The Guidebook to

Membrane Desalination Technology

Reverse Osmosis, Nanofiltration and Hybrid Systems
Process, Design and Applications

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Mark Wilf

With chapters by Leon Awerbuch, Graeme Pearce,
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Introduction

Mark Wilf* *Chapters 1–15*

The reverse osmosis (RO) technology is at present the most versatile desalination method. It is effective in treating water of any salinity, staring with low salinity brackish water up to high salinity seawater. RO units are applied to produce ultrapure water for the semiconductor industry and potable water from high salinity seawater in a cost effective way.

The RO technology started as a scientific experiment at the University of Florida in the 1950s where Reid and Breton (1) were able to demonstrate desalination properties of a cellulose acetate film. At that time reverse osmosis was just one of a number of desalination processes that were evaluated for commercial use (2). After the development of the first asymmetric membrane material from cellulose acetate by Loeb and Sourirajan in the 1950s (3, 8), the subsequent progress included the development of composite membranes with better performing and more robust polyamide based membrane chemistry (4). This was followed by development and optimization of membrane module configurations and the improvement of the process and RO system design. Within the next five decades of continuous development, inventions and improvements, the RO process has been transferred from a scientific curiosity into a self-supporting, rapidly growing industry.

The scientific experiment of the 1950s, which produced a few drops of desalted water per hour, results today (16, 19) in a worldwide network of RO plants of a combined desalting capacity of about 20 million cubic meters per

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day (5.2 billion gallons per day) out of a total of 40 million m^3/d. Reverse osmosis technology is used today in large municipal plants to produce potable water quality from brackish and seawater sources, reclaim municipal effluents and reduce water salinity for industrial applications. At the other end of the application spectrum reverse osmosis membrane elements are used in small under the sink units to produce a few liters per day of drinking water.

A wide variety of membrane material chemistry and membrane module configurations have been developed over the years. However, current commercial membrane modules are almost exclusively made of composite aromatic polyamide membrane materials in a spiral wound configuration. A small fraction of RO elements are still made of cellulose acetate polymers in spiral wound and capillary configuration.

Significant efforts have been made to reduce the investment and operating cost of desalination systems. Better understanding of feed water quality requirements and the introduction of organic scale inhibitors resulted in simplification of the feed water pretreatment process and the increase of feed to permeate conversion ratio.

The development of high flux membrane elements and incorporation of variable speed drivers and power recovery equipment into RO system configuration resulted in significant reductions of specific power consumption. One of the remaining unresolved problems of RO technology is effective control of biofouling in RO systems equipped with polyamide membrane elements. This is especially important in applications involving treatment of feed water from surface or wastewater sources.

The initial cellulose acetate membrane manufactured in the late sixties had a specific permeate flux of about 0.32 $l/m^2/h/bar$ (0.013 gfd/psi) of net driving pressure and salt transport coefficient of 1.5E-05 cm/s. The early RO membranes required a net driving pressure of over 70 bar (1000 psi) in order to produce a permeate flux rate of 25.5 $l/m^2/h$ (15 gfd). The latest generation of polyamide brackish water membranes has specific flux of 5.9 $l/m^2/h/bar$ (0.24 gfd/psi) and salt transport coefficient below 1.0E-05 cm/s. The corresponding net driving pressure required to produce a flux rate of 25.5 (15 gfd) is only 4.3 bar (62 psi) with higher salt rejection (about 35% lower salt passage) than the initial CA membranes. This improvement of specific permeate flux translates into over a twenty fold reduction of the specific power consumption of the RO process pumps. The evolution of membrane performance is summarized in Table 1 and Figs. 1-1 and 1-2.

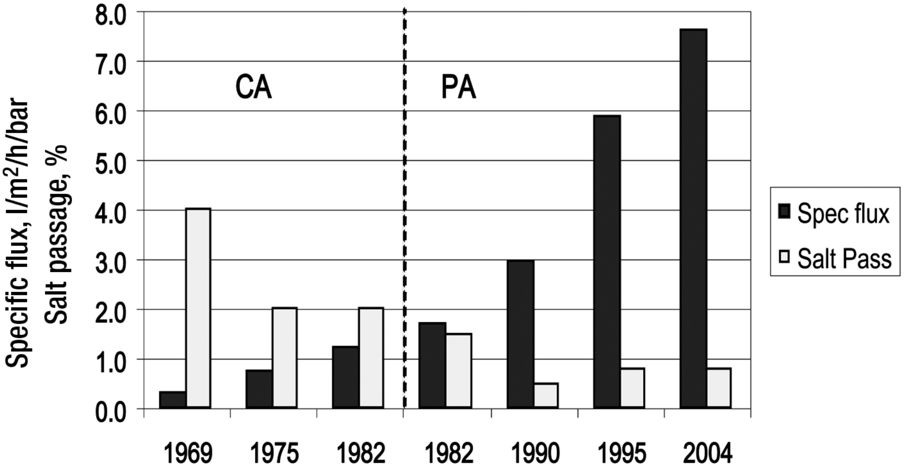


FIG. 1.1 Evolution of performance of brackish RO membranes.

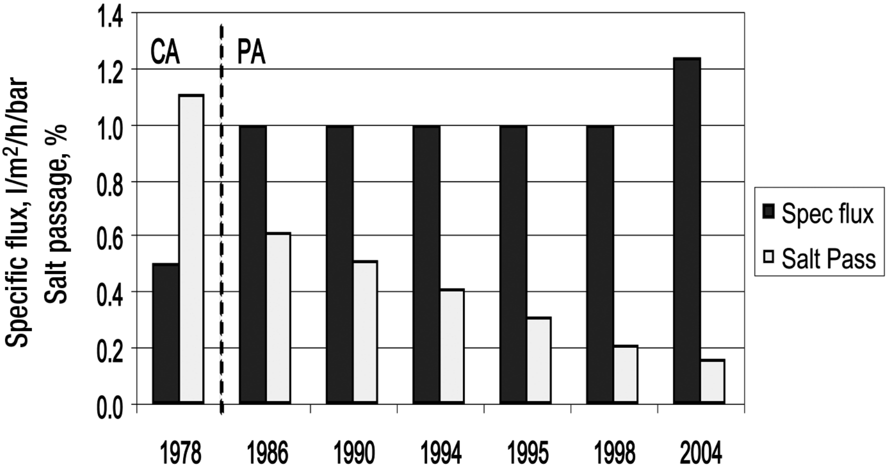


FIG. 1.2 Evolution of performance of seawater RO membranes.

TABLE 1.1
Evolution of performance of flat sheet brackish membranes.

Year	1969	1975	1982	1982	1990	1995	2004
Membrane type	CA asymmetric			PA composite			
Specific flux, l/m^2/h/bar (gfd/psi)	0.32 (0.013)	0.76 (0.031)	1.24 (0.05)	1.73 (0.0.7)	2.96 (0.12)	5.93 (0.24)	8.15 (0.31)
Salt rejection, %	96.0	98.0	98.0	98.5	99.5	99.0	99.2
Salt transport, cm/s	1.5E-5	1.2E-5	1.9E-5	2.0E-5	5.8E-6	6.0E-6	
NDP required for flux 26 l/m^2/h (15 gfd), bar (psi)	79.3 (1150)	33.3 (483)	20.7 (300)	14.8 (214)	8.6 (125)	4.3 (62.5)	3.1 (45.5)

CA asymmetric: Cellulose acetate membrane material, asymmetric structure of membrane layer.
PA composite: Polyamide membrane material, composite structure. Ultra thin membrane barrier cast on a porous support.

A comprehensive review of early development work and theory of the RO process is included in a book edited by Merten (6). An early review of RO applications and RO systems design is included in a publication edited by Buros (5).

The interesting point is that the basic composition of aromatic polyamide membrane has not changed since its invention in the 1980s. The improvements of performance are the result of better morphology of the composite membrane structure. They were achieved mainly through better control of the manufacturing parameters due to improved understanding of the membrane formation process.

2

Introduction to reverse osmosis: basic terms

2.1. Water salinity—concentration units

The concentration of dissolved ions in water is expressed as parts per million (ppm), which is equivalent to milligrams per liter (mg/l) or grams per cubic meter (g/m^3). Sometimes concentration is expressed as a percent of weight of dissolved ions in the total weight of solution. Approximately 10,000 ppm is equal to 1% weight concentration. Concentration can also be expressed as milliequivalents per liter (meqv/l). Miliequivalent concentration is calculated by dividing concentration, expressed as ppm, by equivalent weight, which is ion weight devided by valency. For example calcium (Ca) ion concentration of 1000 ppm is equivalent to meq concentration of 1000/(40/2) = 50 meq/l. Natural water sources contain dissolved ions at various concentrations ranging from a few hundreds ppm for low salinity well or surface water to 35,000–45,000 ppm for a seawater source.

Direct determination of total dissolved solids (TDS) concentration is quite cumbersome. It requires evaporation to dryness of accurately weighted volume of water solution and weight determination of the dry residue. The most common approach to measurement of concentration of total dissolved solids in solution is through analytical determination of concentration of dissolved ions and summation of individual ion concentrations. Approximate determination of the concentration of dissolved ions can be accomplished by measurement of electric conductivity of the water solutions according to the equation:

$$\mathrm{TDS} = K \cdot \mathrm{EC} \qquad (1)$$

where TDS is concentration in parts per million (ppm), K is a conversion factor and EC is electric conductivity in µS/cm.

The composite conductivity of ions in solution is easily measurable. It is customarily expressed as micro Siemens per cm (µS/cm). Electric conductivity of solution is affected by temperature and ion composition. Conductivity varies about 2% per degree C and is customarily reported as a corrected value at 25°C. Different ions have different contributions to conductivity. Therefore conductivity conversion to TDS has to be calibrated with different TDS determinations for each site. The common values of conversion factor K are in the range of 0.55 for an RO permeate up to 0.75 for seawater concentrate (7).

2.2. The osmotic process

Osmosis is a natural process involving fluid flows across a semipermeable membrane barrier. It is selective in the sense that the solvent passes through the membrane at a faster rate than the passage of dissolved solids. The difference of passage rate results in solvent-solids separation. The direction of solvent flow is determined by its chemical potential which is a function of pressure, temperature, and concentration of dissolved solids. Pure water in contact with both sides of an ideal semipermeable membrane at equal pressure and temperature has no net flow across the membrane because the chemical potential is equal on both sides. If a soluble salt is added to water on one side of the membrane, the chemical potential of this salt solution is reduced. Osmotic flow from the pure water side across the membrane to the salt solution side will occur until the equilibrium of chemical potential is restored (Figure 2.1a). Equilibrium occurs when the hydrostatic pressure differential resulting from the volume changes on both sides is equal to the osmotic pressure. This is a solution property independent of the membrane. Application of an external pressure to the salt solution side, which is equal to the osmotic pressure, will also cause equilibrium. Additional pressure will raise the chemical potential of the water in the salt solution and cause a solvent flow to the pure water side, because it now has a lower chemical potential. This phenomenon is called reverse osmosis (Figure 2.1b).

The osmotic pressure, P_{osm}, of a solution can be determined indirectly by measuring the concentration of dissolved salts in solution:

$$P_{\mathrm{osm}} = R(T + 273)\ \Sigma(m_i) \tag{2}$$

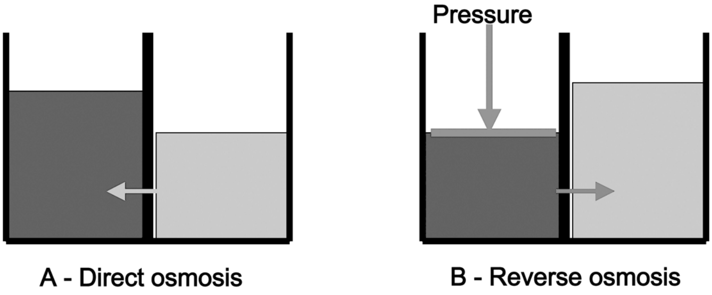


FIG. 2.1 The direct and reverse osmosis process.

where P_{osm} is osmotic pressure (in bar), R is the universal gas constant (0.082 L · atm/mol °K), T is the temperature (in °C), and $\Sigma(m_i)$ is the sum of molar concentration of all constituents in a solution.

An approximation for P_{osm} can be made by assuming that 1000 ppm concentration of TDS equals about 0.77 bar (11 psi) of osmotic pressure. For example, in an RO unit operating at 75% recovery rate, feed salinity is 3,000 ppm TDS and concentrate salinity is about 11,500 ppm TDS. Accordingly, osmotic pressure of the feed is 2.3 bar (33 psi) and the concentrate is 8.7 bar (126 psi). Eq. 2 only holds for dilute salt solutions and temperatures close to 25°C. At significantly different conditions a more rigorous calculation that takes into consideration ion activities rather than concentrations has to be applied.

2.3. Permeate recovery rate (conversion)

Permeate recovery is one of the more important parameters in the design and operation of RO systems. Recovery or conversion rate of feed water to product (permeate) is defined by Eqs. 3 and 4 and illustrated in Fig. 2.2:

$$R_p = 100\% \, (Q_p/Q_f) \tag{3}$$

$$R_p = 100\% \, (Q_p/(Q_p + Q_c)) \tag{4}$$

where R_p is the product recovery rate (in %), Q_p is the product water flow rate, Q_f is the feed water flow rate and Q_c is the concentrate flow rate.

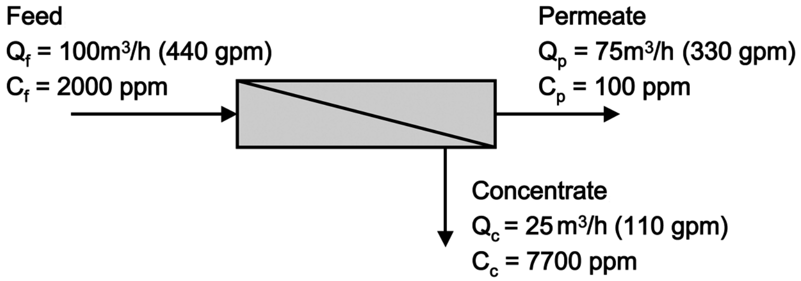


FIG. 2.2 Mass balance in reverse osmosis unit. Recovery rate 75%.

The recovery rate affects salt passage and product flow. As the recovery rate increases, the salt concentration on the feed-brine side of the membrane increases, which causes an increase in salt flow rate across the membrane. Also, a higher salt concentration in the feed-brine solution increases the osmotic pressure, reducing the net driving pressure available and consequently reducing the product water flow rate.

Using Eq. 4 and mass balance a relation can be developed between recovery and concentration of various streams:

$$R_p = Q_p/(Q_p + Q_c)$$

$$Q_f C_f = Q_p C_p + Q_c C_c$$

$$Q_f C_f = Q_p C_p + ((Q_f - Q_p) C_c)$$

$$Q_f (C_f - C_c) = Q_p (C_p - C_c)$$

$$R_p = Q_p/Q_f = (C_c - C_f)/(C_c - C_p) \qquad (5)$$

where C_f is feed concentration, C_p is permeate concentration and C_c is concentrate concentration.

The above relations (Eq. 5) can be applied to determine recovery rate from concentration values of ions in the feed, permeate and concentrate stream. Usually these calculations are based on concentrations of chloride or calcium ions which can be determined easily and with a high degree of accuracy.

Example 1

Designed recovery rate 75%
Concentrations, ppm of Cl^-
Feed = 1000

Concentrate = 3800
Permeate = 200
Actual recovery rate, $R_p = (3800 - 1000)/(3800 - 200) = 0.78$ (78%)

In a multistage RO system the product recovery rate is defined for each stage and for a combined system (Fig. 2.3).

Example 2

Designed recovery rate 0.75 (75%)
Feed concentration, ppm $Cl^- = 1000$
Permeate concentration, ppm $Cl^- = 200$
Calculation of concentrate concentration: $C_c = (C_f - R_p C_p)/(1 - R_p)$
$C_c = (1000 - 0.75 \times 200)/(1 - 0.75) = 3400$

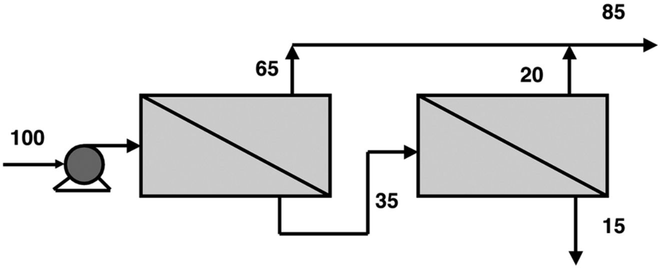


FIG. 2.3 Recovery rate in a two stage RO unit.
Stage 1 $R_{p1} = 65\%$
Stage 2 $R_{p2} = 57\%$ (20%)
Total $R_t = R_{p1} + (1 - R_{p1}/100) \cdot R_{p2}$
$R_t = 65\% + (1 - (65/100)) \cdot 57\% = 85\%$

2.4. Average feed salinity

Average feed salinity (AFS) is a representative value of feed concentration used for calculation of performance of membrane element or RO system. AFS accounts for the phenomenon of salinity increase in the RO system from the salinity of feed water at the entrance to the RO device to the final salinity of concentrate leaving the system. The AFS is calculated as an arithmetic (Eq. 6) or logarithmic mean (Eq. 7). AFS salinity can be expressed as a function of recovery rate, assuming at first approximation that ions are totally rejected by RO membranes:

$$\text{AFS} = 0.5\, C_f\,(1 + 1/(1 - R)) \tag{6}$$

$$\text{AFS} = C_f\, \text{Ln}\,(1/(1 - R))/R \tag{7}$$

where R = recovery rate expressed as a decimal fraction.

The arithmetic mean is usually applied for calculations in cases of low recovery (single element calculations). In case of a high recovery rate a logarithmic mean is used for performance calculations. The values of concentration factors for the arithmetic average, logarithmic average and for concentrate are shown in Fig. 2.4.

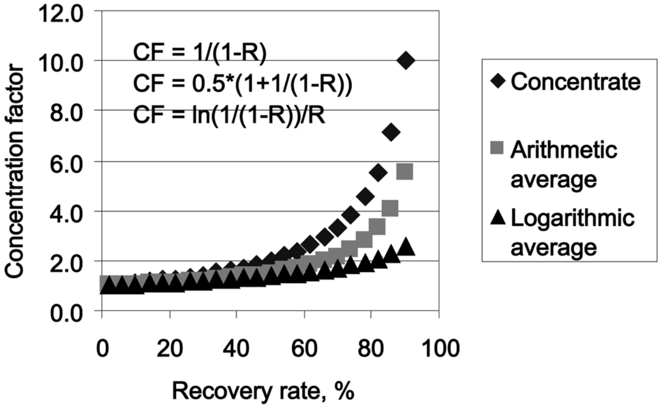


FIG. 2.4 Concentration increase vs. recovery rate.

2.5. Net driving pressure

The net driving pressure (NDP) is the driving force of the water transport through the semipermeable membrane. The value of NDP decreases along the RO unit. Therefore, for the purpose of membrane performance calculations it is defined as an average NDP. The NDP is defined as the fraction of the applied pressure in excess of average osmotic pressure of the feed and any pressure losses in the system according to Eq. 8.

$$\text{NDP} = P_f - P_{os} - P_p - 0.5P_d\,(+\, P_{osp}) \tag{8}$$

where P_f = feed pressure
P_{os} = average feed osmotic pressure
P_p = permeate pressure
P_d = pressure drop across RO elements
P_{osp} = osmotic pressure of permeate.

In regular RO applications osmotic pressure of permeate is negligible. However, in NF applications, where salt rejection is relatively low, permeate salinity is significant compared to the feed concentration. Therefore, osmotic pressure of permeate has to be considered in calculation of NDP in NF systems.

Example 3

Brackish RO system operates at recovery rate of 85%
Feed salinity: 2000 ppm TDS
Permeate salinity = 25 ppm TDS
Concentrate salinity = (2000 – 0.85 × 25)/(1 – 0.85) = 13191 ppm
Feed pressure: 13 bar (188.5 psi)
Pressure drop across the system: 3.0 bar (45 psi)
Permeate pressure: 1.0 bar (14.5 psi)
Average feed salinity = 0.5 (2000 + 13191) = 7596 ppm
Average feed osmotic pressure = 0.77 (7596/1000) = 5.9 bar (85.5 psi)
Permeate osmotic pressure = 0.77 (25/1000) = 0.02 bar (0.3 psi)
Average net driving pressure: NDP = 13 – 5.9 – 1.0 – 0.5 × 3.0 = 4.6 bar (67 psi)

Example 4

Nanofiltration RO system operates at recovery rate of 85%
Feed salinity: 1000 ppm TDS
Permeate salinity = 350 ppm TDS
Concentrate salinity = (1000 – 0.85 × 350)/(1 – 0.85) = 4683 ppm
Feed pressure: 7 bar (101 psi)
Pressure drop across the system: 3.0 bar (45 psi)
Permeate pressure: 1.0 bar (14.5 psi)
Average feed salinity = 0.5 (1000 + 4683) = 2842 ppm
Average feed osmotic pressure = 0.77(2842/1000) = 2.2 bar (32 psi)
Permeate osmotic pressure = 0.77 (350/1000) = 0.3 bar (4 psi)
Average net driving pressure: NDP = 7 – 2.2 – 1.0 – 0.5 × 3.0 + 0.3 = 2.6 bar (38 psi)

2.6. Water reversed salt separation by reverse osmosis

The mechanism of water and salt separation by reverse osmosis is not fully understood. Current scientific thinking suggests two water transport models: porosity and diffusion. That is, transport of water through the membrane may be through physical pores present in the membrane (porosity), or by diffusion from one bonding site to another within the membrane. The theory suggests that the chemical nature of the membrane is such that it will absorb and pass water preferentially to dissolved salts at the solid/liquid interface. This may occur by weak chemical bonding of the water to the membrane surface or by dissolution of the water within the membrane structure. Either way, during this process a salt concentration gradient is formed across the solid/liquid interface (Fig. 2.7).

The chemical and physical nature of the membrane determines its ability to allow for preferential transport of solvent (water) over solute (salt ions). It is also known that part of dissolved species rejection is a result of size discrimination,, i.e., larger molecules are better rejected by the RO membranes than the small ones. Another part of the rejection process is a result of the repulsion of dissolved ions due to presence of charges (usually negative) on the membrane surface. The membrane repels ions having the same charges as those present on the membrane surface. Usually ions of multiple negative valency (high ion charges) are better rejected than single valency ions. For example, rejection of sulfate ions is higher than rejection of chloride ions. Due to the condition of maintaining electro-neutrality of solutions on both sides of the membrane, repulsion of one type of ion hinders passage of the co-ion and increases overall rejection. Conversely, presence of ions with high passage through the membrane will increase passage of the corresponding co-ion. The extent of this charge-dependant rejection will vary with the composition of solution treated (see also chapter 10. Feed water composition).

2.7. Water transport

The rate of water passage through a semipermeable membrane is defined in Eq. 9:

$$Q_w = (\Delta P - \Delta P_{osm})\, K_w\, (S/d) \tag{9}$$

where Q_w is the rate of water flow through the membrane, ΔP is the hydraulic pressure differential across the membrane, ΔP_{osm} is the osmotic pressure differential across the membrane, K_w is the membrane permeability coefficient for water, S is the membrane area, and d is the membrane thickness.

This Eq. can be simplified to:

$$Q_w = A \cdot S \cdot NDP \tag{10}$$

where *A* is water transport coefficient and represents a unique constant for each membrane material type, and *NDP* is the net driving pressure or net driving force for the mass transfer of water across the membrane.

A units are: $g/cm^2/s$ ($gal/ft^2/d/psi$)

2.8. Salt transport

The rate of salt flow through the membrane is defined by Eq. 11:

$$Q_s = \Delta C \cdot K_s\ (S/d) \qquad (11)$$

where Q_s is the flow rate of salt through the membrane, K_s is the membrane permeability coefficient for salt, ΔC is the salt concentration differential across the membrane, *S* is the membrane area, and *d* is the membrane thickness.

K_s units are: cm/s (ft/s)

Eq. 11 can be simplified to:

$$Q_s = B \cdot S\ (\Delta C) \qquad (12)$$

Where *B* is the salt transport coefficient and represents a unique constant for each membrane type, and ΔC is the concentration gradient which is the driving force for the transfer of dissolved ions through the membrane.

Eqs. 10 and 12 show that for a given membrane:

a) Rate of water flow through a membrane is proportional to the *NDP* differential across the membrane.
b) Rate of salt flow is proportional to the concentration differential across the membrane and is independent of applied pressure.

Salinity of the permeate, C_p, depends on the relative rates of water and salt transport through the reverse osmosis membrane:

$$C_p = Q_s/Q_w \qquad (13)$$

The fact that water and salt have different mass transfer rates through a given membrane creates the phenomena of water–salt separation and salt rejection. No membrane is ideal in the sense that it absolutely rejects salts; rather the different transport rates of water and dissolved ions create an apparent rejection. Eqs. 9–12 explain important design considerations in RO systems. For example, an increase in operating pressure will increase water flow without significantly affecting salt flow, thus resulting in lower permeate salinity. On the other hand a higher recovery rate will increase the concentration gradient and result in higher permeate salinity.

2.9. Salt passage and salt rejection

Salt passage is defined as the ratio of concentration of salt on the permeate side of the membrane relative to the average feed concentration. Mathematically, it is expressed in Eq. 14:

$$SP = 100\% \, (C_p/C_{fm}) \tag{14}$$

where SP is the salt passage (in%), C_p is the salt concentration in the permeate, and C_{fm} is the mean salt concentration in the feed stream.

Applying the fundamental equations of water flow and salt flow illustrates some of the basic principles of RO membranes. For example, apparent salt passage is an inverse function of pressure; that is, the salt passage increases as applied pressure decreases. This is because with reduced feed pressure permeate flow rate decreases and hence dilution of salt, on the permeate side of the membrane, decreases as well (the salt flows through the membrane at a constant rate as the rate of flow is independent of pressure).

Salt rejection is the opposite of salt passage, and is defined by Eq. 15:

$$SR = 100\% - SP \tag{15}$$

where SR is the salt rejection (expressed as a percent), and SP is the salt passage as defined in Eq. 14. Salt rejection is an important performance parameter of RO membranes, determining suitability of given membranes for various applications. The above relations for water and salt transport imply constant values of transport rates. However, salt and water transport rates are strongly affected by temperature, changing at similar rates with temperature fluctuations.

Example 5

RO element is tested at 15 recovery rate, $R_p = 0.15$

Feed salinity: 1500 ppm NaCl

Permeate salinity: 4.5 ppm NaCl

Average feed salinity = $0.5\ C_f\ (1 + 1/(1 - R_p)) = 0.5 \times 1500 \times (1 + 1/(1 - 0.15)) = 1632$ ppm

Salt passage = 100% (4.5/1632) = 0.28%
Salt rejection = 100% – 0.28% = 99.72%

2.10. Temperature effect on transport rate

Feed water temperature effect rate of diffusive flow through the membrane. For RO calculations the following equation is being used to calculate temperature correction factor (TCF), applied for calculation of water permeability:

$$\mathrm{TCF} = 1/\exp(C\,(1/(273 + t) - 1/298)) \tag{16}$$

where t is temperature °C, C is constant, characteristic of membrane barrier material. For polyamide membranes C values of 2500–3000 are being used.

It is customary for RO applications to use a temperature of 25°C (77°F) as the reference temperature, for which TCF = 1.0. The water and salt transport increases about 3% per °C. Fig. 2.5 shows a value of TCF in the temperature range of 5–50°C. By comparison, relative values of water viscosity are also included. There is striking similarity of both curves, suggesting that changes of water permeability with temperature are a result of viscosity changes.

The results in Fig. 2.5 suggest that due to increased permeability with temperature increase the operating feed pressure should be lower at higher temperature. This is indeed the situation in case of processing of low salinity feed (brackish applications). This is also the case for RO seawater applications in the low range of feed water temperatures. However, at feed water temperatures

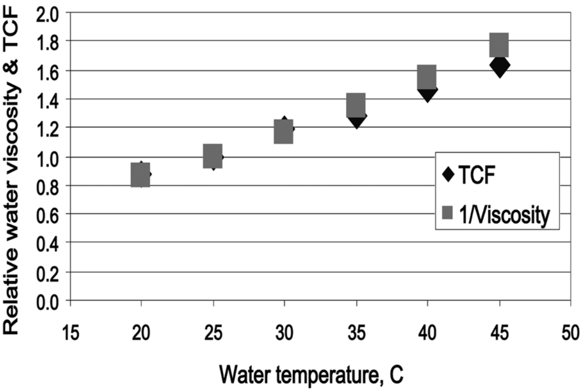


FIG. 2.5 Temperature effect on water permeability and viscosity.

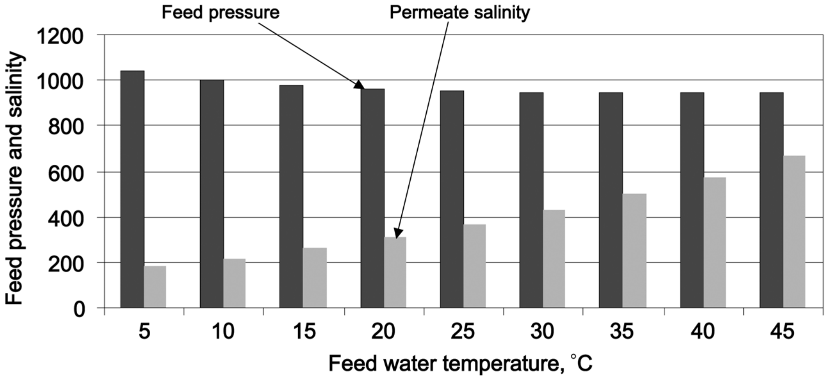


FIG. 2.6 Temperature effect on feed pressure and permeate salinity in seawater RO unit.

above 30°C the subsequent decrease of required feed pressure levels off. The effect of increased permeability is reduced by increased osmotic pressure of the seawater feed. In addition, increase of salt passage and potential need for partial second pass processing, may actually result in higher overall power consumption at the high end of feed water temperature (Fig. 2.6).

Example 6

RO system operates in temperature range of 15–28°C.

At feed water temperature of 28°C the NDP_1 required for the design capacity is 7.5 bar (108.9 psi). What NDP will be required at 15°C to maintain the design permeate capacity?

Value of constant C (Eq. 16) for the membrane used is 2700.

For 28°C the $TCF_1 = 1/(\exp(2700(1/(273 + 28) - 1/298) = 0.914$
For 15°C the $TCF_2 = 1/(\exp(2700(1/(273 + 15) - 1/298) = 1.370$
$NDP_2 = NDP_1\ TFC_2/TFC_1 = 7.5 \times 1.370/0.914$
$NDP_2 = 11.2$ bar (162.4 psi)

2.11. Average permeate flux (APF)

Average permeate flux is another important design parameter of the RO process. APF is combined permeate flow divided by the total membrane area installed in the RO unit. Units: $l/m^2/h$ or gfd ($gal/ft^2/d$)

$$APF = Qp/(EN \times MA) \tag{17}$$

where Q_p = permeate flow rate
EN = number of elements in the system
MA = membrane area per element

Example 7

RO system produces 400 m^3/d (105,700 gal/d). Membrane array consists of 3 pressure vessels, each housing 6 membrane elements. Each element has 37 m^2 of membrane area (400 ft^2). APF is calculated as follows:

$APF = 400{,}000\ l/d/(3 \times 6 \times 37m^2 \times 24) = 25.0\ l/m^2/h$

$APF = 105{,}700\ gal/d\ /(3 \times 6 \times 400\ ft^2) = 14.6\ gfd$

Conversely, design APF is used to determine the required number of membrane elements in the RO system for required permeate capacity.

Example 8

RO system will produce 400 m^3/h (1,760 gal/min). The design flux rate is 25 $l/m^2/h$ (14.7 gfd). Membrane element type selected for this system is ESPA2 with 37 m^2 of membrane area (400 ft^2) per element.

Number of membrane elements required (NEPV):

$NEPV = 400{,}000\ l/h/(25\ l/m^2/h \times 37\ m^2) = 432$ elements

$NEPV = 1{,}760 gal/min \times 1440\ min/d\ /(14.7\ gal/ft^2/d \times 400\ ft^2) = 431$ elements. The number of elements required will be rounded up according to the number of elements per vessel. Assuming seven elements per vessel:

431/7 = 61.6 pressure vessels

Number of elements required: $62 \times 7 = 434$ elements

2.12. Specific permeability of a membrane

Specific permeability, or specific flux (SF), characterizes the membrane material in terms of water flux rate driven by the gradient of applied net driving pressure.

$$SF = APF/NDP \tag{18}$$

Specific permeability depends on the resistance of the membrane to water flow. This resistance is a composite of the flow resistance of the membrane barrier, support layers and any foulant layer on the membrane surface. It is usually calculated for the feed water temperature of 25°C. Specific flux units are: $l/m^2/h$-bar (gfd/psi)

Example 9

RO membrane element is tested at the following test conditions:
Feed salinity = 1500 ppm NaCl
Recovery rate = 15% (0.15)
Feed pressure = 10.3 bar (150 psi)
Pressure drop = 0.2 bar (2.9 psi)
Permeate pressure: 0.1 bar (1.5 psi)
Permeate flow: 41.6 m^3/d (11,000 gpd)
Membrane area: 39.5 m^2 (430 ft^2)
Average permeate flux = $41.6 \times 1000/(24 \times 39.5) = 43.9$ $l/m^2/h$ (25.8 gfd)
Average feed salinity = $1500 \times 0.5 \times (1 + 1/(1 - 0.15)) = 1632$ ppm NaCl
Average osmotic pressure = $1632/1000 \times 0.77 = 1.3$ bar (19 psi)
Net driving pressure = $10.3 - 1.3 - 0.1 - 0.5 \times 0.2 = 8.8$ bar (128.0 psi)
Specific flux, SF = $43.9/8.8 = 4.99$ $l/m^2/h/bar$ (0.20 gfd/psi)

2.13. Concentration polarization

As water flows through the membrane and salts are rejected by the membrane, a boundary layer is formed near the membrane surface. In this layer the salt concentration exceeds the salt concentration in the bulk solution. This in-

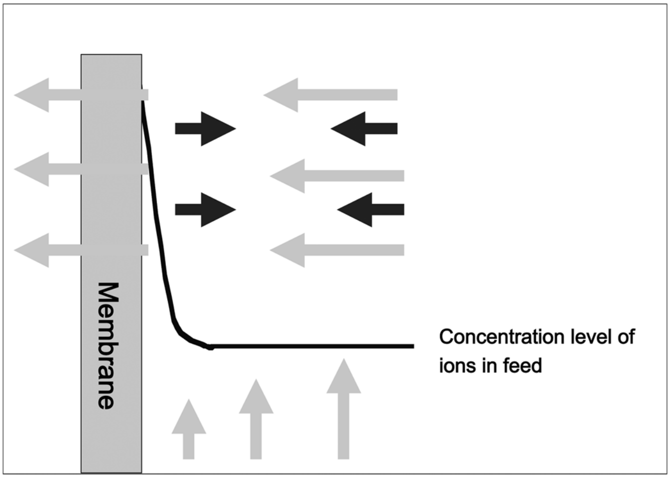


FIG. 2.7 Concentration polarization phenomena.

crease of salt concentration at the membrane surface is called concentration polarization. As shown on Fig. 2.7, during the RO process there is a convective flow of water and ions toward the membrane surface. Ions rejected by the membrane, diffuse back to the bulk due to the concentration gradient. The observed effect of concentration polarization is reduction of actual product water flow rate and salt rejection versus theoretical estimates.

Concentration polarization has the following effect on the RO process:

1. Increased osmotic pressure at the membrane surface than in the bulk feed solution, ΔP_{osm}, and reduced net driving pressure differential across the membrane ($\Delta P - \Delta P_{osm}$).
2. Reduced water flow across the membrane (Q_w).
3. Increased salt flow across the membrane (Q_s).
4. Increased probability of exceeding solubility of sparingly soluble salts at the membrane surface, and the distinct possibility of precipitation causing membrane scaling.

The Concentration Polarization Factor (CPF) is defined as a ratio of salt concentration at the membrane surface (C_s) to bulk concentration (C_b):

$$\mathrm{CPF} = C_s / C_b \tag{19}$$

An increase of permeate flux will increase the delivery rate of ions to the membrane surface and increase C_s. An increase of feed flow, parallel to the membrane surface, increases turbulence and reduces the thickness of the high concentration layer near the membrane surface. Therefore, the CPF will increase with increase of permeate flow (Q_p) due to increase of permeate flux, and will decrease with increase of the average feed flow (Q_{favg}).

$$\mathrm{CPF} = K_p \exp(Q_p / Q_{favg}) \tag{20}$$

Where K_p is a constant depending on membrane element geometry, namely configuration and dimensions of feed channels and of feed spacer.

Using the arithmetic average of feed and concentrate flow as an average feed flow, the CPF can be expressed as a function of the permeate recovery rate of a membrane element (R_i).

$$\mathrm{CPF} = K_p \exp(2R_i / (2 - R_i)) \tag{21}$$

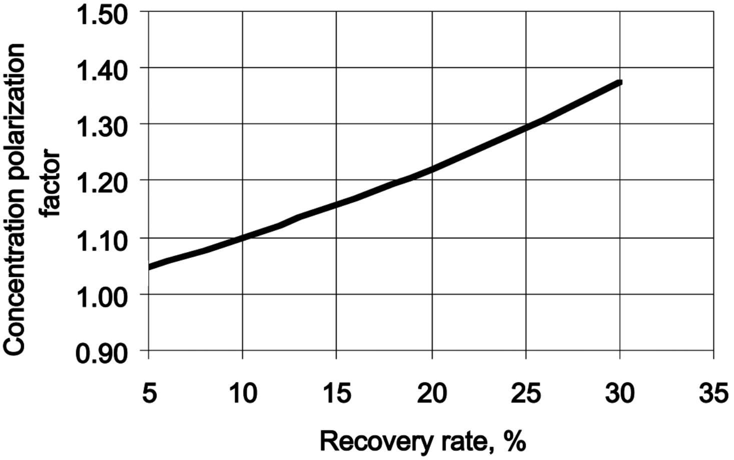


FIG. 2.8 Relative concentration polarization factor vs. membrane element recovery rate.

The value of the Concentration Polarization Factor of 1.2, which is the recommended limit by some membrane manufacturers, corresponds to 18% permeate recovery for a 1 m (40″) long membrane element, as shown in Fig. 2.8. The value of CPF is applied in calculations of RO elements performance to express excess concentration adjacent to the membrane surface.

3

Commercial RO/NF membrane technology

The semipermeable membrane for nanofiltration and reverse osmosis applications consists of a film of polymeric material composed of a skin layer several thousands angstroms thick and spongy supporting layer approximately 0.25–0.50 mm (0.001–0.002″) thick cast on a fabric support. The overall thickness of membrane is 0.15–0.20 mm (0.06–0.08″). The schematic configuration of membrane layers is shown in Fig. 3.1. Scanning electron microscopy (SEM) picture of cross section of cellulose acetate membrane, at 500 times magnification is shown in Fig. 3.2. Corresponding SEM picture of cross section of polyamide membrane is shown in Fig. 3.3.

The commercial grade RO membrane must have high water permeability and a high degree of semipermeability; that is, the rate of water transport must be much higher than the rate of transport of dissolved ions. The membrane material must be stable over a wide range of pH and temperature, and have good mechanical integrity. The stability of membrane performance: permeability and salt rejection, over a period of time at field conditions defines the commercially

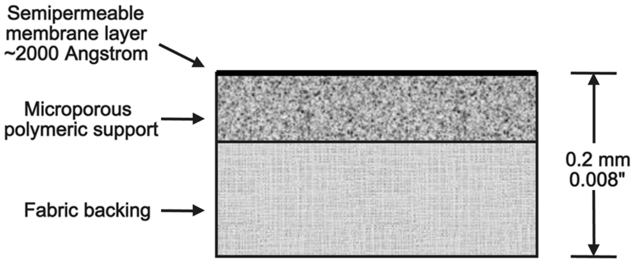


FIG. 3.1 Cross section configuration of flat sheet RO-NF membrane.

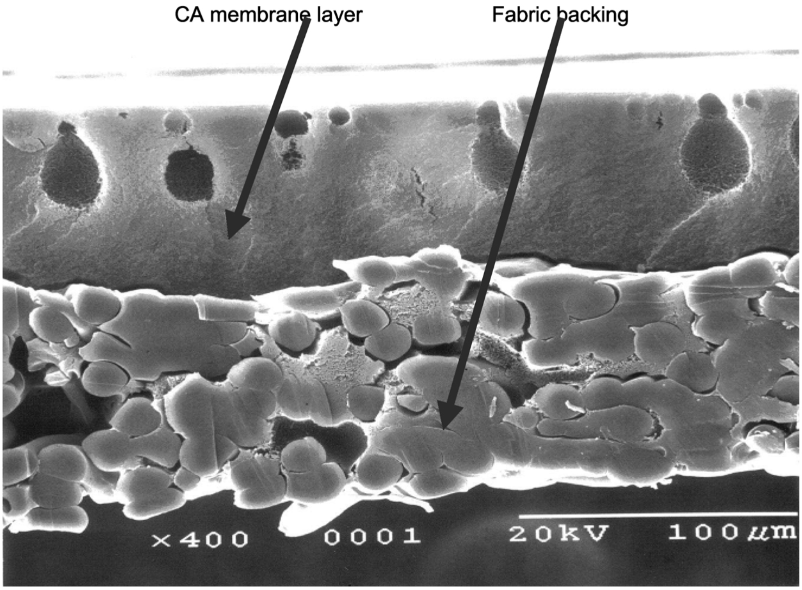


FIG. 3.2 SEM picture of cross section of cellulose acetate membrane.

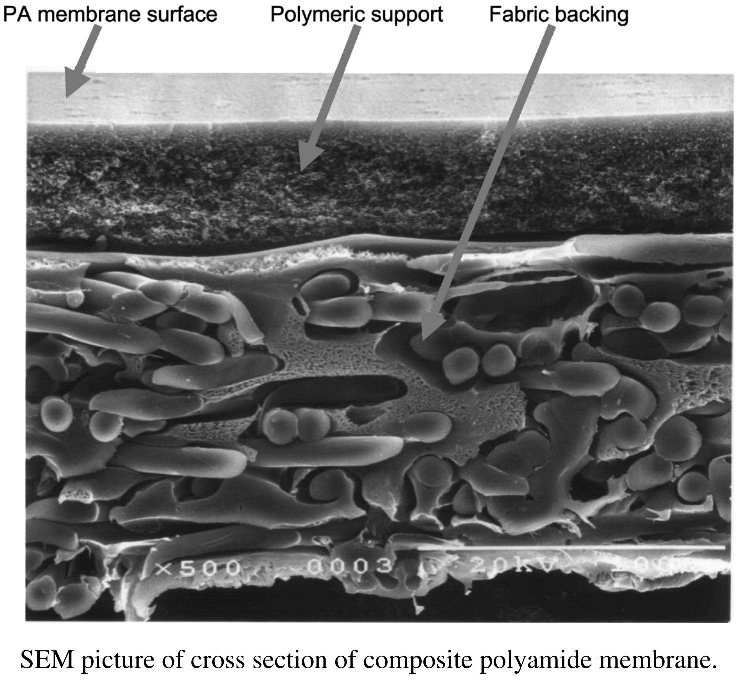


FIG. 3.3 SEM picture of cross section of composite polyamide membrane.

useful membrane life, which is, for the current commercial membranes, in the range of 5 to 10 years.

There are two major groups of polymeric materials, which are used to produce commercial reverse osmosis membranes: cellulose acetate (CA) and polyamide (PA). Polymer structure (Fig. 3.4), chemical tolerance, membrane manufacturing, operating conditions, and performance differ significantly for each group of polymeric material.

3.1. Cellulose acetate membranes

The original cellulose acetate membrane, developed in the late 1950s by Loeb and Sourirajan, was made from cellulose diacetate polymer (3). Current CA membrane is usually made from a blend of cellulose diacetate and triacetate. The membrane is formed by casting a thin film acetone-based solution of cellulose acetate polymer with swelling additives from a trough onto a nonwoven polyester fabric (Figs. 3.5 and 3.6).

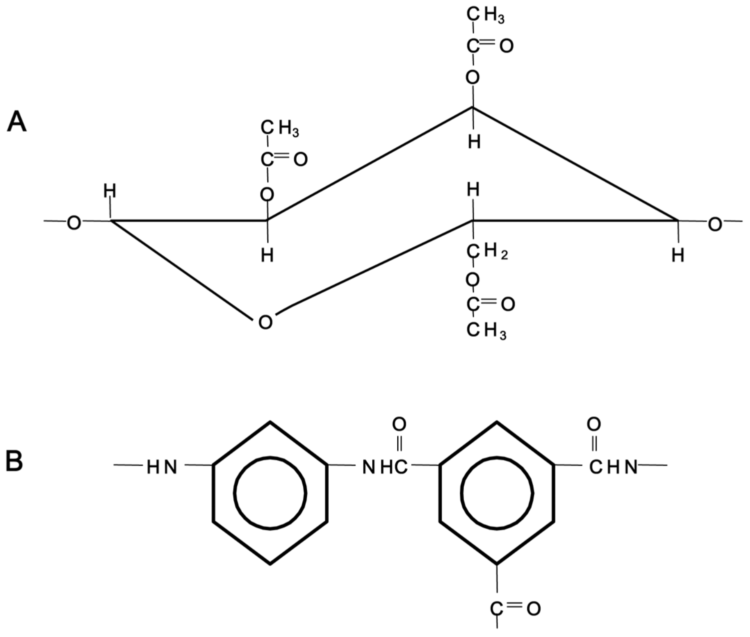


FIG. 3.4 Chemical structure of cellulose triacetate (A) and polyamide (B) membrane material.

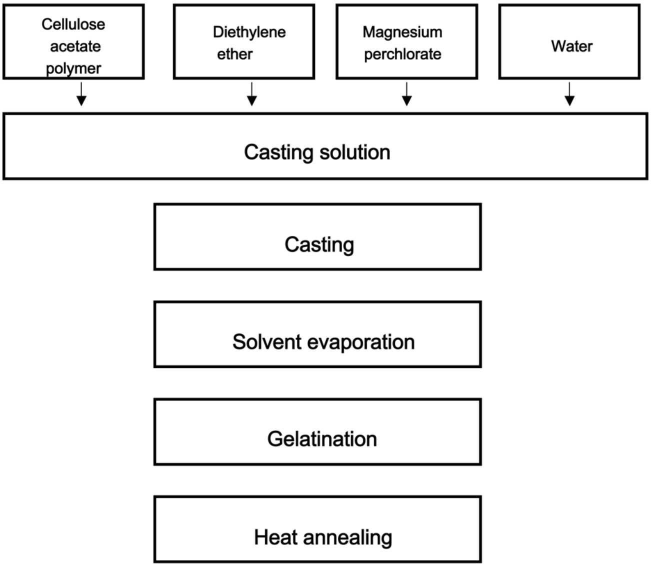


FIG. 3.5 Manufacturing sequence of cellulose acetate membrane.

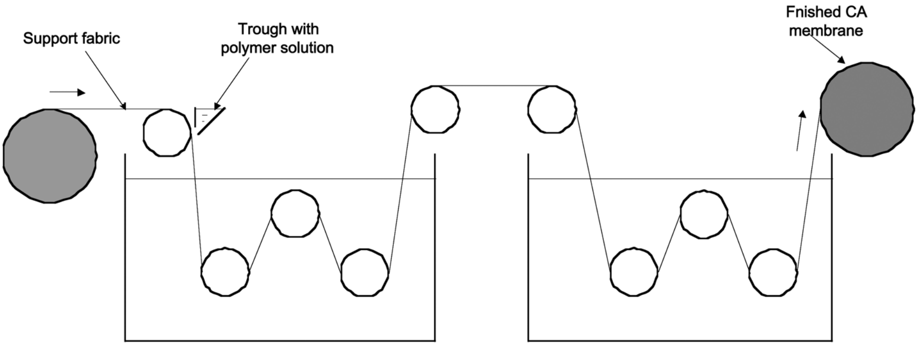


FIG. 3.6 Manufacturing process of cellulose acetate membrane.

Two additional steps, a cold bath followed by high temperature annealing, complete the casting process. During casting, the solvent is partially removed by evaporation. After the casting step, the membrane is immersed into a cold water bath which removes the remaining acetone and other leacheable compounds. Following the cold bath step, the membrane is annealed in a hot water bath at a temperature of 60–90°C. The annealing step improves the semipermeability of the cellulose acetate membrane resulting in a decrease of water transport and a significant decrease of salt passage. After processing, the cellulose membrane has an asymmetric structure with a dense surface layer of about 1000–2000 angstrom (0.1–0.2 micron) which is responsible for the salt rejection property. The rest of the membrane film is spongy and porous and has high water permeability. Salt rejection and water flux of a cellulose acetate membrane can be controlled by variations in temperature and duration of the annealing step. Description of manufacturing process of cellulose acetate membranes and its properties can be found in number of publications (6, 8, 9, 10).

Cellulose acetate membrane polymer hydrolyzes rapidly at extreme pH. Therefore, the operating feed water pH range for CA membrane is 6–8. Accordingly, cellulose acetate membrane elements can be only cleaned in the narrow range of pH close to neutral (pH: 6–8). However, CA membrane polymer has sufficient tolerance to free chlorine that enables operation with chlorinated feed water and on line disinfection to control bacterial growth. For this reason cellulose acetate is still membrane of choice for applications where frequent disinfection of RO system with free chlorine is practiced, such as pharmaceutical industries and some food applications. Also, one of membrane manufacturers presently produces capillary RO membranes for seawater desalting using cellulose acetate polymer. Except for the previously mentioned applications, rest of the desalination market is dominated by the composite polyamide membranes in spiral wound configuration.

3.2. Composite polyamide membranes

The manufacturing process of composite polyamide membranes consists of two distinct steps (Fig. 3.7). First, a polysulfone support layer is cast onto a non-woven polyester fabric. The process of application of polysulfone and formation of ultrafiltration membrane is very similar to the process of manufacturing of the cellulose acetate membrane (Fig. 3.6). The polysulfone polymer solution is applied from a trough onto a moving polyester backing fabric. After polysulfone application and formation of UF membrane layer the fabric travels through water bath to remove solvent and is collected on a drum.

The polysulfone layer is very porous and is not semipermeable;, i.e., does not have the ability to separate water from dissolved ions solution. However, it has high water permeability In the next process step, the drum with polysulfone membrane is moved to the second machine where interfacial polarization takes place (Fig. 3.8). There, a semipermeable membrane skin is formed on the polysulfone

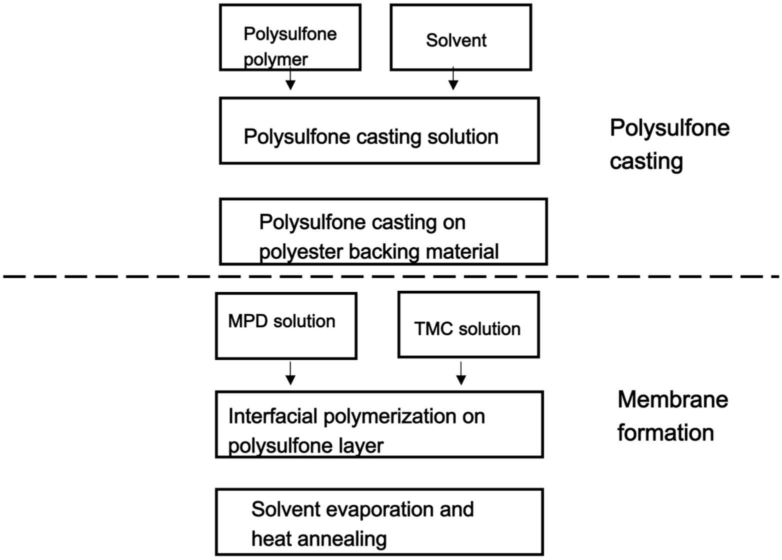


FIG. 3.7 Manufacturing sequence of composite polyamide membrane.

substrate by interfacial polymerization of two monomers, one: metaphenylenediamine (MPD) containing amine groups and the other: trimesoyl chloride (TMC) provides carboxylic acid chloride functional groups. The polymerization reaction is very rapid and takes place on the surface of the polysulfone support forming a barrier, 1000–2000 angstrom thick. This barrier is responsible for the semipermeable property: passage of water and rejection of dissolved species. Following polymerization zone, membrane web enters a rinse bath. The membrane is rinsed to remove excess reagents and passed trough the oven to dry (4).

This manufacturing procedure enables independent optimization of the distinct properties of the membrane support and salt rejecting skin. The resulting composite membrane is characterized by higher specific water flux and lower salt passage than cellulose acetate membranes.

Polyamide composite membranes are stable over a wider pH range than the cellulose acetate membranes. However, polyamide membranes are susceptible to oxidative degradation by free chlorine, while cellulose acetate membranes can tolerate limited levels of exposure to free chlorine. Compared to a polyamide membrane, the surface of cellulose acetate membrane is smooth and

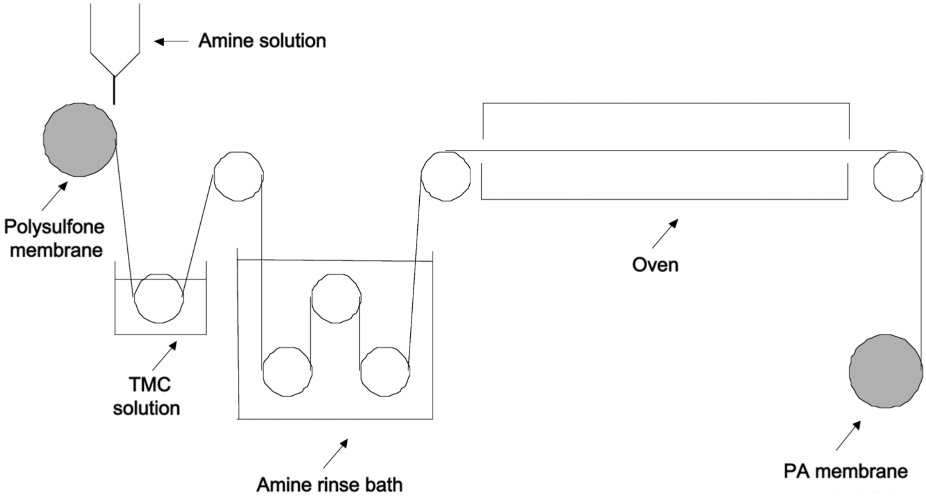


FIG. 3.8 Manufacturing process of composite polyamide membrane barrier on a polysulfone support.

has little surface charge (Figs. 3.9 and 3.10). Because of the neutral surface and tolerance to free chlorine, cellulose acetate membranes will usually have a more stable performance than polyamide membranes in applications where the feed water has a high fouling potential due to presence of dissolved organics such as with municipal effluent and surface water supplies.

The early composite membranes made of aliphatic polymers (9) were very sensitive to presence of oxidants and suffered from inadequate stability of performance in field conditions. The later generation of composite membranes, based on aromatic polyamide invented by Cadotte (4, 12) have some tolerance to free chlorine, good stability in wide range of feed pH (2–10) and shows excellent long term performance stability with majority of feed water types. This type of membrane material is used today almost exclusively to manufacture commercial RO membrane elements.

The variety of types of membranes made of composite aromatic polyamide includes seawater, brackish and nanofiltration membrane elements. Composite membranes are used in all areas of applications: seawater and brackish water desalting, potable water softening, wastewater reclamation, food processing and other industrial applications.

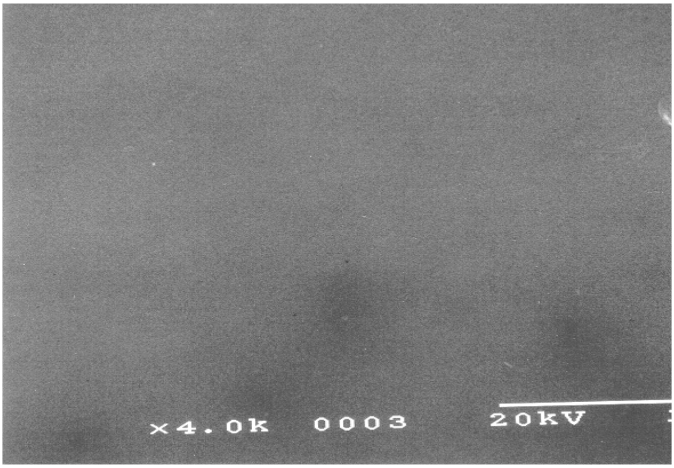


FIG. 3.9 SEM picture of surface of cellulose acetate membrane.

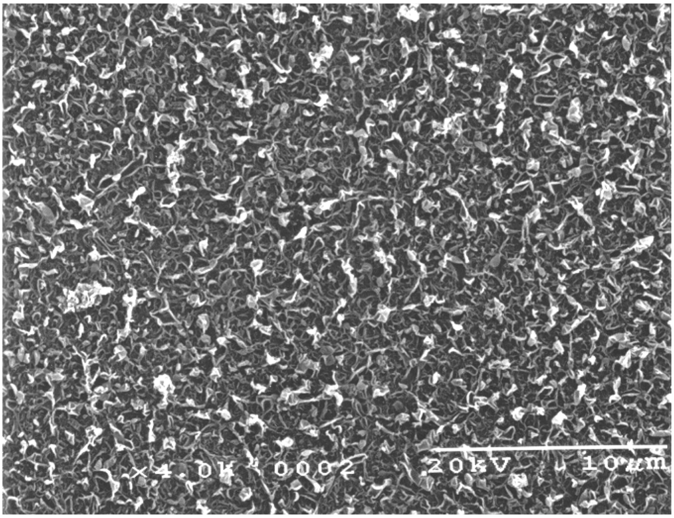


FIG. 3.10 SEM picture of surface of composite polyamide membrane.

4

Membrane module configurations

The reverse osmosis technology started with tubular and plate and frame configurations. Due to low packing density, these initial module configurations were gradually phased out of potable applications and at present are being very infrequently used in conventional reverse osmosis applications. However, new configurations of plate and frame modules are still being used in food processing applications and for treatment of waste streams including land fill leaches. In the past, the two major membrane module configurations used for reverse osmosis applications were hollow fiber and spiral wound. At present, the majority of RO membrane manufacturers offer elements in spiral wound configuration only.

4.1. Plate and frame configuration

The plate and frame configuration has been introduced at the early stages of development of reverse osmosis technology (15) and later on almost abandoned in favor of higher packing density spiral wound and hollow fiber configurations. Today the plate and frame modules are still used in applications where spiral wound and hollow fiber modules can not provide sufficient reliability or performance. These include treatment of streams with high concentration of suspended solids. One of such applications is reduction of volume of land fill leachate (63). In modern plate and frame configuration, the flow regime provides turbulent flow and short feed flow path. Therefore, the tendency for membrane scaling or fouling is significantly reduced. Due to the high cost of membrane modules, the plate and frame configuration is not used in commercial potable applications.

4.2. Hollow fine fiber membrane elements

The concept of hollow fine fiber (HFF) configuration module has been introduced by Mahon (13) in the early sixties. The HFF configuration utilizes semipermeable membrane in the form of hollow fibers which have been extruded from cellulosic or non-cellulosic materials (14). The fiber is asymmetric in structure and is as fine as a human hair, about 40–80 µ (0.0016–0.0030 in.) I.D. and 85–150 µ (0.0033–0.060 in.) O.D. (Fig. 4.1). Millions of these fibers are formed into a bundle and folded in half to a length of approximately 120 cm (4 ft).

A perforated plastic tube, serving as a feed water distributor is inserted in the center and extends the full length of the bundle. The bundle is wrapped and both ends are epoxy sealed to form a sheet-like permeate tube end and a terminal end which prevents the feed stream from bypassing to the brine outlet. The hollow fiber membrane bundle, 10–20 cm (4–8 in.) in diameter, is contained in a cylindrical housing or shell approximately 137 cm (54 in.) long and 15–30 cm (6–12 in.) in diameter. The assembly is called a permeator (Fig. 4.2). The pressurized feed water enters the permeator feed end through the center distributor tube, passes through the tube wall, and flows radially around the fiber bundle toward the outer permeator pressure shell. Water permeates through the outside wall of the fibers into the hollow core or fiber bore, through the bore to the tube

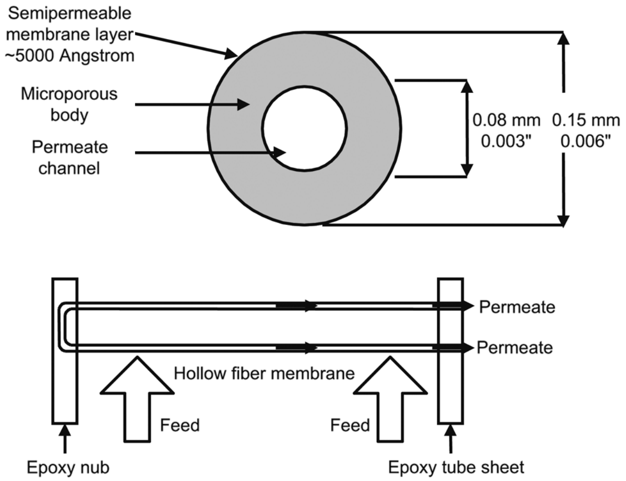


FIG. 4.1 Configuration of the hollow fiber membrane and module.

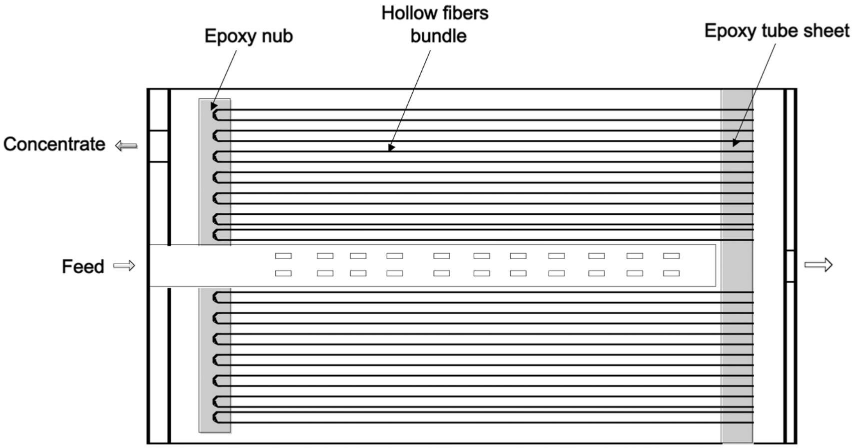


FIG. 4.2 Configuration of the hollow fiber membrane module.

sheet or product end of the fiber bundle, and exits through the product connection on the feed end of the permeator.

In a hollow fiber module, the permeate water flow per unit area of membrane is low, and therefore, the concentration polarization is not high at the membrane surface. The net result is that hollow fiber units operate in a non-turbulent or laminar flow regime. The HFF membrane must operate above a minimum reject flow to minimize concentration polarization and maintain even flow distribution through the fiber bundle. Typically, a single hollow fiber permeator can be operated at up to 50% recovery and still satisfy the minimum reject flow required. The hollow fiber module configuration allows a large membrane area per unit volume of permeator which results in compact systems.

Hollow fiber membrane modules are available for brackish and seawater applications. Membrane materials are cellulose acetate blends and aramid (a proprietary polyamide type material in an anisotropic form). Because of very close packed fibers and tortuous feed flow inside the module, hollow fiber modules require feed water of better quality (lower concentration of suspended solids) than the spiral wound module configuration. The hollow fiber modules are used mainly for desalting of seawater and treatment of good quality brackish water (well water). Due to fouling susceptibility of the conventional hollow fiber configuration, these module types are not used for desalting of municipal wastewater.

4.3. Spiral wound elements

The concept of the spiral wound membrane element device was introduced shortly after the invention of the hollow fiber configuration (15). In a spiral wound configuration two flat sheets of membrane are separated with a permeate collector channel material to form a leaf. The leaf assembly is sealed on three sides with the fourth side left open for permeate to exit (Fig. 4.3). A feed/brine spacer material sheet is added to the leaf assembly. A number of these assemblies or leaves are wound around a central plastic permeate tube. The permeate tube is perforated to collect the permeate from the multiple leaf assemblies (Fig. 4.3). During the element assembly process membrane leaves are rolled around the permeate tube in a spiral configuration (Fig. 4.4). The membrane leaves are kept in this form with a tape wrapped around the element and the outer shell, which is usually made of reinforced fiberglass.

The typical commercial spiral wound membrane elements are approximately 100 or 150 cm (40 or 60 in.) long and 10 or 20 cm (4 or 8 in.) in diameter (Fig. 4.6). The feed/brine flow through an element is in a straight axial path from the feed end to the opposite brine end, running parallel to the membrane surface. Fraction of the feed permeates through the membrane and flows through the permeate carrier fabrics to the central permeate tube. The remaining fraction of feed water continue to flow through the feed channel and becomes a concentrate (Fig. 4.4). The feed channel spacer is in the form of a two level (biplanar) net. The strands in each level are parallel and crossing at about 90 de-

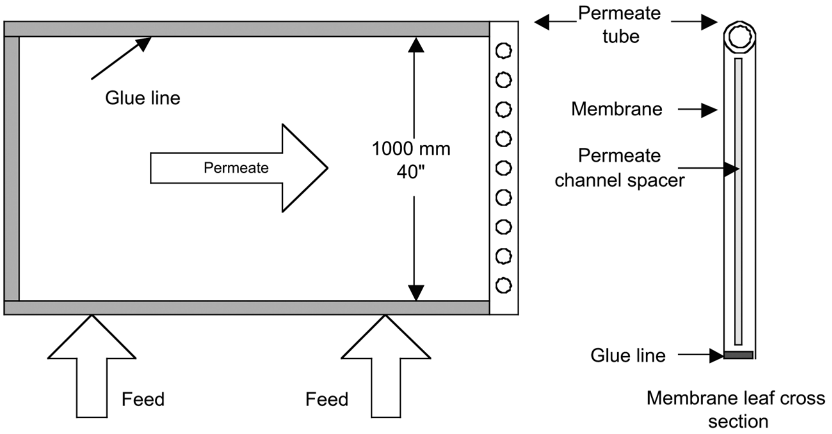


FIG. 4.3 Configuration of flat sheet membrane leaf.

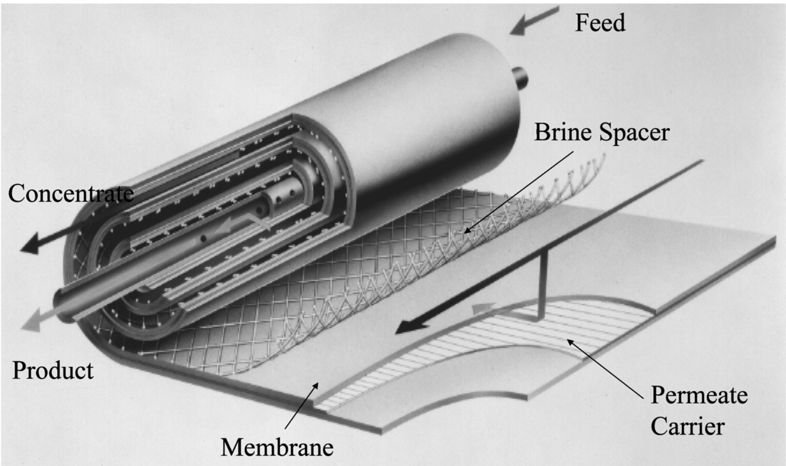


FIG. 4.4 Configuration of spiral wound membrane element.

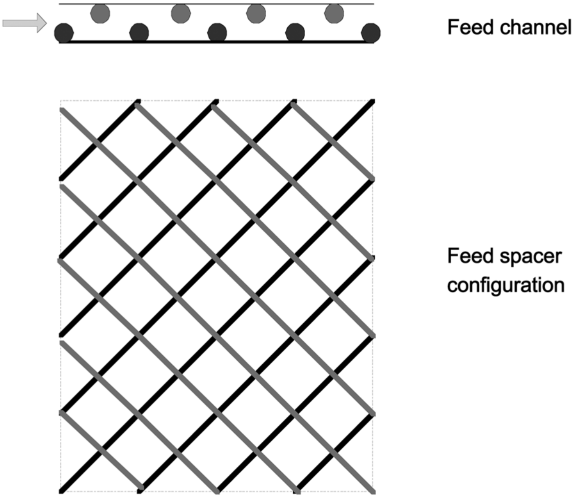


FIG. 4.5 Configuration of feed-brine spacer in a spiral wound element.

gree strands in the other level (Fig. 4.5). This two level net separates membranes from adjacent leaves and induces turbulence in the feed stream to reduce concentration polarization. The thickness of the feed channel is in the range of 0.7–09 mm (0.028–0.034″). However, the cross section of feed channel open to flow is much smaller, due to the presence of feed spacer.

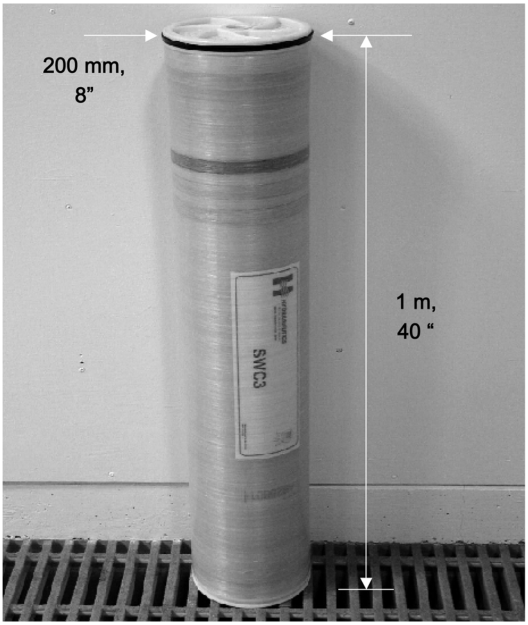


FIG. 4.6 Commercial spiral wound membrane element.

Membrane manufacturers specify concentrate flow rate requirements to control concentration polarization by limiting recovery rate (or conversion) per element to 10–20%. Therefore, recovery rate is a function of the feed-brine path length. In order to operate at acceptable recoveries, spiral systems are usually staged with three to eight membrane elements connected in series in a pressure tube (Fig. 4.7). The concentrate stream from the first element becomes the feed to the following element, and so on for each element within the pressure tube.

Each element contains brine seal, which is in the form of flexible o-ring, usually position at the front end of element. The brine seal seals the space between the element outer wall and inner wall of the pressure tube. Brine seal prevents feed water to bypass the element, which would otherwise result in low flow through element and high recovery rate. Concentrate stream from the last element exits the pressure tube to the next processing stage or to waste. The permeate tubes of each element are connected to adjacent element through permeate interconnector, forming a common permeate tube. The first and the last element in the pressure vessel is connected through an adaptor to the pressure vessel permeate outlet (Fig. 4.8). Permeate from all elements in the pressure vessel exits the vessel as a common permeate stream. A single pressure vessel

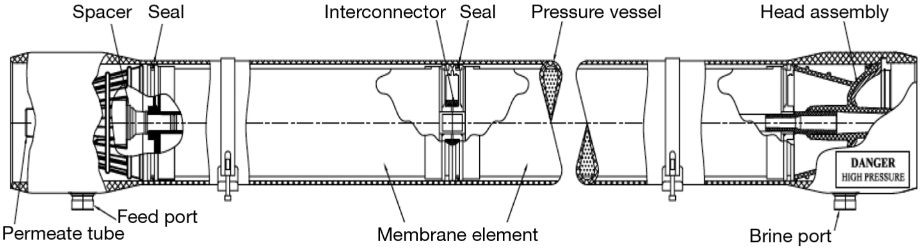


FIG. 4.7 Configuration of pressure vessel with membrane elements (courtesy of Bell Industries).

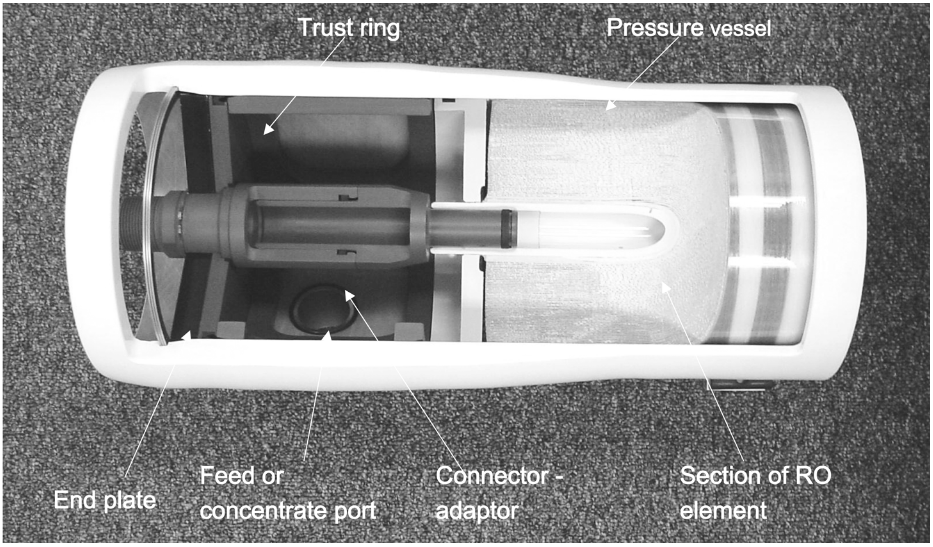


FIG. 4.8 Membrane element in pressure vessel in the first or last position (courtesy of Protec Pressure Vessels).

with six to eight membrane elements connected in series can be operated at up to 50–70% recovery under normal design conditions.

The dimensions and geometry of spiral wound membrane elements is highly standardized. Spiral wound membrane elements produced by various manufacturers are of very similar configuration and outer dimensions. They can be operated in the same pressure vessels and are easily interchangeable. More recently a committee composed of representatives of major manufacturers of spiral wound elements has been evaluating feasibility of large scale commercial production of large diameter (>200 mm, 8″) elements (63). The conclusion of the evaluation was that the optimum size of the future large elements should be 406 mm diameter by 1016 mm long (16″ by 40″ L). Such elements would have about 4 times the membrane area (and the permeate flow) of the current, 200 mm (8″) diameter element and would provide some reduction of RO system capital cost. One of the membrane manufactures (Koch Membrane Systems) provides already limited offering of 16–18″ (406 mm–457 mm) diameter membrane elements, which are used in small to medium size RO units. It is expected that by 2008 most of the major manufacturers of membrane elements will include large diameter membrane elements in their product offering.

4.3.1. Spiral wound element categories

Large desalination systems utilize membrane elements that are 1 m (40″) long and 200mm (8″) in diameter. Smaller, 100 mm diameter elements are used for small systems (light commercial, small potable) and pilot testing. The spiral wound membranes are used commercially in three major application categories: potable water softening (nanofiltration), brackish water desalting and seawater desalting. They are categorized accordingly as nanofiltration, brackish and seawater elements. Although operated at different feed pressure range, they are of almost identical configuration and utilize the same materials of construction. One of the major difference between spiral elements used in low and high feed pressure applications is feed permeate spacer. The type used in high pressure, seawater elements is of tighter type, reducing membrane embossing. Traditionally, brackish water elements were manufactured with slightly higher (about 10%) membrane area then seawater elements. At present, this difference is still maintained to some extend but membrane area of elements in all categories has increased.

The representative properties and nominal performance of nanofiltration, brackish and seawater elements are listed in Tables 1.1, 4.1, and 4.2. The nominal performances are measured during testing of a single element at nominal test conditions. The nominal test conditions include feed salinity (as NaCl), feed pressure, recovery rate and temperature (25°C). At field conditions, where operating parameters are different then the conditions during the factory tests, the elements are operating at a different performance level. In softening and brackish applications 200 mm (8″) diameter element will produce about 24 m^3/d (6400 gal/d). In seawater applications a single element of the same size will produce about 12 m^3/d (3200 gal/d). In brackish water RO systems the reduction of feed salinity will be about 97% i.e., permeate salinity will be about 3% of the feed salinity. The corresponding values for seawater systems are about

TABLE 4.1
Representative nominal performance of softening membrane elements.

Element model	Hydracore	ESNA-LF	SU620F	NF-90	NF-270
Element dimensions	1m long, 200 mm diameter (40″ × 8″)	1m long, 200 mm diameter (40″ × 8″)	1m long, 200 mm diameter (40″ × 8″)	1m long, 200 mm diameter (40″ × 8″)	1m long, 200 mm diameter (40″ × 8″)
Membrane area, m^2 (ft^2)	36.8 (400)	36.8 (400)	36.8 (400)	36.8 (400)	35 (380)
Permeate flow, m^3/d (gpd)	31.0 (8,200)	29.5 (7,800)	22.0 (5,800)	38.8 (10,000)	47.3 (12,500)
Salt rejection, %	50.0	80.0	55.0	97.0 (*)	97.0 (*)
Test feed pressure, bar (psi)	5.2 (75)	5.2 (75)	3.4 (50)	4.8 (70)	4.8 (70)
Test feed salinity, ppm NaCl	500	500	500	2000 (*)	2000 (*)
Test recovery rate, %	15	15	15	15	15
Test flux rate, l/m^2/h (gfd)	34.9 (20.5)	33.2 (19.5)	24.7 (14.5)	42.5 (25)	55.9 (32.9)
Permeability, l/m^2/h/bar (gfd/psi)	7.6 (0.31)	7.3 (0.29)	8.6 (0.35)	11.9 (0.48)	15.8 (0.63)
Relative salt transport value	1743	663	1109	128	168

(*) Na_2SO_4 used as a test solution

99.0% reduction of feed salinity or producing permeate salinity of about 1.0% concentration of the feed salinity. In softening (nanofiltrtaion) applications the actual system salt rejection will very much depend on type of elements selected. It can be as high as 90% for systems designed for salinity and hardness reduction and as low as practically insignificant for systems designed to remove color and dissolved organics only.

TABLE 4.2
Representative nominal performance of brackish membrane elements.

Element model	ESPA2+	ESPA4+	TMG20-430	BW30-LE440	BW30-LE-440
Element dimensions	1m long, 200 mm diameter (40″ × 8″)	1m long, 200 mm diameter (40″ × 8″)	1m long, 200 mm diameter (40″ × 8″)	1m long, 200 mm diameter (40″ × 8″)	1m long, 200 mm diameter (40″ × 8″)
Membrane area, m^2 (ft^2)	39.5 (430)	39.5 (430)	39.5 (430)	40.5 (440)	40.5 (440)
Permeate flow, m^3/d (gpd)	41.6 (11,000)	49.2 (13,000)	41.6 (11,000)	48.0 (12,700)	48.0 (12,700)
Salt rejection, %	99.60	99.60	99.50	99.0	99.3
Test feed pressure, bar (psi)	10.3 (150)	6.7 (100)	7.6 (110)	6.7 (100)	10.3 (150)
Test feed salinity, ppm NaCl	1500	500	500	500	2000
Test recovery rate, %	15	15	15	15	15
Test flux rate, l/m^2/h (gfd)	43.5 (25.6)	51.4 (30.2)	43.5 (25.6)	49.1 (28.9)	49.1 (28.9)
Permeability, l/m^2/h/bar (gfd/psi)	4.9 (0.20)	8.2 (0.33)	6.2 (0.25)	7.8 (0.31)	5.9 (0.24)
Relative salt transport value	17	21	22	49	34

The nominal performances of membrane elements listed in Table 4.1 through 4.3 are difficult to compare directly. The one reason being is that the nominal test conditions are somewhat different for different model elements. The test conditions differ even for the elements listed in the same application category. The other reason is that performances in field conditions are very much different than the nominal performance values. Still, the nominal values are the basis for calculation of projected system performance. The projected field performances are calculated using membrane manufactures provided computer programs. The results of computer calculations are quite accurate for brackish and seawater applications. At present, the calculated results for soften-

TABLE 4.3
Representative nominal performance of seawater membrane elements.

Element model	SWC4+	SWC5	TM820-400	SW30HR-LE	SW30-XLE
Element dimensions	1m long, 200 mm diameter (40″ × 8″)	1m long, 200 mm diameter (40″ × 8″)	1m long, 200 mm diameter (40″ × 8″)	1m long, 200 mm diameter (40″ × 8″)	1m long, 200 mm diameter (40″ × 8″)
Membrane area, m^2 (ft^2)	36.8 (400)	36.8 (400)	36.8 (400)	35.0 (380)	36.8 (400)
Permeate flow, m^3/d (gpd)	24.6 (6,500)	30.3 (8,000)	24.6 (6,500)	26.5 (7,000)	34.0 (9,000)
Salt rejection, %	99.8	99.8	99.75	99.75	99.70
Test feed pressure, bar (psi)	55.2 (800)	55.2 (800)	55.2 (800)	55.2 (800)	55.2 (800)
Test feed salinity, ppm NaCl	32,000	32,000	32,000	32,000	32,000
Test recovery rate, %	10	10	8	8	8
Test flux rate, l/m^2/h (gfd)	27.6 (16.3)	34.0 (20.0)	27.6 (16.3)	31.3 (18.4)	38.3 (22.5)
Permeability, l/m^2/h/bar (gfd/psi)	1.0 (0.04)	1.3 (0.05)	1.0 (0.04)	1.2 (0.05)	1.4 (0.06)
Relative salt transport value	6	7	7	8	11

ing applications are approximate and manual corrections based on field experience (usually pilot unit operation) have to be applied.

The membrane performance parameters that provide some insight into expected performance of membrane elements in field conditions are water permeability and relative salt transport value. The permeability is indicative of the required feed operating pressure. High permeability will results in low feed pressure required for a given flux rate during initial system operation. During the course of field operation the permeability may change due to fouling and/or membrane compaction. Relative salt transport value (RSTV), which is a product of multiplication of test flux rate by nominal salt passage, provides indication of the salt passage. For the brackish and seawater elements low RSTV's are associated with low overall salt passage.

For nanofiltration membranes, due to high salt passage there is strong interaction of various ions during the transport process. In addition, some of the nanofiltration membrane surfaces are strongly charged. Therefore, salt passage for a given mixed ions composition could be significantly different than the nominal salt passage determined using single salt solution. These issues will be discussed in more details in the chapters dealing with membrane performance calculations and system design (see chapter 6 Calculation of system performance). Performances of nanofiltration elements are discussed in details in chapter 16: Nanofiltration technology and applications.

5

RO system configuration

Configuration of RO system is affected to some extend by type of feed water being processed. Fig. 8.1 shows configuration diagram of RO system processing well water. This configuration is representative of brackish plant or seawater plants receiving feed water from wells. In such systems, feed water filtration is usually very limited, including only cartridge filtration. Systems processing surface water from open intake, brackish or seawater, require more extensive filtration of the raw water (Fig. 8.9). In such systems it may consist of single or two-stage media filtration, combined with flocculation and/or clarification. Wastewater reclamation systems utilize, almost universally, membrane pretreatment: ultrafiltration or microfiltration. Membrane pretreatment is being evaluated and extensively piloted for seawater desalination systems. However, no large capacity RO seawater plant with such pretreatment has been operated for a significant period of time so far but some systems are in an advanced design stage.

RO systems consist of the following basic equipment components:

- Feed water supply unit
- Pretreatment system
- High pressure pumping unit
- Membrane element assembly unit
- Instrumentation and control system
- Electric power supply system
- Permeate treatment and storage unit
- Cleaning unit

The configuration and operation of all components of RO system are designed to produce and maintain adequate quality of feed water to the membrane elements, maintain stable performance of all system components, produce design permeate flow and quality and maintain design process economics.

5.1. Membrane assembly unit

The membrane assembly unit (RO train) is the "heart" of the RO system. This is where the separation between water and dissolved species takes place. It consists of a stand supporting pressure vessels, interconnecting piping, and feed, permeate and concentrate manifolds. RO train also includes an instrumentation panel with local display of flow pressure and conductivities. In some systems permeate sampling panel is also included. This panel is a collecting point for permeate sampling tubing from individual pressure vessels.

Membrane elements are installed in the pressure vessels.

A pressure vessel has a permeate port on each end, located in center of the end plate, and feed and concentrate ports, located on the opposite ends of the vessel (Fig. 4.7). Each pressure vessel may contain from one to eight membrane elements connected in series. The permeate tube of the first and the last membrane element in the vessel is connected to the end plates of the pressure vessel (Fig. 4.8). Permeate tubes of all elements in the pressure vessel are connected to each other using interconnectors, practically forming one long permeate pipe inside the pressure vessel. On one end of each membrane element there is a brine seal (Fig. 4.7). The brine seal, in the form of V-shape o-ring, closes the passage between outside rim of the element and inside wall of the pressure vessel. This seal prevents feed water from bypassing the membrane module, and forces it to flow through the feed channels of the element. As feed water flows through each subsequent membrane element part of the feed volume passes through the membrane, and is removed as a permeate. The salt concentration of the remaining feed water increases along the pressure vessel (more extensive discussion on permeate salinity distribution is included in Chapter 5.5). Permeate tubes conduct the permeate from all connected elements. The collected permeate has the lowest salinity at the feed end of the pressure vessel, and increases gradually in the direction of the concentrate flow.

The RO system is divided into groups of pressure vessels, called concentrate stages. In each stage pressure vessels are connected in parallel, with respect to the direction of the feed/concentrate flow. The number of pressure vessels in each subsequent stage decreases in the direction of the feed flow. The configuration is usually in the ratio of 2:1. Thus, one can visualize that the flow of feed water through the pressure vessels of a system resembles a pyramid structure: a high volume of feed water flows in at the base of pyramid, and a relatively small volume of concentrate leaves at the top. The decreasing number of parallel pressure vessels from stage to stage compensates for the decreasing volume of feed flow, which is continuously being partially converted to permeate. Permeate from all pressure vessels in each stage, is combined together into a common permeate manifold.

The objective of the taper configuration of pressure vessels is to maintain a similar feed/concentrate flow rate per vessel through the length of the system and to maintain feed/concentrate flow within the limits specified for a given type of membrane element. Very high flow through a pressure vessel will result in a high pressure drop and possible structural damage of the element. Very low flow will not provide sufficient turbulence, and may result in excessive salt concentration at the membrane surface. The limits of maximum feed flow and minimum concentrate flow are specified by membrane manufacturers for a given membrane element type depending mainly on combined height of the feed channels in the element and type of feed spacer net used.

Pressure vessels used in RO commercial systems are highly engineering products, especially design for RO applications. Configuration of pressure vessels manufactured to house spiral wound elements is highly standardized in respect of port-to-port dimensions, internal length and inside diameter. In this respect they are interchangeable, i.e., almost any commercial spiral wound element made by varius manufacturers will fit into all commercial pressure vessel of any manufacturer. Materials of construction of pressure vessels are adapted to high salinity corrosive environment. Pressure vessel tube is made of epoxy based FRP materials. Feed and concentrate ports are made of corrosion resistant stainless steel. Pressure vessels are manufactured in accordance to ASME code. Maximum pressure rating of pressure vessels is according to application category. For brackish applications pressure vessels are rated up to 40 bar (600 psi). Pressure rating of pressure vessels for seawater application is up to 83 bar (1200 psi). For nanofiltrtaion applications, a low pressure less expensive, products are also available.

5.2. Concentrate staging

A commercial RO unit usually consists of single pump and a multistage array of pressure vessels. A simplified block diagram of a two stage RO unit is shown in Fig. 5.1. The concentrate from the first stage becomes the feed to the

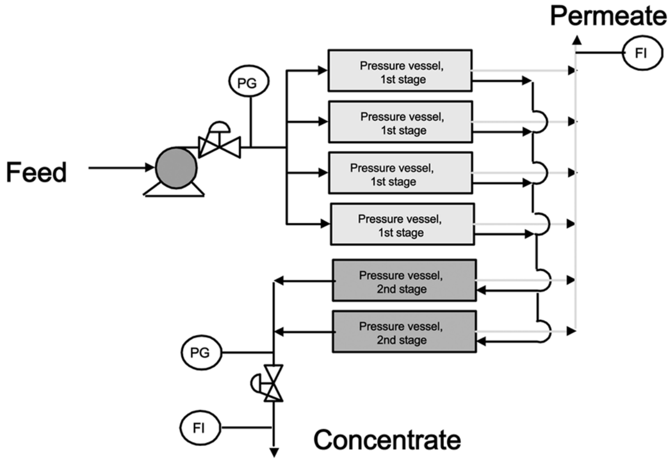


FIG. 5.1 Configuration of two stage RO system.

second stage; this is what is meant by the term "concentrate staging." The flows and pressures in the multistage unit are controlled with the feed and concentrate valves. The feed valve, after the high pressure pump, controls feed flow to the unit. The concentrate valve, at the outlet of RO block, controls feed pressure inside the unit.

For a given RO unit, the number of concentrate stages will depend on the permeate recovery ratio and the number of membrane elements per pressure vessel. In order to avoid excessive concentration polarization at the membrane surface, the recovery rate per individual membrane element should not exceed 18%. It is common engineering practice to design brackish RO systems so that the average recovery rate per 40-inch-long membrane element will be about 6–8%. Accordingly, the number of concentrate stages for an RO unit having 6 elements per pressure vessel would be two stages for recovery rates over 60%, and three stages for recovery rates over 75%. With pressure vessels containing seven to eight elements, a two stage configuration would be sufficient for recovery rates up to about 85%.

Figure 5.2 shows drawing of an RO train in a single stage configuration. Feed, permeate and concentrate manifold are clearly indicated.

Figure 5.3 shows a picture of commercial, two stage, brackish train. The array is 32:14 pressure vessels with 7 elements per vessel. The picture shows two parallel feed manifold with 4 × 8 pressure vessel connected. Unit configuration is eight vessels high and six vessels wide. This translates into unit dimensions of 4.0 m high, 2.9 m wide and 8.0 m long (13.1′ × 9.5′ × 26′). The first

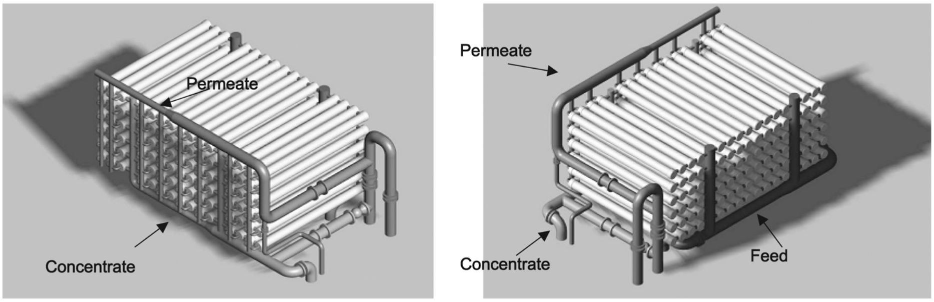


FIG. 5.2 Configuration of a single stage RO train. Courtesy of CH_2M Hill.

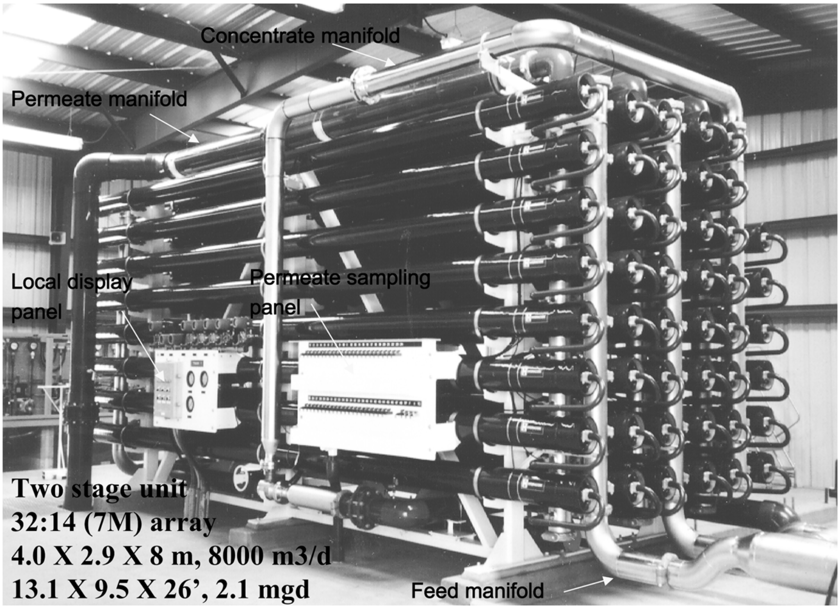


FIG. 5.3 Picture of a two stage RO train.

stage concentrate is collected from the 32 first stage pressure vessels and fed to 14 second stage vessels. The concentrate outlets from the second stage vessels are connected to the concentrate manifold. The concentrate throttling valve located on the concentrate pipe is shown as well. The train is equipped with permeate sampling panel, which enables sampling permeate conductivity from individual vessels and local display panel of flow, pressure, feed temperature and conductivities.

The unit contains total of 322 membrane elements, each 200 mm (8″) diameter, 1000 mm (40″) long, and is capable of about 8,000 m^3/d (2.1 mgd) of permeate at recovery rate of 80%. The unit shown on Fig. 5.3 is equipped with pressure vessels that have one feed and one concentrate port. This leads to train configuration where two pressure vessels are connected to the feed or concentrate manifold at the given level, on each side. New version of pressure vessels, being introduced recently, are of multiport design. The "multiport" vessels have two large, entry-exit ports, at each end of pressure vessels. This configuration enables pressure vessel to be connected to feed or concentrate manifold and then to each other horizontally (or vertically), practically creating horizontal (or vertical) feed and concentrate lines (Fig. 5.4). Feed flow starts from feed manifold and then flows to the group of pressure vessels connected in parallel. The same situation, with reverse flow direction, takes place at the concentrate end of the train. Currently, the number of "multiport" pressure vessels that pressure vessel manufacturers recommend to connect together is five. Therefore number of pressure vessels connected to feed or concentrate manifold at given level could be increased to ten vessels (five vessels on each side of feed or concentrate pipe). This type of pressure vessels enables train design with lower requirement of high pressure piping, contributing to reduction of equipment cost.

5.3. Flow distribution

In some cases it is necessary to equilibrate permeate flow between stages, i.e., decrease permeate flow from the first stage and increase permeate flow from the last stage. This can be accomplished in one of two design configurations. One solution is to install a valve on the permeate line from the first stage. By

FIG. 5.4 Array of "multiport" pressure vessels (courtesy of Protec Pressure Vessels).

throttling this valve, permeate back pressure will increase, reducing net driving pressure and reducing permeate flux from the first stage (see Eq. 8). To compensate for lower permeate flow from the first stage the differential permeate flow is produced from the second stage by operating the RO unit at a higher feed pressure then would be required without permeate throttling. The other solution is to install a booster pump on the concentrate line between the first and the second stage (Fig. 5.5). The booster pump will increase feed pressure to the second stage. This configuration will result in lower permeate flow from the first stage and higher permeate flow from the second stage,, i.e., more uniform permeate flux distribution. The advantage of the permeate throttling design is simplicity of the RO unit configuration and lower capital cost. However, this design results in additional power losses due to permeate throttling and higher power consumption. The interstage pump design requires modification of the interstage manifold and an additional pumping unit. The investment cost is higher than in the first design configuration, but the power consumption is lower.

5.4 Permeate staging (two pass systems)

For some applications, the single pass RO system may not be capable of producing permeate water of a required salinity. Such conditions could be encountered in two types of RO applications:

- Seawater RO systems, which operate on a very high salinity feed water, at high recovery ratio and/or at high feed water temperature.
- Brackish RO applications which require very low salinity permeate such as supply of makeup water for pressure boilers or production of rinse water for microelectronics applications.

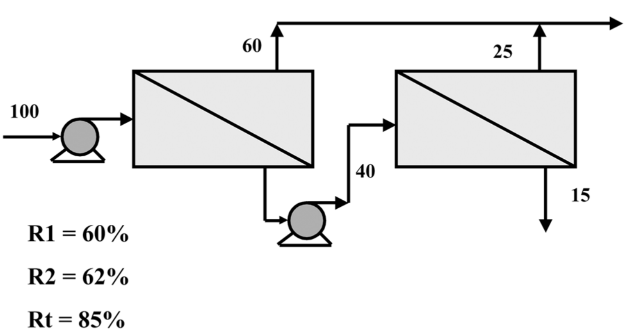


FIG. 5.5 Configuration of two stage RO unit with interstage booster pump.

To achieve an additional reduction in permeate salinity, the permeate water produced in the first pass is processed again in a second RO system. This configuration is called a two pass design, or “permeate staging.” Fig. 5.6 contains schematic diagram of a two pass system.

Depending on permeate quality requirements, part or all of the first pass permeate volume is desalted again in the second pass unit. The system configuration is known as a complete or partial two pass system according to whether all of the 1st pass permeate or only some fraction is fed to the second pass unit. The first pass permeate is a very clean water. It contains very low concentrations of suspended particles and dissolved salts. Therefore, it does not require any significant pretreatment. The second pass RO unit can operate at a relatively high average permeate flux and high recovery rate without concerns of concentration polarization and scaling. The common design parameters for the second pass RO unit are average flux rate of 34 $l/m^2/h$ (20 gfd) and recovery rate of 85–90%. In a two pass system, the permeate from the first pass flows through a storage tank or is fed directly to the suction of the second pass high pressure pump. It is a common procedure in a two pass seawater RO systems to return concentrate from the second pass unit to the suction of the high pressure pump of the first pass unit. The dissolved salts concentration in the concentrate from the second pass is usually lower the concentration of the feed to the first pass unit. Therefore, blending feed water to the first pass with small flow rate of the second pass concentrate, reduces slightly the salinity of the feed to the first pass, and increases the overall utilization of the feed water (Fig. 5.7).

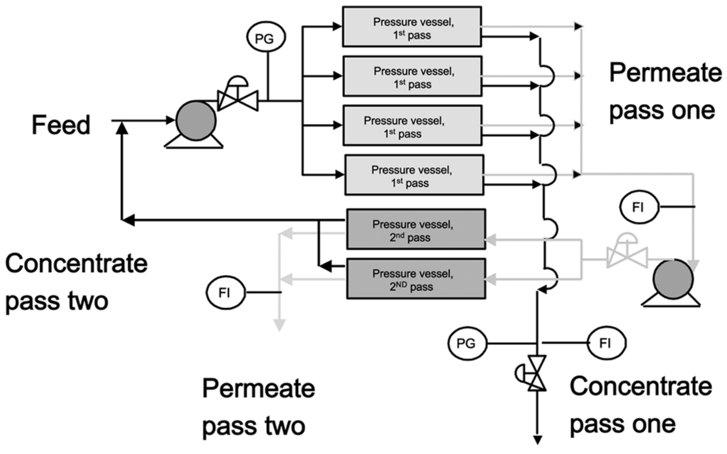
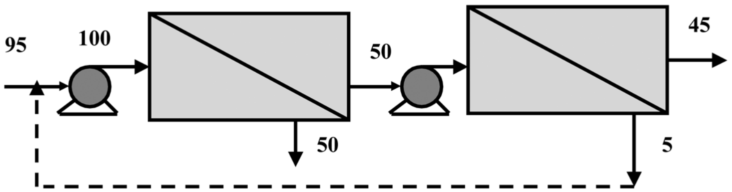


FIG. 5.6 Configuration of a two pass RO unit.



R1 = 50.0 %

R2 = 90.0 %

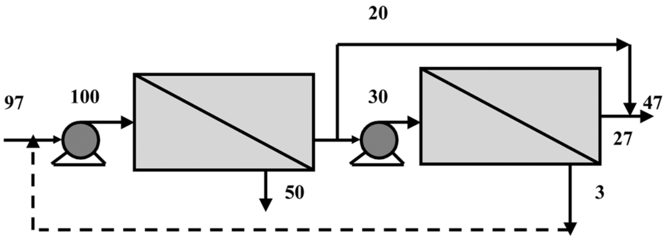
Rt = 45.0% (w/o recirculation)

Rt = 47.3% (w recircirculation)

FIG. 5.7 Recovery rate in a two pass RO unit.

5.5. Partial two pass configuration

There are number of possible configuration of the two pass RO units. One configuration, which is a partial two pass system, is shown in Fig. 5.8. In this configuration the first pass permeate is split into two streams. One stream is processed by the second pass unit, and it is then combined with the unprocessed part of the permeate from the first pass. Provided that the partial two pass system can produce the required permeate quality, this configuration results in smaller second pass unit, therefore lower capital and operating costs, as well as



R1 = 50.0%

R2 = 90.0%

Rt = 47.0% (w/o recirculation)

Rt = 48.4% (w recirculation)

FIG. 5.8 Recovery rate in a partial two pass RO unit.

higher combined permeate recovery rate (utilization of the feed water), compared to a complete two pass system.

Another partial two pass configuration, which takes advantage of internal salinity distribution of RO permeate in the pressure vessel, is becoming increasingly popular in seawater systems. This unique concept of partial two pass system, designated as "split partial", has been proposed in the past (51) but only lately is being implemented in large seawater systems (47, 52). In the RO system each subsequent element, in the direction of feed flow, produces water of increasingly higher salinity (Fig. 5.9). This is due to increasing feed salinity along the pressure vessel and decreasing permeate flux (Fig. 5.10) affected by the decrease of NDP. The NDP decrease results from increasing osmotic pressure and gradual decrease of feed pressure along the pressure vessel, due to friction losses. As the elements are connected together through the permeate tube, the permeate from the individual elements is mixed together and leaves pressure vessel through permeate port at composite salinity.

In a split partial configuration the first pass system is configured as a single stage unit. The permeate is collected from both ends of pressure vessels. The fraction collected from the feed end is of lower salinity and flows directly to permeate storage tank. The fraction collected at the concentrate end is processed with the second pass RO unit (Fig. 5.11). The salinity distribution between the two fractions of first pass permeate depends on their relative flow and

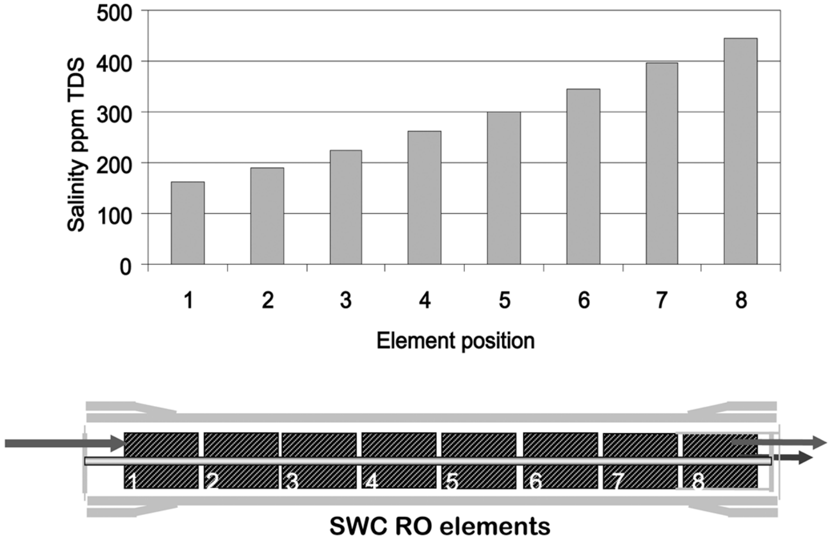


FIG. 5.9 Combined permeate salinity vs. element position in a pressure vessel.

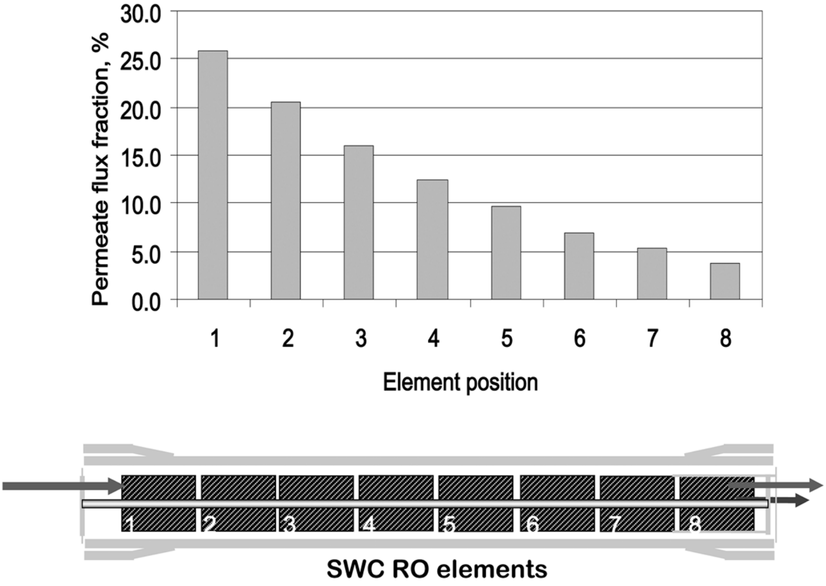


FIG. 5.10 Flux distribution from individual elements vs. element position in a pressure vessel.

is shown in Fig. 5.12. As illustrated in Fig. 5.12, the salinity of both fractions increases, as the fraction taken from the low salinity end (feed end) increases. In other words, the salinity of the fraction to be processed by the second pass RO increases as the permeate flow rate taken from the concentrate side decreases. This is also the direction of increasing effectiveness of the split partial process in reduction of the overall system size and power consumption. The

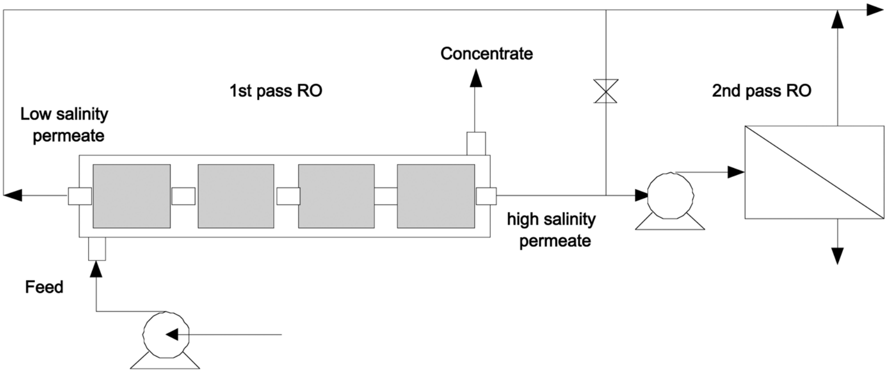


FIG. 5.11 Configuration of a “split partial” two pass unit.

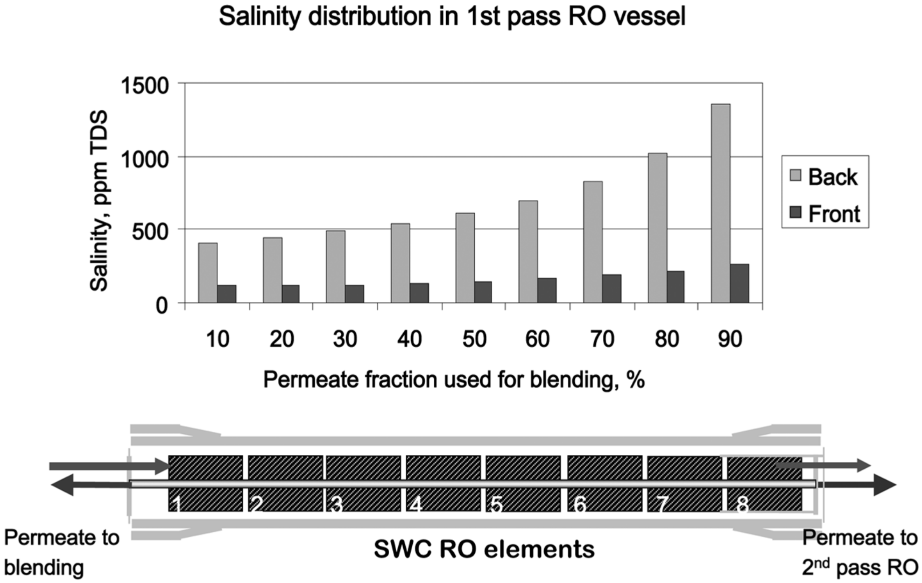


FIG. 5.12 Salinity distribution in a "split partial" two pass unit as a function of permeate flow distribution.

smaller the fraction of the first pass permeate that has to be processed with the second pass RO the higher the benefits of "split partial"

The objective of the second pass unit is to remove salt (reduce TDS) from the first pass permeate stream. It is intuitively obvious that to achieve given salinity reduction, it will be more effective to process small flow of high salinity rather the large flow of low salinity. Therefore, the "split partial" process takes advantage of internal separation of permeate salinity in the RO unit. In the conventional partial two pass configuration, after the permeate leaves the pressure vessel it has a uniform salinity and additional work has to be expended to reverse the mixing and achieve the separation.

Table 5.1 shows design parameters for seawater two pass system operating in a split partial configuration. The table lists the fraction of the high salinity permeate flow (fraction leaving pressure vessel at the concentrate side) that would be processed by the second pass RO and corresponding salinities of the low salinity, high salinity and blended stream.

For comparison the last column lists the fraction of second pass RO required to achieve the same blended flow salinity in conventional partial two

TABLE 5.1
Partial two pass system design. Feed salinity 40,200 ppm TDS, 28°C

Seawater split partial two pass system. First pass feed: 40, 120 ppm TDS, 50% recovery. Second pass 90% recovery					Conventional two pass
High salinity flow fraction (feed to 2nd pass RO), %	2nd pass capacity, %	Low salinity fraction, ppm TDS	High salinity, ppm TDS	Blended flow salinity ppm TDS	2nd pass RO capacity required,%
90	81	108	351	18	87
80	72	108	383	31	84
70	63	113	422	43	81
60	54	125	469	60	76
50	45	134	532	78	72
40	36	152	609	102	65
30	27	173	717	131	57
20	18	203	879	172	47
10	9	249	1149	232	30

pass system. Two important observations could be concluded from the presented results:

1. Split partial configuration requires smaller second pass system capacity than the conventional partial two pass unit.
2. The capacity difference between this two configurations decreases with increased flow being processed by the second pass unit.

The flow rate processed by the second pass RO unit affects required capacity of the primary RO. With increased capacity of the secondary RO larger fraction of primary RO permeate is discharged as concentrate of the secondary RO. This has to be compensated by increased capacity of the primary RO. Therefore, it affects both the capital and operating cost (mainly power consumption but also some contribution from chemical consumption and membrane replacement cost). The benefits of split partial configuration have to be evaluated against somewhat higher cost resulting from additional permeate piping and more complex process control. Table 5.2 provides comparison of the membrane

TABLE 5.2
Comparison of design parameters for split partial and conventional partial two pass system configurations

	First pass	Second pass
Conventional design		
Permeate flow, m^3/h	4,312.5	1958.3
Processed by second pass, %	45.4	
Recovery ratio, %	50.0	90.0
Feed pressure, bar	65.4	14.1
No. of pressure vessels	1030	190
No. of elements	8,240	1,520
Power requirement, kWh	12,525	826
Combined power req., kWh	13,351	
Split partial design		
Permeate flow, m^3/h	4,166	62.5
Processed by second pass, %	18.9	
Recovery ratio, %	50	90.0
Feed pressure, bar	66.5	13.4
No. of pressure vessels	1,000	60
No. of elements	8,000	480
Power requirement, kWh	12,307	321
Combined power req., kWh	12,628	
Difference of no. of elements, (%)	240 (2.9)	1,040 (68)
Power saving, kWh (%)	723 (5.4)	

arrays of commercial RO seawater system of 100,000 m^3/day (26.45 mgd) capacity, for split partial and conventional partial two pass configuration. For this particular design the split partial configuration results in approximately 13% savings of number of RO elements and pressure vessels and about 5% reduction of power consumption.

6

Calculation of system performance

6.1. Manual method of membrane system performance calculations

The common approach to projecting performance of RO system is to calculate permeate flow according to the net driving pressure model and base calculations of permeate salinity on salinity gradient between feed and permeate as a driving force of the salt transport. The reference conditions are the nominal element performance, as tested at standard test conditions, defined by membrane manufactures. Single point calculations for basic system configuration can be conducted manually (as shown below). However repeated calculations, required for optimization of process design, are conducted using computer programs available from all major membrane manufacturers. The manual calculations process goes through the following steps:

1. According to the type of feed water source select membrane element type to be used and the value of system average permeate flux (APF).
2. Using nominal test conditions and nominal element performance calculate specific permeability of the selected membrane element (SP).
3. Using the above values of APF and SP calculate the required net driving pressure (Eq. 18).
4. Based on feed water composition, feed water type or project specifications select system recovery ratio and calculate average feed salinity (Eq. 6 or 7).
5. Calculate corresponding average feed osmotic pressure (Equation 2 or salinity–osmotic pressure relations).

6. Make assumptions regarding system array, pressure drop per stage and permeate back pressure.
7. Calculate required feed pressure (Eq. 8).
8. Calculate permeate salinity based on average feed salinity, average system permeate flux, nominal element salt passage and element permeate flux at nominal test conditions.

Example 11

Brackish two stage RO system. Feed salinity 2500 ppm TDS. Recovery rate: 85%. Average flux rate 27.2 $l/m^2/h$ (16 gfd). Feed water temperature 25°C.

Calculations of specific element performance
Element type: ESPA2, membrane area: 36.8 m^2
Nominal element performance: 34.07 m^3/d @ pressure 10.3 bar
Salt rejection 99.6% @ flux rate 38.6 l/m^2h (22.7 gfd)
Nominal test conditions: feed salinity 1500 ppm NaCl, recovery rate 15%
Average feed salinity during nominal test: $1500 \times 0.5\ (1 + 1/(1 - 0.15)) =$ 1632 ppm NaCl
Average feed osmotic pressure: 1.25 bar (18.1 psi)
Nominal NDP: $10.3 - 1.25 = 9.05$ bar (131.2 psi)
Specific permeability: $38.6/9.05 = 4.26$ $l/m^2/h/bar$ (0.17 gfd/psi)

Calculation of system performance
System NDP required: $27.2/4.26 = 6.4$ bar (93 psi)
Friction pressure drop per stage 2 bar (29 psi), total for system 4 bar (58 psi).
Permeate back pressure 0.5 bar (7.2 psi).
Feed salinity 2500 ppm TDS, osmotic pressure 1.9 bar (28 psi)
Average feed osmotic pressure: $1.9 \times 0.5\ (1 + 1/(1 - 0.85)) = 7.3$ bar (105.8 psi)
Required system feed pressure: $6.4 + 7.3 + 4 + 0.5 = 18.2$ bar (264 psi)

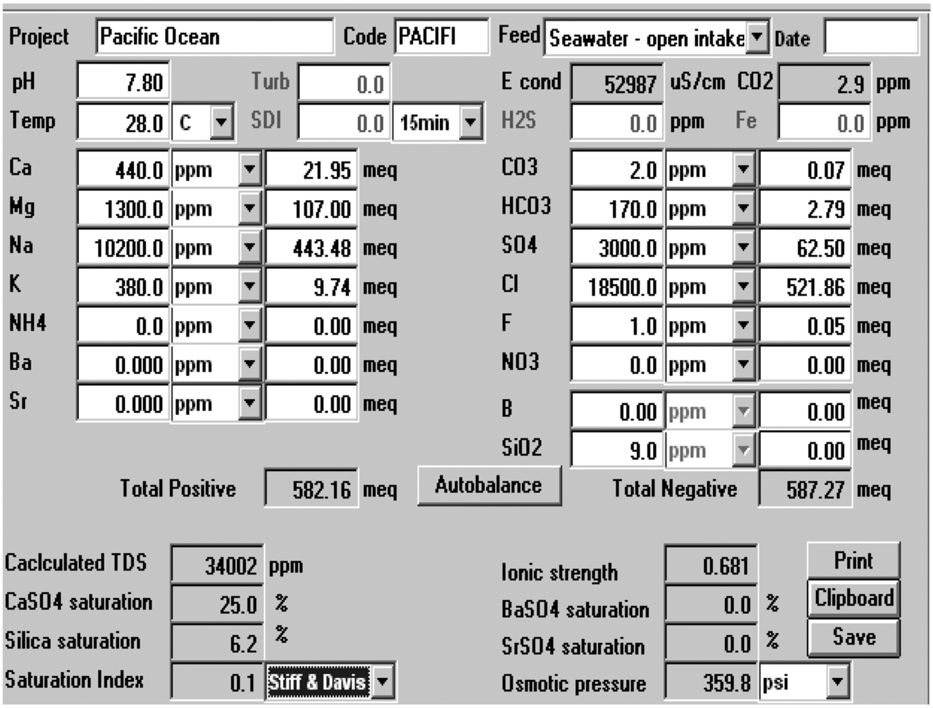
Permeate salinity
Average feed salinity: $0.5 \times (2500 + 2500/(1 - 0.85)) = 9583$ ppm
Permeate salinity: $9583 \times (1 - 99.6/100) \times (38.6/27.2) = 54$ ppm

Permeate salinity is function of average feed salinity and operating permeate flux rate as compared to the nominal flux (Eq. 13).

Additional corrections that should be applied to these calculations include correction for system configuration (flux distribution), temperature and element age. The above calculations are close approximation for two stage system treating brackish water. However, they were conducted treating a whole system as an average element. Therefore, significant errors can be made when manually calculating RO unit performance at extreme values of feed salinity, temperature or permeate recovery. Computer programs have built in routine to provide adjustment for the above mentioned parameters. In seawater systems the apparent permeability is strongly affected by concentration polarization at high salinities. Therefore, additional "salinity" correction has to be applied in calculations of NDP and feed pressure.

6.2. Use of computer programs for projections of membrane performance

To enable large number of calculations required during design of RO systems, computer programs have been developed by membrane manufacturers. Various performance projection programs are quite similar in functionality, design of user interface, input values required and output format. The calculation procedure usually starts with input of feed water analysis (Fig. 6.1). After



Project Pacific Ocean
Code PACIFI
Feed Seawater - open intake
Date

pH 7.80
Temp 28.0 C
Turb 0.0
SDI 0.0 15min
E cond 52987 uS/cm
H2S 0.0 ppm
CO2 2.9 ppm
Fe 0.0 ppm

Ca	440.0 ppm	21.95 meq		CO3	2.0 ppm	0.07 meq
Mg	1300.0 ppm	107.00 meq		HCO3	170.0 ppm	2.79 meq
Na	10200.0 ppm	443.48 meq		SO4	3000.0 ppm	62.50 meq
K	380.0 ppm	9.74 meq		Cl	18500.0 ppm	521.86 meq
NH4	0.0 ppm	0.00 meq		F	1.0 ppm	0.05 meq
Ba	0.000 ppm	0.00 meq		NO3	0.0 ppm	0.00 meq
Sr	0.000 ppm	0.00 meq		B	0.00 ppm	0.00 meq
				SiO2	9.0 ppm	0.00 meq

Total Positive 582.16 meq
Total Negative 587.27 meq

Caclculated TDS 34002 ppm
CaSO4 saturation 25.0 %
Silica saturation 6.2 %
Saturation Index 0.1 Stiff & Davis
Ionic strength 0.681
BaSO4 saturation 0.0 %
SrSO4 saturation 0.0 %
Osmotic pressure 359.8 psi

FIG. 6.1 Computer projection program—feed water analysis entry screen.

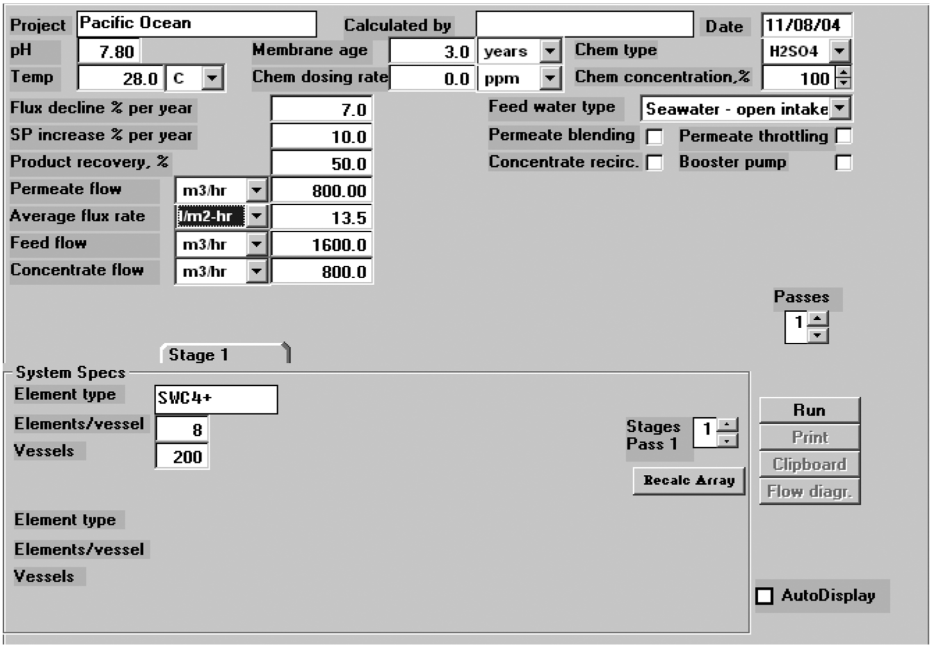
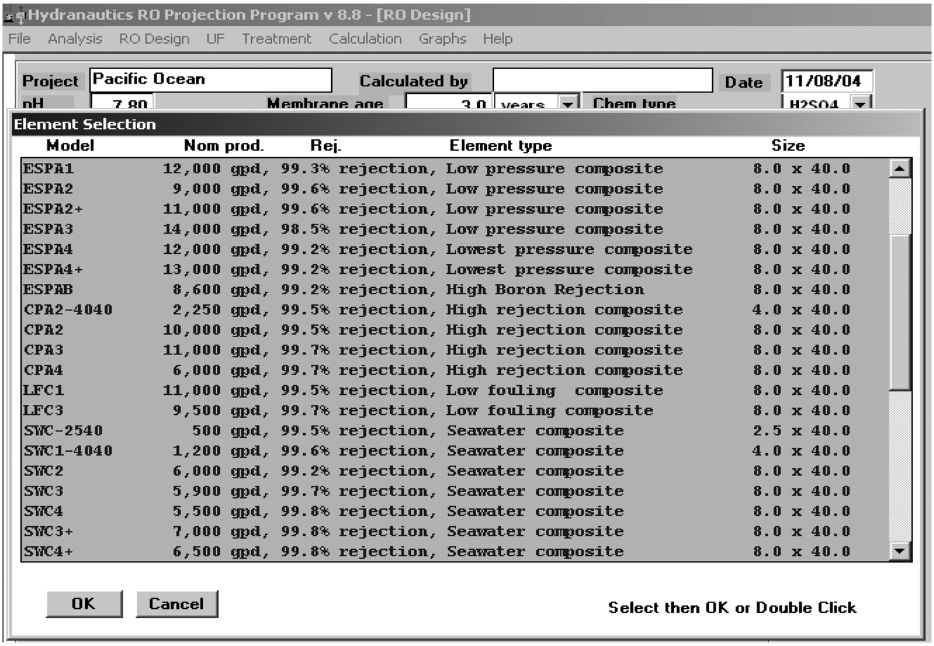

Project: Pacific Ocean
Calculated by:
Date: 11/08/04
pH: 7.80
Membrane age: 3.0 years
Chem type: H2SO4
Temp: 28.0 C
Chem dosing rate: 0.0 ppm
Chem concentration,%: 100
Flux decline % per year: 7.0
SP increase % per year: 10.0
Product recovery, %: 50.0
Feed water type: Seawater - open intake
Permeate blending ☐
Permeate throttling ☐
Concentrate recirc. ☐
Booster pump ☐
Permeate flow: m3/hr 800.00
Average flux rate: l/m2-hr 13.5
Feed flow: m3/hr 1600.0
Concentrate flow: m3/hr 800.0
Passes: 1
Stage 1
System Specs
Element type: SWC4+
Elements/vessel: 8
Vessels: 200
Stages Pass 1: 1
Element type
Elements/vessel
Vessels
☐ AutoDisplay

FIG. 6.2 Computer projection program—single pass unit design parameters entry screen.

completing input to the water analysis screen the next screen includes entry of process and system design parameters (Fig. 6.2).These include:

- Feed temperature
- Feed water pH
- Membrane age
- Permeate flow
- Recovery rate
- Membrane model
- Membrane array
- Salt passage increase factor
- Flux decline coefficient

The element type is selected from elements look up table (Fig. 6.3).



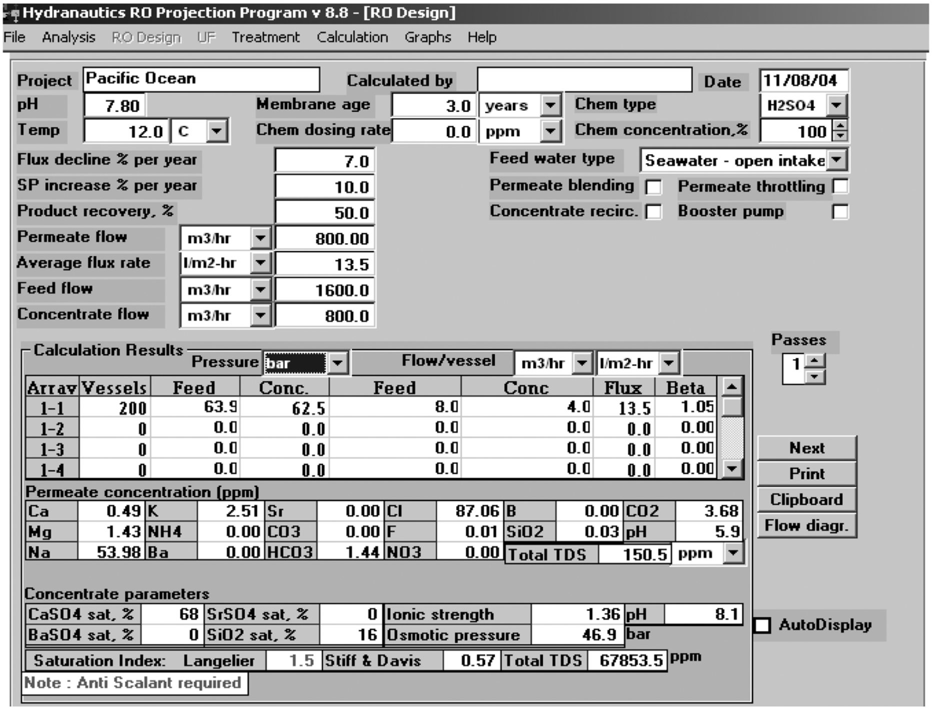
Project: Pacific Ocean
Calculated by:
Date: 11/08/04
pH: 7.80
Membrane age: 3.0 years
Chem type: H2SO4

Model	Nom prod.	Rej.	Element type	Size
ESPA1	12,000 gpd,	99.3% rejection,	Low pressure composite	8.0 x 40.0
ESPA2	9,000 gpd,	99.6% rejection,	Low pressure composite	8.0 x 40.0
ESPA2+	11,000 gpd,	99.6% rejection,	Low pressure composite	8.0 x 40.0
ESPA3	14,000 gpd,	98.5% rejection,	Low pressure composite	8.0 x 40.0
ESPA4	12,000 gpd,	99.2% rejection,	Lowest pressure composite	8.0 x 40.0
ESPA4+	13,000 gpd,	99.2% rejection,	Lowest pressure composite	8.0 x 40.0
ESPAB	8,600 gpd,	99.2% rejection,	High Boron Rejection	8.0 x 40.0
CPA2-4040	2,250 gpd,	99.5% rejection,	High rejection composite	4.0 x 40.0
CPA2	10,000 gpd,	99.5% rejection,	High rejection composite	8.0 x 40.0
CPA3	11,000 gpd,	99.7% rejection,	High rejection composite	8.0 x 40.0
CPA4	6,000 gpd,	99.7% rejection,	High rejection composite	8.0 x 40.0
LFC1	11,000 gpd,	99.5% rejection,	Low fouling composite	8.0 x 40.0
LFC3	9,500 gpd,	99.7% rejection,	Low fouling composite	8.0 x 40.0
SWC-2540	500 gpd,	99.5% rejection,	Seawater composite	2.5 x 40.0
SWC1-4040	1,200 gpd,	99.6% rejection,	Seawater composite	4.0 x 40.0
SWC2	6,000 gpd,	99.2% rejection,	Seawater composite	8.0 x 40.0
SWC3	5,900 gpd,	99.7% rejection,	Seawater composite	8.0 x 40.0
SWC4	5,500 gpd,	99.8% rejection,	Seawater composite	8.0 x 40.0
SWC3+	7,000 gpd,	99.8% rejection,	Seawater composite	8.0 x 40.0
SWC4+	6,500 gpd,	99.8% rejection,	Seawater composite	8.0 x 40.0

FIG. 6.3 Computer projection program—element selection look up table.

The array and number of elements are manually adjusted targeting permeate flux to be within the range suitable for the feed water type being processed. Computer calculations start with calculation of specific flux for the selected element. This value is corrected for temperature and element age (or fouling factor). Similar corrections are applied to nominal salt rejection (salt passage). Next, the program conducts calculations of permeate flow and salinity for a single element, in a element by element mode, staring with the lead element and progressing toward the concentrate end of the system.

The calculations are based on projected feed salinity to a given element, specific permeability and value of feed pressure selected from some predefined range. The resulting permeate flow is used to calculate the recovery rate. The calculated recovery rate is used in turn to correct the average feed salinity, which again is used to calculate new permeate flow and permeate salinity. Then the subsequent results of two calculations of permeate flow are compared, and if the difference do not exceed preset value, the calculations are conducted for the next element. After completing calculations of permeate flow and permeate salinity for all elements in series, and combining them as the corresponding cumulative



Project	Pacific Ocean	Calculated by		Date	11/08/04
pH	7.80	Membrane age	3.0 years	Chem type	H2SO4
Temp	12.0 C	Chem dosing rate	0.0 ppm	Chem concentration,%	100

Flux decline % per year		7.0
SP increase % per year		10.0
Product recovery, %		50.0
Permeate flow	m3/hr	800.00
Average flux rate	l/m2-hr	13.5
Feed flow	m3/hr	1600.0
Concentrate flow	m3/hr	800.0

Feed water type: Seawater - open intake
Permeate blending ☐ Permeate throttling ☐
Concentrate recirc. ☐ Booster pump ☐
Passes 1

Calculation Results
Pressure bar Flow/vessel m3/hr l/m2-hr

Array	Vessels	Feed	Conc.	Feed	Conc	Flux	Beta
1-1	200	63.9	62.5	8.0	4.0	13.5	1.05
1-2	0	0.0	0.0	0.0	0.0	0.0	0.00
1-3	0	0.0	0.0	0.0	0.0	0.0	0.00
1-4	0	0.0	0.0	0.0	0.0	0.0	0.00

Permeate concentration (ppm)

Ca	0.49	K	2.51	Sr	0.00	Cl	87.06	B	0.00	CO2	3.68
Mg	1.43	NH4	0.00	CO3	0.00	F	0.01	SiO2	0.03	pH	5.9
Na	53.98	Ba	0.00	HCO3	1.44	NO3	0.00	Total TDS	150.5	ppm	

Concentrate parameters

CaSO4 sat, %	68	SrSO4 sat, %	0	Ionic strength	1.36	pH	8.1
BaSO4 sat, %	0	SiO2 sat, %	16	Osmotic pressure	46.9	bar	

Saturation Index: Langelier 1.5 Stiff & Davis 0.57 Total TDS 67853.5 ppm

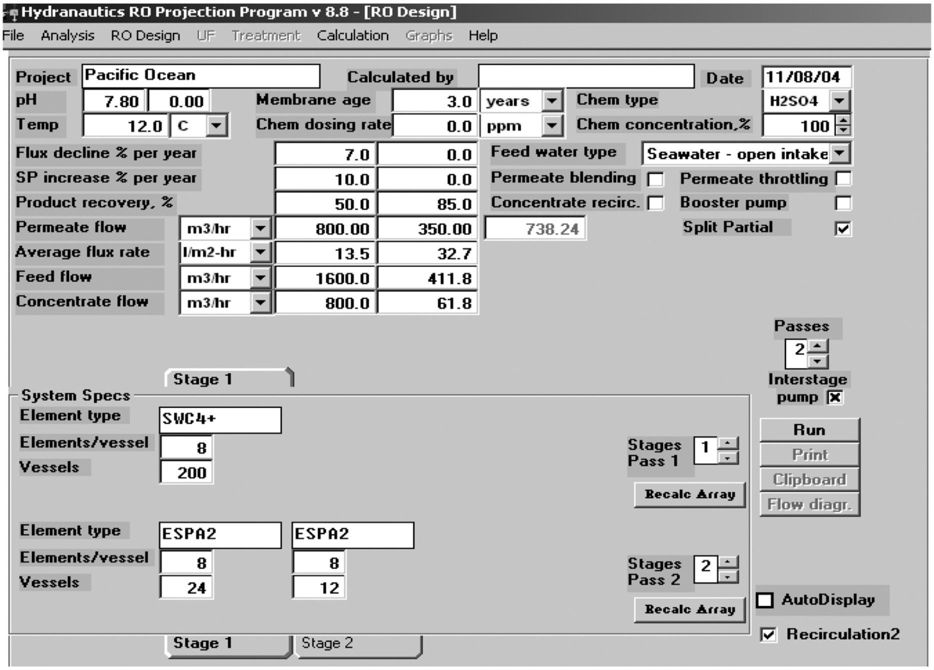
Note : Anti Scalant required

☐ AutoDisplay

FIG. 6.4 Computer projection program—performance calculation results display screen.

values, the resulting permeate flow is compared with the input value. If difference is too large, new calculations are conducted with a new feed pressure range, adjusted accordingly to the difference between previous calculations and the input value of permeate flow. If the cumulative calculated value is sufficiently close to the input value, the calculations are completed and the results are displayed on the screen. The display (Fig. 6.4) includes major input values and results of calculations of pressures, flows, some more critical saturation values and composition of permeate.

At some site conditions it is necessary to design seawater system as partial two pass configuration. Most computer programs are flexible enough to enables calculations for multipass-multistage system configurations (Fig. 6.5). Additional configurations options may include permeate blending, permeate throttling, concentrate recirculation and interstage booster pump. Some programs may have auxiliary routines that enable calculation of power consumption associated with projected operating parameters (Fig. 6.6), calculation of operating cost or conduct simulation of chemical treatment of permeate.



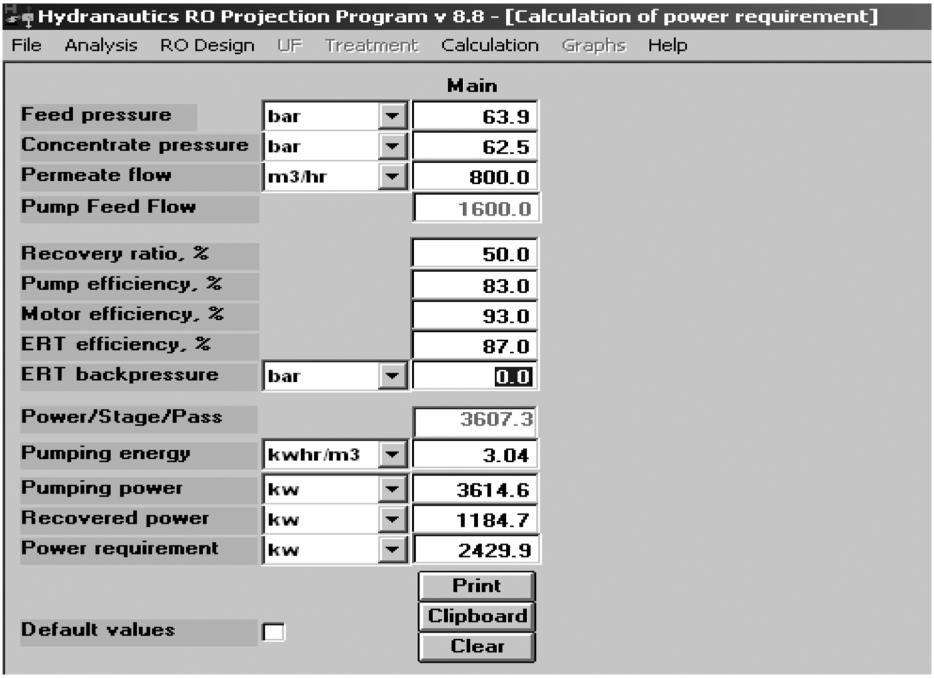
Project: Pacific Ocean
Calculated by:
Date: 11/08/04
pH: 7.80 0.00
Membrane age: 3.0 years
Chem type: H2SO4
Temp: 12.0 C
Chem dosing rate: 0.0 ppm
Chem concentration,%: 100
Flux decline % per year: 7.0 0.0
Feed water type: Seawater - open intake
SP increase % per year: 10.0 0.0
Permeate blending ☐
Permeate throttling ☐
Product recovery, %: 50.0 85.0
Concentrate recirc. ☐
Booster pump ☐
Permeate flow: m3/hr 800.00 350.00 738.24
Split Partial ☑
Average flux rate: l/m2-hr 13.5 32.7
Feed flow: m3/hr 1600.0 411.8
Concentrate flow: m3/hr 800.0 61.8
Passes: 2
Interstage pump ☒

System Specs
Element type: SWC4+
Elements/vessel: 8
Vessels: 200
Stages Pass 1: 1

Element type: ESPA2 ESPA2
Elements/vessel: 8 8
Vessels: 24 12
Stages Pass 2: 2

☐ AutoDisplay
☑ Recirculation2

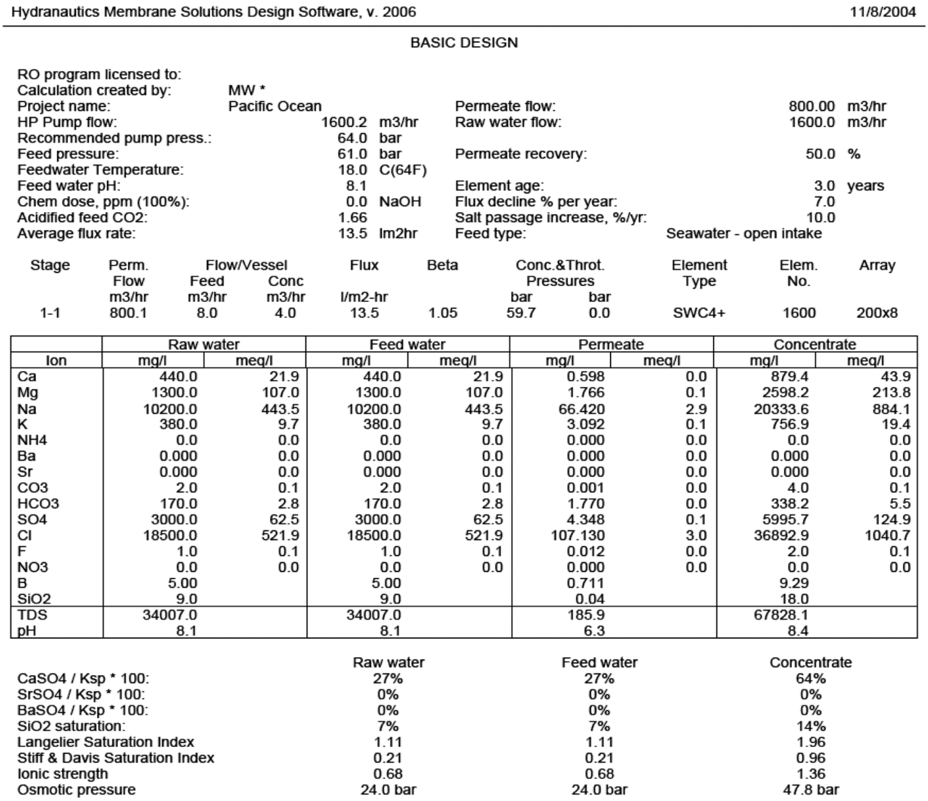
FIG. 6.5 Computer projection program—two pass unit design parameters entry screen.



		Main
Feed pressure	bar	63.9
Concentrate pressure	bar	62.5
Permeate flow	m3/hr	800.0
Pump Feed Flow		1600.0
Recovery ratio, %		50.0
Pump efficiency, %		83.0
Motor efficiency, %		93.0
ERT efficiency, %		87.0
ERT backpressure	bar	0.0
Power/Stage/Pass		3607.3
Pumping energy	kwhr/m3	3.04
Pumping power	kw	3614.6
Recovered power	kw	1184.7
Power requirement	kw	2429.9

Default values ☐

FIG. 6.6 Computer projection program—calculation of power requirement screen.



```
Hydranautics Membrane Solutions Design Software, v. 2006                                   11/8/2004
                                         BASIC DESIGN

RO program licensed to:
Calculation created by:       MW *
Project name:                 Pacific Ocean               Permeate flow:                       800.00  m3/hr
HP Pump flow:                               1600.2  m3/hr  Raw water flow:                      1600.0  m3/hr
Recommended pump press.:                      64.0  bar
Feed pressure:                                61.0  bar    Permeate recovery:                     50.0  %
Feedwater Temperature:                        18.0  C(64F)
Feed water pH:                                 8.1         Element age:                            3.0  years
Chem dose, ppm (100%):                         0.0  NaOH   Flux decline % per year:                7.0
Acidified feed CO2:                           1.66         Salt passage increase, %/yr:           10.0
Average flux rate:                            13.5  lm2hr  Feed type:                 Seawater - open intake

  Stage    Perm.      Flow/Vessel       Flux      Beta     Conc.&Throt.      Element    Elem.    Array
           Flow     Feed      Conc                          Pressures        Type       No.
           m3/hr    m3/hr     m3/hr    l/m2-hr              bar      bar
   1-1     800.1     8.0       4.0      13.5      1.05      59.7     0.0      SWC4+      1600     200x8

|         |      Raw water      |     Feed water      |      Permeate       |     Concentrate     |
|   Ion   |   mg/l   |  meq/l   |   mg/l   |  meq/l   |   mg/l   |  meq/l   |   mg/l   |  meq/l   |
| Ca      |    440.0      21.9  |    440.0      21.9  |    0.598       0.0  |    879.4      43.9  |
| Mg      |   1300.0     107.0  |   1300.0     107.0  |    1.766       0.1  |   2598.2     213.8  |
| Na      |  10200.0     443.5  |  10200.0     443.5  |   66.420       2.9  |  20333.6     884.1  |
| K       |    380.0       9.7  |    380.0       9.7  |    3.092       0.1  |    756.9      19.4  |
| NH4     |      0.0       0.0  |      0.0       0.0  |    0.000       0.0  |      0.0       0.0  |
| Ba      |    0.000       0.0  |    0.000       0.0  |    0.000       0.0  |    0.000       0.0  |
| Sr      |    0.000       0.0  |    0.000       0.0  |    0.000       0.0  |    0.000       0.0  |
| CO3     |      2.0       0.1  |      2.0       0.1  |    0.001       0.0  |      4.0       0.1  |
| HCO3    |    170.0       2.8  |    170.0       2.8  |    1.770       0.0  |    338.2       5.5  |
| SO4     |   3000.0      62.5  |   3000.0      62.5  |    4.348       0.1  |   5995.7     124.9  |
| Cl      |  18500.0     521.9  |  18500.0     521.9  |  107.130       3.0  |  36892.9    1040.7  |
| F       |      1.0       0.1  |      1.0       0.1  |    0.012       0.0  |      2.0       0.1  |
| NO3     |      0.0       0.0  |      0.0       0.0  |    0.000       0.0  |      0.0       0.0  |
| B       |     5.00            |     5.00            |    0.711            |     9.29            |
| SiO2    |      9.0            |      9.0            |     0.04            |     18.0            |
| TDS     |  34007.0            |  34007.0            |    185.9            |  67828.1            |
| pH      |      8.1            |      8.1            |      6.3            |      8.4            |

                                        Raw water             Feed water            Concentrate
CaSO4 / Ksp * 100:                         27%                   27%                    64%
SrSO4 / Ksp * 100:                          0%                    0%                     0%
BaSO4 / Ksp * 100:                          0%                    0%                     0%
SiO2 saturation:                            7%                    7%                    14%
Langelier Saturation Index                 1.11                  1.11                   1.96
Stiff & Davis Saturation Index             0.21                  0.21                   0.96
Ionic strength                             0.68                  0.68                   1.36
Osmotic pressure                         24.0 bar              24.0 bar               47.8 bar
```

FIG. 6.7 Computer projection program—example of computer printout for seawater unit design.

The computer projection program produces summary of input and calculated parameters in a form of a printout (Fig. 6.7), which serves as a basis for RO process design.

The important information on the printout form includes basic process design information as summarized in Table 6.1. The information is divided into user input and calculated results. The first group of calculated results are of primary importance and my affect the design process directly. The results in the second category of values are of secondary importance, providing only general information about the process.

Operating parameters for number of representative system configurations, calculated using computer projection program, are listed in Appendix A.

TABLE 6.1
Computer program information processing

User input parameters	Water analysis, feed water temperature and pH, permeate flow, recovery ratio, flux decline coefficient, salt passage increase coefficient, membrane age, membrane element type, array.
Primary calculated results	Feed flow, feed pressure, concentrate pressure, permeate composition, feed pH, osmotic pressure, concentrate composition, pH, osmotic pressure, saturation values of scaling constituents, average permeate flux, concentration polarization factor, feed and concentrate flow per vessel.
Secondary calculated results	Recommended pump pressure, acid dosage, pH of permeate and concentrate, permeate flux, recovery rate and permeate salinity for individual elements.

7

Normalization of system performance

Performance of RO/NF system is a result of aggregate performance of individual membrane elements. Each subsequent element in a pressure vessel, connected in series, operates at different values of feed salinity and feed pressure. Along the system feed salinity increases and feed pressure decreases. The performance and operating conditions are significantly different then the corresponding nominal values. In addition, system performances are affected by fluctuation of operating parameters: feed salinity, temperature, recovery rate and feed pressure.

In order to identify intrinsic changes of membrane performance such as permeability or salt passage, at the early stages of membrane deterioration process, system operational data are recorded at frequency at least once per day and normalized performance are calculated. The generic method of RO performance normalization is described in the ASTM procedure (42). In actual commercial RO applications the following normalization approaches are adopted:

1. Normalization to the reference (initial) operating conditions of the plant.
2. Normalization to the nominal element(s) test conditions
3. Calculation of water transport and salt transport values for the membrane elements in operation.

In the normalization calculations process each set of plant (or desalting stage) flows, pressures and salinities data is initially reduced to the average values. These average values are assumed to be representative for an element positioned somewhere in the middle of the system, on the feed–concentrate cross section line:, i.e., element that process an average feed salinity at an average applied feed pressure and produces average permeate flow. The averages are calculated based on feed–concentrate values. Then based on this data the water and salt permeability are calculated. In normalization approach 1 every set of the performance data of the system are being recalculated to the initial operating conditions: temperature, average feed salinity and NDP.

Any of the above performance normalization method will provide good presentation of membrane unit performance trend. Some advantage of the first method is that, in addition to normalized permeate flow and salt passage, it usually also provides trend of the pressure drop. Pressure drop is an important indicator of early stage of fouling, which results in blockage of the element feed channels. In normalization approach 2 the performance of the system are calculated and presented as a performance of an average element, it would perform, if tested at the nominal test conditions. Normalization approach 3 is very similar to the first one. In this calculations performance of RO system is reduced to performance of an average element. Then based on this data the water and salt permeability are calculated.

Normalization of salt passage and permeate flow is derived from the transport relations: Eqs. 10, 11 and 12. According to these equations the salt passage is function of salinity gradient and quantity of permeate available for dilution (permeate flux rate). Therefore, salt passage at a given operating conditions SP(1) is related to different operating conditions accordingly to the corresponding average permeate flux rate (APF):

$$\mathrm{SP}(2) = \mathrm{SP}(1)\ \mathrm{APF}(2)/\mathrm{APF}(1) \tag{22}$$

Permeate flow (Q_p) at condition 1 can be related to operating conditions 2 accordingly to corresponding net drawing pressures (ND) and temperature correction factors (TCF)

$$Q_p(2) = Q_p(1)\ \mathrm{NDP}(1)/\mathrm{NDP}(2)\ \mathrm{TCF}(1)/\mathrm{TCF}(2) \tag{23}$$

System information required and relations used for performance normalization calculations are summarized in Table 7.1.

Example 12

Normalization of system performance to the initial operating conditions (see Table 7.2)

Recovery rate – R_p

$R_p = Q_p/(Q_p + Q_r)$

TABLE 7.1
Summary of data and relations used in the normalization calculations

Membranes information	Total number of membrane elements, membrane area per element, nominal element flow and salt rejection.	
System operating data	Feed salinity or feed composition, feed temperature, permeate flow, concentrate flow, feed, concentrate and permeate pressure, feed, permeate and concentrate salinity	
Normalized calculation results	**Parameter**	**Relation**
	System recovery rate	Equation 4
	System average feed salinity	Equation 6
	System average osmotic pressure	Equation 2
	System NDP	Equation 8
	Temperature correction factor	Equation 16
	System average flux rate	Equation 17
	System salt passage	Equation 14
	Element nominal flux rate	Equation 17
	System normalized salt passage	Equation 21
	System normalized permeate flow	Equation 22
	System normalized pressure drop	Equation 52
	Element normalized salt passage	Equation 21
	Element normalized permeate flow	Equation 22
	Water transport coefficient	Equation 10
	Salt transport coefficient	Equation 12

$R(1) = 200/(200 + 50) = 0.80$
$R(2) = 180/(180 + 60) = 0.75$

Concentration factor – CF
$CF = \ln(1/(1 - \mathrm{R}))/\mathrm{R}$
$CF(1) = \ln\,(1/(1 - 0.80))/0.80 = 2.01$
$CF(2) = \ln\,(1/(1 - 0.75))/0.75 = 1.85$

Average feed salinity, ppm – C_{favg}
$C_{favg} = \mathrm{Cf\ CF}$
$C_{favg}(1) = 2000 \times 2.01 = 4020$
$C_{favg}(2) = 2500 \times 1.85 = 4625$

TABLE 7.2
System configuration and operating parameters

Brackish RO unit. Array 20:10. Number elements per vessel – 7

		Initial values of operating parameters, condition 1	Current values of operating parameters, condition 2
Feed salinity	ppm TDS	2000	2500
Permeate salinity	ppm TDS	30	50
Feed pressure,	bar (psi)	14 (203)	16 (232)
Concentrate pressure	bar (psi)	10.5 (152)	11 (160)
Feed–concentrate pressure drop	Bar (psi)	3.5 (51)	5 (72)
Permeate pressure	bar (psi)	1.5 (22)	1.5 (22)
Feed temperature	C	17	18
Permeate flow	m^3/h (gpm)	200 (880)	180 (792)
Concentrate flow	m^3/h (gpm)	50 (220)	60 (264)
Element type	8040 ESPA2	8040 ESPA2	
Number of elements	210	210	
Membrane area per element	m^2 (ft^2)	37 (400)	37 (400)

Average osmotic pressure, bar (psi)–Posmavg
Posmavg = 0.77 C_{favg}/1000
Posmavg(1) = 0.77 (4020/1000) = 3.1 (45)
Posmavg(2) = 0.77 (4625/1000) = 3.5 (51)

Average permeate flux, l/m^2/h (gfd)–APF
APF = Q_p 1000/(A EN)
APF(1) = 200 × 1000/(37 × 210) = 25.7 (15.1)
APF(2) = 180 × 1000/(37 × 210) = 23.2 (13.6)

Temperature correction factor–TCF
TCF = exp (2700 (1/(273 + t) – 1/298))
TC F(1) = exp (2700 (1/(273 + 22) – 1/298) = 1.284
TC F(1) = exp (2700 (1/(273 + 18) – 1/298) = 1.243

Net driving pressure bar (psi)–NDP
NDP = P_f – 0.5 (P_f – P_c) – P_p – Posmavg

NDP(1) = 14.0 – 0.5 (14.0 – 10.5) – 1.5 – 3.1 = 7.65 (111)
NDP(2) = 16.0 – 0.5 (16.0 – 11.0) – 1.5 – 3.5 = 8.5 (123)

Specific flux $l/m^2/h/bar$ (gfd/psi)–SF
SF = APF TCF/NDP
SF(1) = 25.7 × 1.284/7.65 = 4.31 $l/m^2/h/bar$ (0.172)
SF(2) = 23.2 × 1.243/8.50 = 3.39 $l/m^2/h/bar$ (0.136)

Salt passage,% – SP
$SP = 100\ C_p/C_{favg}$
SP(1) = 100 × 30/4020 = 0.74
SP(2) = 100 × 50/4625 = 1.08

Normalized salt passage,% – NSP
NSP(2) = SP(2)APF(2)/APF(1)
NSP(2) = 1.08 × 23.2/25.7 = 0.97%

Average feed flow m^3/h (gpm)
$Q_{fav} = (Q_f + Q_c)/2$
$Q_{fav}(1) = (200 + 50)/2 = 125$ (550)
$Q_{fav}(2) = (180 + 60)/2 = 120$ (528)

Normalized pressure drop, bar (psi)
NPD(2)= DP1 (Q_{fav1}/Q_{fav2})^1.4
NPD(2) = 5 (125/120)^1.4 = 5.29 (77)

TABLE 7.3
Summary operating parameters normalized to initial performances

Normalized value	Initial values	Current values	Difference,%
Specific flux, $l/m^2/h/bar$ (gfd/psi)	4.31 (0.172)	3.39 (0.136)	–21
Salt passage, %	0.74	0.97	+31
Pressure drop, bar (psi)	3.5 (51)	5.29 (77)	+51

Example 13

Normalization of system performance (Table 9, conditions 2) according to the nominal test conditions

Element type:	8040ESPA2
Membrane area:	37 m^2 (400 ft^2)

Nominal permeate flow:	34 m^3/d (9000 gpd)
Nominal salt rejection:	99.5%
Test pressure:	10.3 bar (150 psi)
Pressure drop	0.3 bar (4 psi)
Feed salinity	1500 ppm NaCl
Recovery rate	15%
Temperature	25°C

Permeate flux at nominal test conditions, l/m^2/h (gfd)
$PFN = 34 \times 1000/(37 \times 24) = 38.2\ (22.5)$

Concentration factor at nominal test conditions
$CFN = \ln (1/(1\text{-}0.15))/0.15 = 1.083$

Average feed salinity at nominal test conditions
$C_{favg}N = 1500 \times 1.083 = 1624$ ppm

Osmotic pressure at nominal test conditions, bar (psi)
$P_{osm}N = 0.77 \times 1624/1000 = 1.2\ (17)$

Net driving pressure at nominal test conditions, bar (psi)
$DDPN = 10.3 - 0.5 \times 0.3 - 1.2 = 8.9\ (129)$

Average element permeate flux in the system (Table 9), l/m^2/h (gfd)
$Fluxavg(1) = 200 \times 1000 /(210 \times 37) = 25.74\ (15.1)$
$Fluxavg(2) = 180 \times 1000 /(210 \times 37) = 23.16\ (13.6)$

Average element permeate flow in the system (Table 9), m^3/d (gpd)
$Q_{Pavg}(1) = 200 \times 24/210 = 22.85\ (6036)$
$Q_{Pavg}(2) = 180 \times 24/210 = 20.57\ (5434)$

Average element permeate flow normalized to nominal test conditions (NDP and TCF from Example 12), m^3/d (gpd)
$Q_{Pavg}N(1) = 22.85 \times 1.284 \times 8.9/7.65 = 34.13\ (9018)$
$Q_{Pavg}N(2) = 20.57 \times 1.243 \times 8.9/8.50 = 26.77\ (7072)$
Permeate flow change between conditions 1 and 2: –22%

Average element salt passage normalized to nominal test conditions (SP(2)) from Example 12
$SPN(1) = 0.74\ (25.74/38.2) = 0.50$
$SPN(2) = 1.08\ (23.16/38.2) = 0.65$
Salt passage change: +30%

Average element salt rejection normalized to nominal test conditions, %
Rej (1) = 100 – 0.50 = 99.50
Rej (2) = 100 – 0.65 = 99.35

Example 14

Calculation of water transport and salt transport values for the membrane elements in operation.

Calculations of water transport value (A) are conducted according to Eq. 10 using values calculated in Example 13.

$$Q_w = A\ S\ \mathrm{NDP} \tag{10}$$

$$A = \mathrm{Flux}_{\mathrm{avg}}/\mathrm{NDP} \tag{10a}$$

The units of A are s^{-1}

A(1) = 25.70 (1000/ 10000 × 3600)/(7.65 × 1000) = 9.3E – 8
A(2) = 23.16 (1000/ 10000 × 3600)/(8.50 × 1000) = 7.6E – 8
Difference of water transport coefficients –19%

Calculations of salt transport value (B) are conducted according to Eq. 12 using values calculated in Example 13.

$$Q_s = B\ S\ \Delta C \tag{13}$$

$$B = \mathrm{Flux}_{\mathrm{avg}}\ (C_p/\Delta C) \tag{13a}$$

The units of B are $\mathrm{g/cm^2\text{-}s}$

At low permeate salinity one can assume that DC equals average feed salinity: C_{favg}
B(1) = 25.70 × 30 (1000/(10000 × 3600))/4020 = 5.32E – 6
B(2) = 23.16 × 50(1000/(10000 × 3600))/4625 = 6.95E – 6
Difference of salt transport coefficients +30%

The results listed in Examples 12–14 illustrate three different methods of performance normalization calculation. All calculations are based on the same principles that permeate flow depends on net driving pressure and temperature, and salt passage is function of salinity gradient. The calculation accuracy can be improved by incorporating the effect of temperature on salt transport and by more rigorous calculation of osmotic pressure.

The above methods of normalization are accurate enough for normalization of performance of brackish membrane elements. In normalization of performance of low rejection nanofiltration elements, salinity of permeate should be included in calculations of NDP and salinity gradient. In system operating with high salinity feed (seawater), concentration polarization has significant effect on permeate flow. This effect should be account for, especially in calculations of normalized permeability. Otherwise, comparison of performance results obtained at different feed salinities will not provide meaningful results.

8

Feed water supply system and pretreatment

8.1. Well water

The configuration of the feed water supply system depends on the type of water source. The common sources of feed water for RO/NF plants are brackish wells, seawater intakes and treated municipal wastewater effluents. Less common sources are brackish surface waters, seawater beach wells or industrial wastewater effluents. Brackish wells providing water to an RO/NF system could be deep aquifer wells, reaching a depth of hundreds of meters, or shallow wells collecting surface water that infiltrated to shallow strata.

The composition of water from deep wells is in most cases very stable and of good quality. The fluctuations of water salinity and temperature are low. This is under conditions that the pumping rate is consistent with the permeability of the formation surrounding the well (production zone), and within the rate of aquifer replenishment.

The allowed pumping rate is determined through hydrologic evaluation of the water producing aquifer. This includes drilling pilot wells, test pumping and determination of step-drawdown for a range of water drawdown rates. In some aquifers, which are under influence of water of different salinities, the salinity could change with time as a result of pumping. In the majority of such cases the change of water quality is gradual. However, these changes should be considered at the system design stage. Shallow wells are more prone to quality fluctuations due to infiltration of seasonal surface run-off.

For RO applications it is important that the water supply well is constructed from materials that are compatible with a saline environment. The major concern is corrosion, therefore the well casing, pump and piping should be made of a corrosion resistant alloy or polymeric materials. However, in the case of plastic materials, especially if epoxy based FRP components are used, it is important that the resin components of FRP material are completely cured so that they do not release even minute concentrations of organic compounds that could otherwise reduce membrane permeability.

Water originating from a properly designed well field has very low concentration of suspended solids. Therefore, in RO/NF systems treating well water, operating in the US, it is common practice not to include media filtration in the pretreatment process and use only cartridge filters on the feed line (Fig. 8.1). Cartridge filters for RO applications have a nominal porosity in the range of 5–15μ. The preferred porosity rating of filtration cartridges is 5μm. Feed water flow through cartridge filters should not exceed 1 m^3/h (4.4 gpm) per 25 cm (10″) cartridge length. The schematic configuration of the cartridge filter housing is shown in Fig. 8.2. In this configuration the filtration cartridges are mounted in a vertical position. In large capacity plants horizontal cartridge filters (Fig. 8.3) are sometimes used. The important part of the cartridge filter housing is the baffle, which protects cartridges from direct impingement of suspended particles. Filtration cartridges are usually made of soft polymeric materials. Exposure to direct impingement of hard particles could result in abrasion of cartridges and eventually loss of integrity.

The role of cartridge filters is mainly to protect equipment located downstream (pumps and membrane elements) from the sudden appearance of particulate matter in the feed water. Such conditions could sometimes be experienced due to sudden sand or silt release from wells or from sand filters (in a system utilizing sand filters). Use of cartridge filters as a feed water filtration step for colloidal matter removal is usually prohibitively expensive in respect of the cartridge replacement cost.

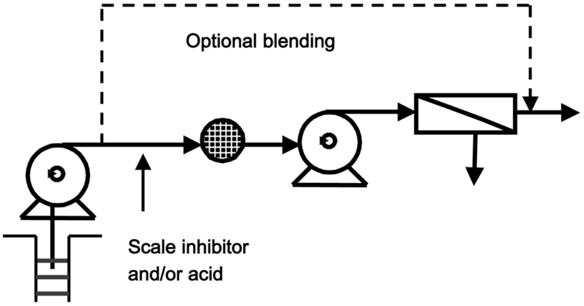
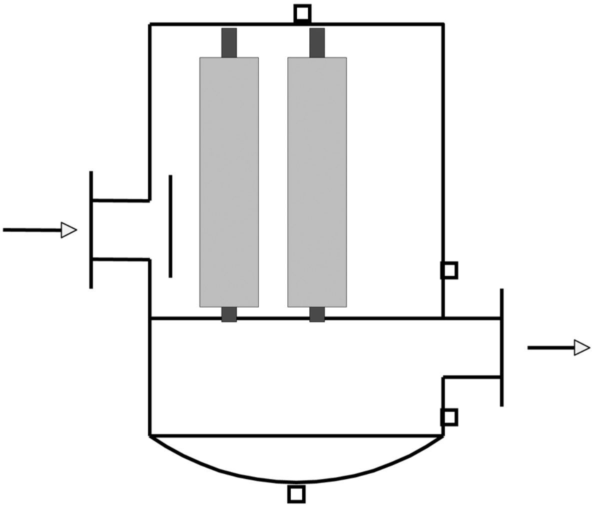


FIG. 8.1 Configuration of an RO unit (brackish or seawater) operating on a well water feed.



Max flow, m^3/h	Number of 25 cm cartridges	Diameter, cm	Height, cm	In-out diameter, cm	Weight, kg
3.5	3	33	70	5.0	57
7.0	6	38	70	5.0	82
27.0	24	38	130	7.5	100
86.0	76	60	150	10.0	200

FIG. 8.2 Configuration of cartridge filters.

FIG. 8.3 Battery of cartridge filters housings in horizontal configuration.

Extensive field experience shows that RO systems treating well water, with cartridge filtration as the only filtration step, operated successfully over the years. In some isolated cases, plants with this configuration experienced release of silt and/or sand from wells, preventing the cartridge filters from operating properly; the particulate matter eventually ended up reaching membrane elements. In almost all reported cases of such events, the solids intrusion and accumulation was limited to the lead elements only. This condition was eventually rectified by flushing of the lead elements (in the reverse flow direction, after turning them around and moving them to the end of the system) with some replacements.

One big advantage of a pretreatment filtration configuration limited to cartridge filters only, is the reduction of exposure of feed water to the outside environment, which is very convenient in the treatment of anaerobic water sources. Anaerobic water sources (for example water from a deep Floridian aquifer) may contain variable quantities of hydrogen sulfide and usually sulfate reducing bacteria are present as well. Had this water been exposed to air there would be a high probability of hydrogen sulfide being partially oxidized to elemental sulfur according to the following reaction:

$$2H_2S + O_2 = 2H_2O + 2S \tag{24}$$

Elemental sulfur has very limited solubility in water or water based solutions and once deposited in the feed channel of RO elements, it cannot be removed. At the early stages of RO technology development attempts were made to oxidize hydrogen sulfide present in feed water with strong oxidants, prior to the RO. This process configuration almost always ended up in either fouling of the membrane elements with elemental sulfur or in oxidative damage of the membrane barrier. The design approach that provides stable system performance is to maintain anaerobic conditions of the water through the RO system. After the RO unit, hydrogen sulfide is removed from the permeate (and sometimes from the concentrate as well) either by aeration or oxidation. If the location of the RO system is close to an urban center the degasifiers can not vent off gasses to the air. The common solution in case of hydrogen sulfide aeration is to follow the degasifiers with an absorption system. In such a system hydrogen sulfide is absorbed on an iron based catalyst and eventually disposed as a solid waste.

For any feed water source, but especially for an anaerobic one, it is important to design a system that assures complete total exclusion of light from the feed water. Even slightly translucent materials used as a part of piping or storage tanks can transmit enough light to provide sufficient energy for bacteria to grow. The problem of bio-growth due to some light transmission is quite common for unpainted FRP piping or water storage tanks made of plastic materials. If the feed water supply system consists of a number of wells, used as a combined source of feed water, it is important to evaluate compatibility of their mixture in respect of potential solids precipitation. According to what was described above about the nature of anaerobic water, water from an anaerobic source cannot be mixed together with water containing dissolved air due to presence of oxygen and possibility of hydrogen sulfide oxidation.

Seawater beach wells, sometimes used as a feed water source for seawater RO systems, are usually quite shallow. They can be built as a regular wells or Ranney wells or as a combination of both configurations (Fig. 8.4).

As is the case of brackish wells, seawater beach wells provide water with a low concentration of suspended solids. One of the major limitations of seawater wells is their limited output capacity, usually in the range of few thousand m^3/day (few MGD). Because of the low recovery rate of seawater systems, e.g., 35–50%, beach wells can only support RO systems of a limited permeate capacity. Another problem with beach wells is in obtaining permits. The general public is quite sensitive about building any structure in the seashore area. At present, obtaining approvals for construction of a large number of beach wells necessary to support a large capacity desalination plant, can be a very difficult task.

As indicated in Fig. 8.1 the pretreatment for a well water based system is usually limited to pH adjustment and/or addition of a scale inhibitor together with cartridge filtration. For some feed water supply wells, which have a history

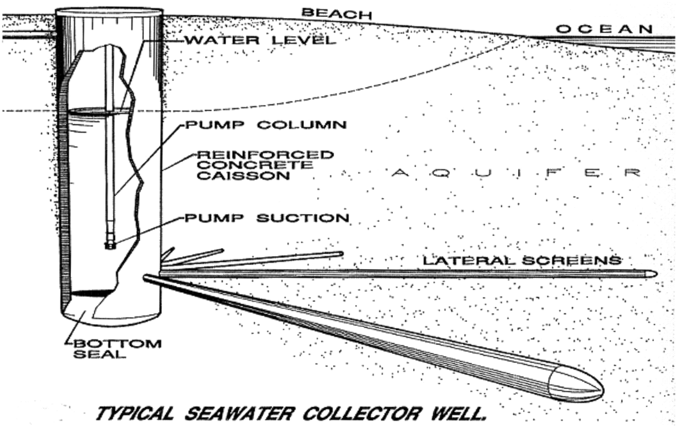


FIG. 8.4 Configuration of beach well (Courtesy of Collector Wells International).

of releasing sand, in line hydro-cyclones or sand screens are used. Water produced by shallow wells sometimes contains significant amount of suspended particles. The process designer may decide, based on pilot operation results, to use media filtration upstream of the cartridge filters. Inclusion of media filtration in treatment of questionable water quality could reduce the rate of RO membrane fouling and decrease the frequency (and cost) of cartridge filters replacement.

8.2. Surface water

8.2.1. Raw water intake and concentrate discharge

Brackish RO systems treating surface water are not very common. Large seawater RO systems, on the other hand, are based almost exclusively on surface water supply. The surface water is provided from an intake structure, which could be located off shore, or pumped from a lagoon or intake canal. The intake structure could be in the form of intake towers, shore intakes, siphon well intakes, floating intakes or submersible intakes. This last configuration is the prevailing one in large seawater RO plants. If possible, the submersible, off shore intake structure should be positioned at location of the most stable water quality. The location for the seawater intake is selected to ensure a 10–15 m (32–49 ft) water depth, even during low tide periods. Depending on the sea floor topography, it may require the intake inlet structure to be located hundreds of meters off shore. The submersible intake structure is usually designed and built to provide a water inlet at a level of 2–5 m (6–16 ft) above the sea floor to reduce the possibility of clogging by silt and sand (Fig. 8.5). The level of the inlet should be selected to consistently provide a good water quality, possibly free of algae and plankton. The screen size and a slow inlet flow rate (<0.3 m/sec, <1ft/sec) should ensure low entrapment and impingement of water born organisms. The water flow in the intake conduit should be at much higher rate (1–1.5 m/s, 3–5 ft/s) to prevent deposition of sediments in the connecting piping. Very comprehensive descriptions of various intake configurations can be found in the book: “Water treatment plant design” (38), edited by AWWA.

In seawater plants about half of the feed flow is returned back to the ocean as a concentrate at a significantly higher concentration. The concentrate discharge structure should be positioned at sufficient distance from the intake to ensure that the water concentration at the intake is not affected by the concentrate salinity (Fig. 8.6). The example of an excess salinity decay profile during concentrate discharge is shown in Fig. 8.7 and discussed in reference 43.

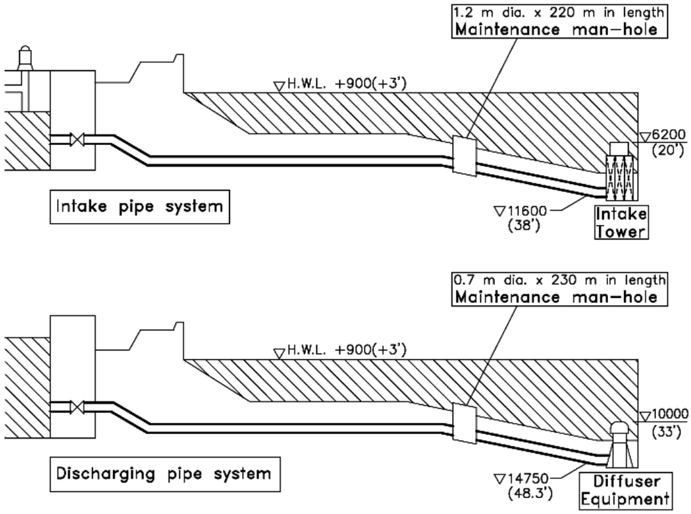


FIG. 8.5 Configuration of seawater intake and concentrate discharge. Adapted from ref. 43.

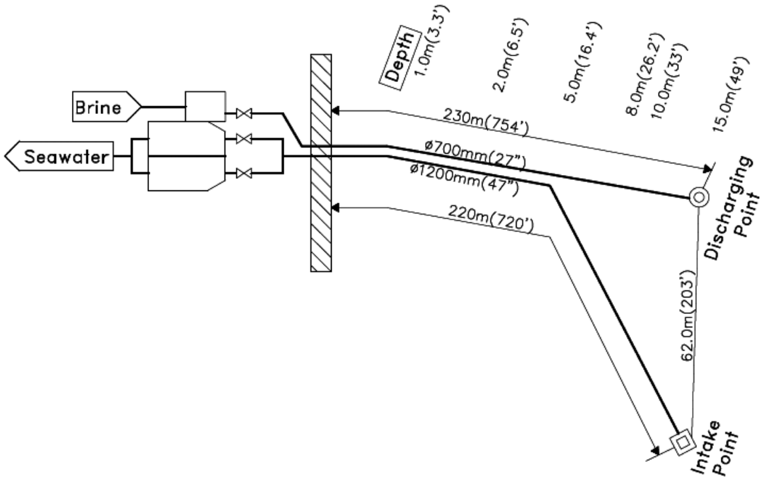


FIG. 8.6 Relative positioning of intake and concentrate discharge structures. Adapted from ref. 43.

Recently, some developers of large seawater projects have considered locating RO plants contiguous to some other larger seawater users. Convenient candidates are electric power generation plants with seawater cooled steam condensers. A schematic diagram of water flow in an RO system located adjacent to a power

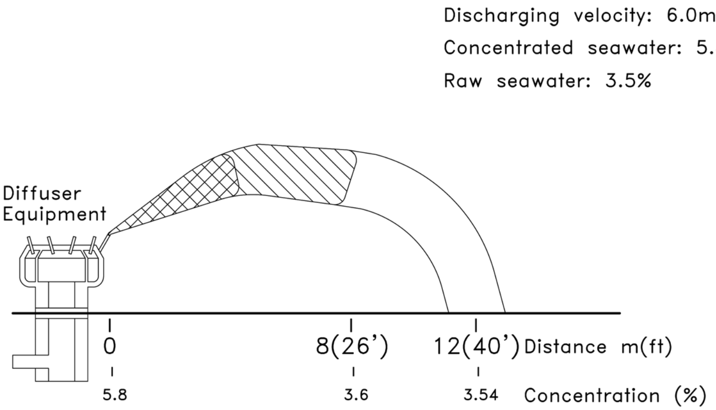


FIG. 8.7 Example of salinity profile vs. distance from concentrate discharge structure. Adapted from ref. 43.

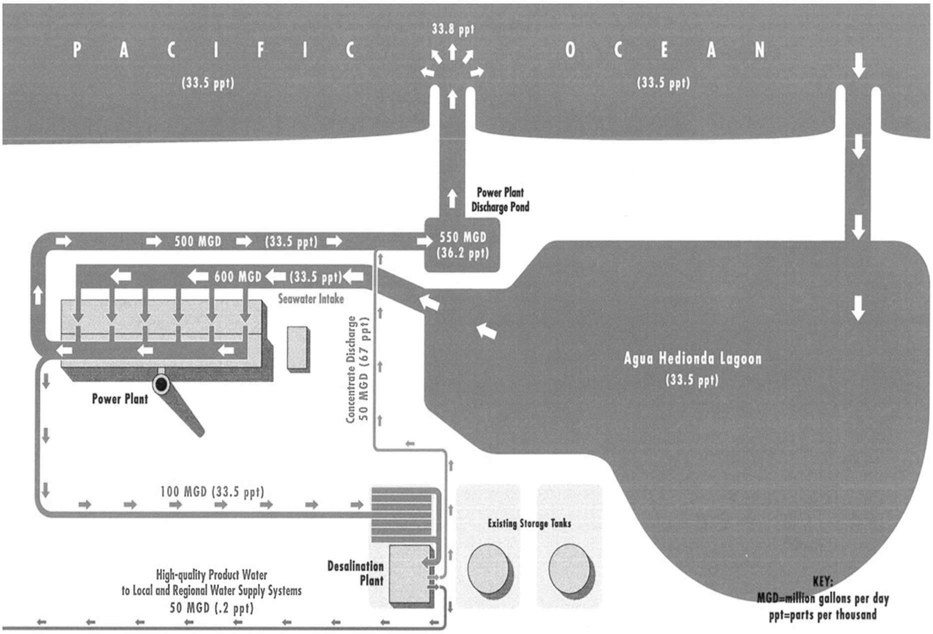


FIG. 8.8 Configuration of feed water supply and concentrate discharge in RO plant co-located with a power plant (Courtesy of Poseidon Corporation).

plant is shown in Fig. 8.8. There are number of advantages and some potential problems with this option of a feed water supply configuration. Supply from a dedicated intake usually implies a dedicated outfall facility as well. Current regulations require careful design that will minimize any potential environmental effect (43). The lengthy permitting process for the construction and operation of intake and outfall facilities makes location of an RO plant adjacent to an on-shore located power plant a very convenient solution. In this process configuration the RO system utilizes the seawater discharged from the heat reject section of the power plant as a feed before it flows to the ocean. In a similar fashion, the RO concentrate is discharged to the same line, downstream of the feed uptake.

The temperature of seawater at the outlet from the power plant is usually higher by 3–10°C (5–18°F) than the water temperature at the intake. For locations with a low seawater temperature (<20°C, <68°F) the temperature increase due to operation of the power plant is beneficial since it increases membrane permeability. This allows the RO system to operate at lower feed pressure. However, at locations with high water temperatures, e.g., above 30°C (86°F), a further increase of the feed water temperature does not result in any significant decrease of feed pressure (Fig. 2.6). Depending on the feed salinity and recovery rate in the temperature range of 30–40°C (86–104°F), the effect of the higher rate of water permeability through the membrane at a higher temperature, is adversely compensated by increased osmotic pressure.

Higher feed water temperature also results in higher salt passage (shown in Fig. 2.6). If this increase of salt passage requires increased operation of the second pass, higher feed water temperature can actually result in higher power consumption of the RO plant. Location of an RO system, contiguous with a power plant may result in some feed water quality problems. It is common practice for a power plant to intermittently chlorinate the intake structure to reduce biogrowth. An additional periodic event, that may affect seawater quality, is cleaning of the heat transfer surfaces of the condenser. As a result of cleaning, small particle fragments are released to the cooling seawater and could end up in the RO feed. Both of these periodic events at power plants, i.e., intake chlorination and cleaning of the condenser heat exchange surfaces, should be addressed in the RO system design and operation to prevent potential membrane damage.

8.2.2. Conventional filtration pretreatment

Concentration of suspended solids in surface water sources is usually much higher then in well water and could fluctuate over a wide range. The fluctuations are due to seasonal weather patterns (rainy seasons) or variability of biological activity (algae blooming periods). Therefore, surface water pretreatment systems have to contain additional solid removal steps (Fig. 8.9), as compared to RO systems treating well water (Fig. 8.1). The treatment steps of surface seawater usually include coagulation, flocculation and media filtration. Sometimes sedimentation

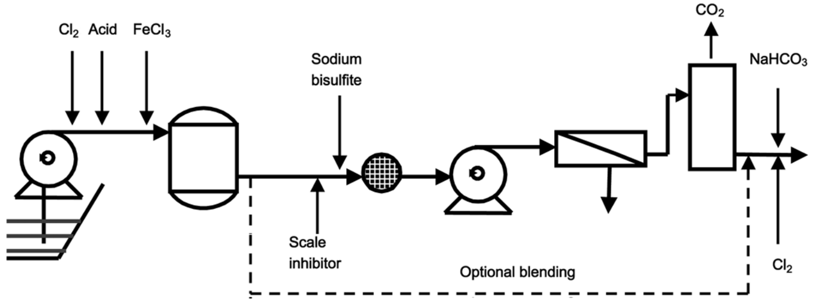


FIG. 8.9 Configuration of brackish RO unit operating on surface water feed.

and additional screening is included as an initial treatment step. As shown in Fig. 8.10 the feed water pretreattment system may include a number of treatment steps to improve water quality.

If disinfection is required, it is recommended to apply it intermittently to prevent bacterial after-growth and biofouling downstream. Disinfection with oxidants (free chlorine) requires subsequent injection of a reducing agent at some point before RO membranes to prevent membrane damage. The reducing agent most frequently used in RO systems, is sodium bisulfite. The stochiometric requirement is about 1.8 ppm of sodium bisulfite for 1 ppm of free chlorine. However, the usual dosing rate of sodium bisulfite in RO pretreatment systems is three times the concentration of free chlorine in the feed water. The reduction

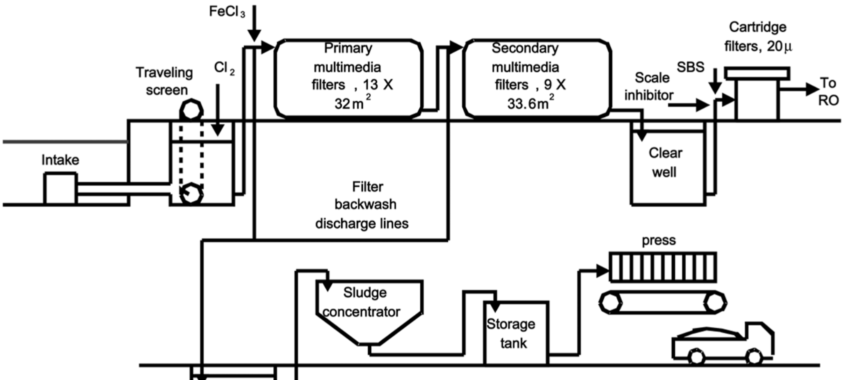


FIG. 8.10 Configuration of two stage media filtration system in seawater RO plant. Adapted from ref. 43.

reaction between bisulfite and free chlorine is very fast and completed in matter of seconds.

After removing large size floating debris with traveling screens and settling, water is pumped to the next treatment step, which is usually in line coagulation and flocculation followed by direct filtration. In the so called direct filtration process, water is treated, as it flows through the delivery piping, with conditioning chemicals (coagulants) and filtered, without prior clarification, by media filters.

Coagulation and flocculation is a combined process of destabilization and conglomeration of colloidal particles to facilitate more effective removal in media filtration process. Colloidal particles in the water stream are negatively charged and electrostatic repulsion helps to maintain them in suspension. Coagulants are positively charged hydrolyzed metal salts that neutralize negative charges of suspended colloids and help to aggregate them into larger, heavier, more filterable solids. For coagulation ferric or alum salts are mainly used. During hydrolysis of these salts a complex polynuclear, positively charged species are formed in a matter of seconds. The solubility of these species is low and they form a dense, suspended flock. The action of coagulants is threefold: they adsorb colloidal particles on the flock surface, neutralize negative charges that surround colloidal particles and also enmesh suspended particles in the body of the flock formed. The effective quantity of coagulant required is specific to water composition, type of colloidal particles, water pH and temperature.

An excessive quantity of coagulant could have the undesirable effect of increasing the stability of colloidal particles. In addition, a high concentration of coagulant may increase dispersion of colloids, due to reversal of surface charges: formation of high density, positive charges on the colloids surface and mutual electrostatic repulsion. The initial estimation of the required dosing rate of the coagulant and the optimum pH range for the process is determined by conducting a “jar test” (described in chapter 12). The dosing rate of the inorganic coagulant is usually in the range of 1–30 ppm with pH in the range of 6–8. Following the results of the jar test the adjustment of the coagulation process parameters is conducted during the initial stages of commercial system operation.

For RO applications ferric salts are preferred over aluminum due to the lower tendency of forming deposits in the membrane elements. The solubility of hydrolyzed species of aluminum and iron depends on pH, which is at a minimum at about pH 6 for aluminum hydroxide and pH 8 for ferric hydroxide. The solubility of hydrolyzed ferric compounds is much lower than of corresponding aluminum species. Therefore, if coagulation is applied, the pretreatment system effluent, and subsequently RO feed water will have a lower concentration of ferric ions than the potential concentration of aluminum ions at corresponding conditions. Accordingly, in case of ferric coagulant there will be a lower potential for precipitation as a result of feed water pH changes and/or due to the increase in concentration of dissolved species that occurs in the RO process.

As mentioned already coagulation is a very rapid process requiring just a few second to complete. However, effective coagulation requires intensive mixing to bring the coagulant in contact with a large number of colloidal particles. In RO applications this is usually achieved by incorporating static mixers positioned downstream of the coagulant injection point. Conversion of metal coagulants to the hydrolyzed form consumes alkalinity in the water, therefore the raw water pH is reduced (0.1–0.3 pH units), in proportion to coagulant dosing and alkalinity present, according to the following equations:

$$FeCl_3 + 3HCO_3^- = Fe(OH)_3 + 3Cl^- + 3CO_2 \tag{25}$$

$$AlCl_3 + 3HCO_3^- = Al(OH)_3 + 3Cl^- + 3CO_2 \tag{26}$$

Coagulation can be also conducted using long chain synthetic organic polymers, which could be of nonionic, anionic or cationic types. The nonionic and anionic polymers destabilize colloids by bridging particles together. The cationic type polymers have a dual action of bridging and neutralization the negative surface charges of the colloids. Cationic organic polymers can be used as primary coagulants. However, in RO pretreatment systems polymers are usually used as additives to enhance the effectiveness of metal based coagulants by binding flock particles together. In most cases, polymers are applied at low dosing rage, below 1 ppm, directly injected to the feed water downstream of the dosing point of the metal coagulant, at the location where hydrolyzed metal flock has been already formed. If polymers are used at a high dosing rate, and a carry over from the sand filters occurs, cationic polymers may react with anionic scale inhibitors and form a fouling layer on the membrane surface.

Flocculation, which follows coagulation, is a process of flock formation during gentle mixing. Flocculation is a slower process than coagulation and takes a number of minutes to complete. During flocculation, colloidal particles and some fraction of the dissolved organics are attached to the flock body, and are eventually retained on the filtration layer in granular media filters. In the granular media filtration process, suspended solids are removed through attachment to the filtration media particles and through blockage/capture by the filtration cake. The preferred process of filtration is capture of suspended solids with significant bed penetration as opposed to surface filtration, since the latter results in a faster increase of pressure loss and therefore shorter filter runs. In a single medium filtration bed, after number of backwash runs, fine size filtration media particles are aggregated at the top of the bed. This reduces penetration of suspended solids and therefore, mainly results in surface bed filtration. A graduation of the filtration bed from coarse to fine particles can be achieved in a dual media configuration by placing fine, high specific gravity, filtration media as the lower filtration layer and coarse, low specific gravity, filtration media as a top layer. Filtration media selection that provides a coarse to fine filtration bed configuration, includes anthracite (specific density 1.5–1.75 t/m^3 (93–110 lb/ft^3), effective size around 1 mm. as a top layer and silica sand (specific density: 2.55–2.65 t/m^3 (159–165 lb/ft^3), effective size around 0.5 mm, as a bottom filtration layer. Effective size means that size of 90% of filter media particles in the given lot is larger than the value indicated.

There is a variety of media filtration equipment configurations used in potable and waste water filtration. In RO applications the frequently used filter types are pressure or gravity filters in single or two stage configurations. Pressure filters are cylindrical pressure vessels filled with a layer(s) of filtration media. They could be configured for horizontal or vertical operation (Fig. 8.11 and 8.12). The diameter of pressure filters is limited by the transportation constraints.

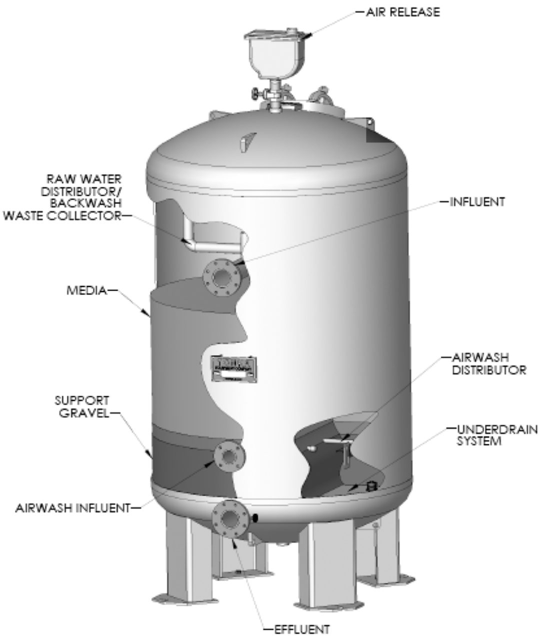


FIG. 8.11 Dual media, vertical pressure filter configuration (Courtesy of Tonka Equipment Company).

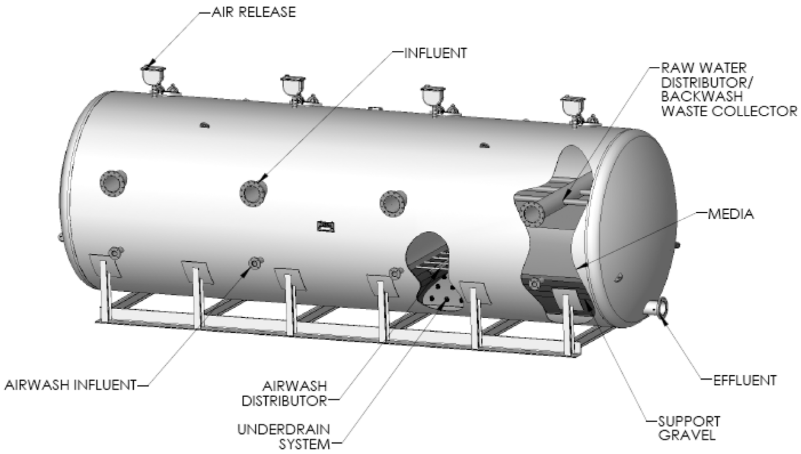


FIG. 8.12 Dual media, horizontal pressure filter configuration (Courtesy of Tonka Equipment Company).

Usually the diameter is limited to 3 m (10 ft). The above constraint limits the use of pressure filters mainly to smaller RO systems.

Some large RO seawater systems utilize pressure filters (Carboneras RO plant, Spain). However, they are of horizontal configuration, which enables larger filtration area per filter compared to the vertical type filters. When horizontal filters are used, the usual concern is to ensure good flow distribution and effectiveness of backwash of filter media. The common solution is to divide horizontal filters into compartments with independent feed and backwash.

Gravity filters have the configuration of rectangular tanks, usually made of concrete. An example of a gravity filter configuration is shown on Fig. 8.13. Influent water to the filter is supplied through a side channel. Filtrate leaves the filter through the effluent outlet located at the bottom of the filter. The same port serves as an entrance for the backwash water. Backwash operation is sometimes augmented by air. Backwash water is collected by the wash water troughs, located above the level that the media expands to during backwash. Some gravity filters utilize washing of the media surface using water jets. A typical dual media filter contains, in the direction of influent flow, 30–70 cm (12–28″) layer of anthracite followed by 20–50 cm (8–20″) of silica sand supported on a 10 cm (4″) layer of gravel.

Dual media filters usually operate at a filtration rate of 7.5–15 $m^3/m^2/h$ (3–6 gpm/ft^2). As filtration progresses, the water passage between particles of the filtration media increases as the interstices are progressively blocked by the deposition of colloidal particles and coagulant flock, and pressure drop (head

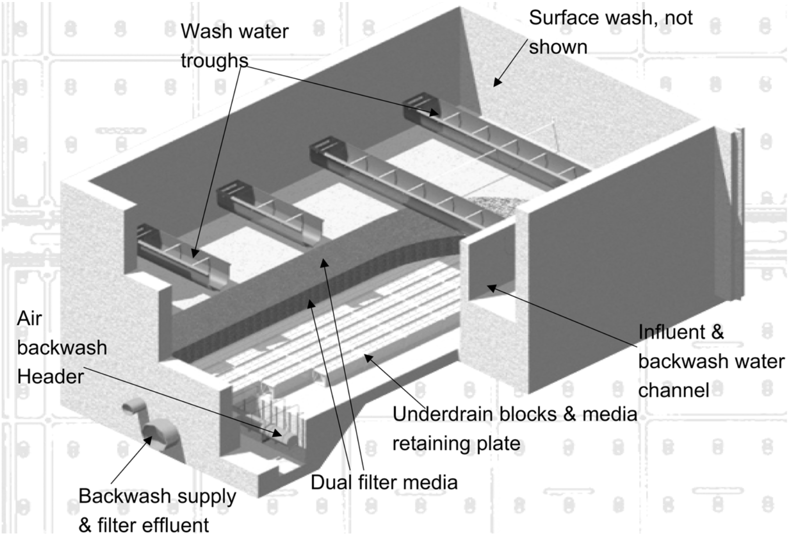


FIG. 8.13 Dual media, gravity filter configuration (Courtesy of Infilco-Degremont).

loss) across filtration media layer increases. Pressure drop has to be maintained below a terminal value, usually below 2.5 m (8 ft), otherwise channeling or media breakthrough can develop.

At the end of a filtration run the filter bed is cleaned from deposits in a backwash step and the initial value of pressure loss restored. The length of filtration run depends on effluent water quality. The filtration run can last from 8 hours to a number of days. During the backwash step water passes through the filtration bed in the reverse direction causing bed expansion and removal of deposits. Backwash operation is conducted at a flow rate of 35–55 $m^3/m^2/h$ (15–23 gpm/ft^2). Bed expansion during backwash is in the range of 20–50%. With increased water temperature, due to the lower buoyancy of the filtration media, the backwash flow rate has to be increased (about 2% per °C) to provide sufficient bed expansion. A special precaution should be applied for pressure filters to ensure that the backwash rate is not excessively high since it may result in removal of filtration media from the filter. In gravity filters a similar problem can be encountered. However, there the conditions of backwash and media level in the filter can be conveniently observed by plant operators.

Backwash is mainly conducted with filtered water. In seawater RO plants RO concentrate is sometimes used for filter backwash to reduce the filtrate requirement. For more effective bed expansion and the removal of deposits, backwash operation is sometimes combined with air scouring. Air mixed with backwash water passes through the filtration bed at rate of 55–90 m/h (3–5 scf/ft^2-min). Air scouring results in good bed expansion and effective mixing of the granular media in the filter. Therefore, for dual media filters, air scouring is followed by high rate water backwash to restore the original configuration of coarse to fine media segregation.

The typical backwash sequence includes: disconnecting the filter from the feed manifold, draining filter tank or vessel from water, backwash using an air and water mixture for 3–10 min followed by high rate backwash for 5–10 min, and reconnecting the filter to the feed manifold. Immediately after the backwash step, the filtrate quality is not as good as during normal operation, since it contains a high concentration of suspended particles. Filtrate quality gradually improves in a "ripening" process that could take up to 30 min. It is up to the plant operator to decide when to redirect filter effluent back to the membrane unit based on effluent quality (turbidity, SDI).

The filtrate volume required for a backwash operation is 2–3% of the system filtrate capacity per filtration stage. A backwash operation is usually conducted every 8–24 h of filter operation. It could be triggered by timer setting or pressure drop across filtration media.

At the majority of locations, the backwash effluent can not be discharged directly to any natural body of water, especially if the backwash water contains metal based coagulants. The current common procedure of backwash water disposal is on site treatment. It includes thickening, to increase solid concentration and dewatering in a filter press (see lower part of Fig. 8.10). Dewatered sludge is disposed as a solid waste and recovered water can be discharged or returned to the beginning of the pretreatment process.

For RO applications filtrate quality is determined through measurement of turbidity and SDI. There is no defined and consistent correlation between turbidity and SDI. However, field experience indicates, that to achieve an SDI below 3, the filtrate turbidity has to be below 0.1 NTU, preferably below 0.05 NTU. Except for relatively clean surface water sources, such low filtrate turbidity is difficult to achieve in a single stage filtration. Additional treatment steps may include clarification or diffused air flotation, prior to filtration.

A more frequent approach is to apply a two stage filtration treatment process to water with high and/or variable turbidity. It is common in a two stage filtration system that the first stage filters serve as roughing filters operating at high filtration rate. The second stage filters serve as a polishing filters operating at a lower filtration rate. However, filtration systems designed for the same filtration rate of both stages, or lower on the first and higher on the second stage, are also encountered. Coagulation and flocculation is usually conducted prior to the first filtration stage. Some two stage filtration systems have provision for an interstage addition of coagulant and/or pH adjustment. The effluent from the media filters is collected in a clear well and subsequently pumped through micron rated cartridge filters and from there to the suction of high pressure pumps. If required for scale prevention, acid or scale inhibitor is added to the feed water prior to cartridge filters. Extensive information on the water treatment processes applied in the conventional RO pretreatment system can be found in number of water treatment handbooks, among them references 45, 46, and 123.

8.2.3 Membrane filtration pretreatment

*Graeme Pearce**

Separation Mechanisms. The objective of pretreatment for an RO or NF system is to remove particles, reduce organics, and provide a feed that will not cause biofouling in the RO/NF elements. Conventional pretreatment technology relies on a combination of chemical treatment and media filtration to achieve conditioning of the feed to make it acceptable as an RO/NF feed.

The separation mechanism for UF and MF membranes differs from conventional treatment devices, such as granular or fibrous media filters. Media filters rely on a gravity removal mechanism. They have a nominal pore size considerably greater than the particles they are capturing. For a granular media filter, the grain size may be >100 micron, creating pores of a similar size. The absolute rating of such a filter will be of the same magnitude.

However, due to the depth of the media, and the tortuous path created for the feed as it moves through the media bed, relatively high removal rates can be achieved for particle sizes below the absolute rating. Media beds of 150 micron sand particles can routinely achieve removal efficiencies of 90-99% for particles down to 10–20 micron, and with coagulant addition, the removal rates are significantly better. The nominal rating of the depth filter means that the removal efficiency is variable, and is dependent on a host of environmental and operating parameters.

In contrast, UF and MF membranes operate by a surface removal mechanism, and resemble a fine screen or sieve. The pore size at the surface of a membrane is highly uniform, with a narrow pore size distribution. Particles larger than the size of the largest pore are rejected by the membrane surface,

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and remain on the feed or concentrate side. The bulk carrier fluid, and any particles finer than the largest pore, can pass through the membrane to the filtrate side.

UF can remove the finest particles commonly found in water supply, with the removal rating dependent upon the pore size of the active layer of the membrane; the typical removal capability of UF used for general water treatment is 0.01–0.02 micron. MF typically operates at a particle size an order of magnitude coarser than UF, e.g. approx 0.1–0.2 micron. UF therefore has the advantage over MF of providing a better disinfection barrier, since the pore size of UF will exclude viruses. However, in the application of RO/NF feed conditioning, both technologies are effective at reducing particle concentrations to a satisfactory level.

Although membranes provided a barrier to particles, colloids, and most microorganisms, dissolved components such as organics can still pass through the membrane. Organics may be a problem to the RO/NF, since they may cause fouling due to surface adsorption, or they may provide a food source to microorganisms. Sometimes therefore, chemical treatment with coagulants can still be beneficial prior to the UF/MF stage to adsorb organics, and be subsequently removed from the feed as fine particulates. If the concentration of dissolved organics is high, coagulant pretreatment would reduce both the fouling of the UF/MF stage, improve fluxes, and reduce cleaning frequencies of the RO/NF. However, coagulation is not needed in every case, and the need for coagulant should be established through pilot trails in the project development stage.

Advantages of membrane pretreatment. Since membranes provide a barrier to particulates, they provide significant benefits to the RO/NF, and to the overall system design. These benefits fall into two broad categories, namely reduced cost of the overall system, and improved on-stream time and security of supply.

The capital cost of membrane pretreatment normally exceeds that of conventional pretreatment by 20–50% depending on flow rate, feed quality, and local factors. However, the improved treated water quality can reduce the size of the RO by allowing a higher RO flux to be used. This advantage tends to be marginal for high salinity waters above 40,000 ppm TDS, since the RO flux at this salinity is limited by the osmotic pressure. However, for any feed below 30,000 ppm TDS, the RO flux could typically be increased by at least 25%, which would not only create a substantial saving for the RO system, but would mean that the pretreatment system would be smaller as well.

In addition to the savings due to the system size, UF/MF pretreatment has the advantage of requiring less space; typically, UF/MF provides a 33% saving in plant area, and this too can often translate into capital savings for the plant.

A further saving that results from the improved RO feed quality from UF/MF is that the replacement rate of the RO will be reduced, typically by 33%. Thus, if a 20% annual RO replacement rate is used for a conventional pretreatment system, the replacement rate following UF/MF may be reduced to 13%. Clearly, many factors influence replacement rate, both technical and commercial, but this gives an indication of the potential savings.

RO cleaning frequencies are typically reduced by UF/MF pretreatment, since SDI's are lower, and the RO feed will have very low particle concentrations. This results in considerable lengthening of the interval between cleans, which not only reduces the cost of cleaning chemicals, and the generation of chemical waste, but improves on-stream time for the RO system. In a typical whole life cost comparison, the savings in RO dosing and cleaning chemicals woll equate to the additional cost of UF/MF.

The UF/MF system itself may have higher running cost than conventional pretreatment. Although the chemical usage and disposal costs of conventional pretreatment will be significantly higher than UF/MF, the equipment and media of the conventional system are long lived, requiring low maintenance. For the UF/MF system, the membranes will need to be replaced from time to time. Membrane life is typically 5–10 years, so this requires a moderate running cost, but this should equate to the savings by using a lower RO replacement. Power cost for UF/MF is comparable to, or in some cases, slightly higher than conventional, but chemical usage in the pretreatment itself is lower.

Improved on-stream time for the RO with UF/MF pretreatment can be seen as both a cost issue and a security of supply issue. Not only does this benefit arise from reduced RO cleaning, but feed variability, occurring due to storms, can result in out of spec feed from conventional pretreatment. Operators then have the choice of shutting down, or operating outside RO warranty limits. It is not always appreciated that the effect of the excursion may be much longer lived than the excursion itself in terms of the effect on the RO membranes.

Cost study for UF/MF vs. conventional pretreatment. The factors listed above have been quantified in a study example, in which the costs are compared for UF/MF vs. conventional pretreatment for a nominal 50 mld feed to an RO system. The approximate overall cost for the complete RO system of this capacity with pretreatment, operating at 50% recovery, would be of the order of \$20M. Pretreatment would be approximately 15–20% of the overall cost, as indicated in the example cost breakdown for UF/MF and conventional pretreatment systems shown in Table 8.1.

TABLE 8.1
Capex cost comparison for UF/MF vs. conventional pretreatment for a 50 mld RO feed

Component	Conv, US $,k	UF/MF, US $,k
Seawater intake and pumping	600	600
Sodium hypochlorite generation	320	320
Coagulant addition	15	15
Sulfuric acid addition	20	20
Antiscalant addition	14	14
SBS addition	10	10
Dual media filters	890	
Pumping	260	
Cartridges	240	
UF modules and racks		2000
Backwash, CEB, CIP systems		330
Instrumentation	30	30
Electrical	100	100
Mechanical installation	150	180
S/U and commissioning	40	60
Engineering and QC	500	500
Total	3189	4179

In this example, UF/MF pretreatment exceeds the cost of conventional pretreatment by approximately $1m.

In terms of running cost, the pretreatment power and chemical costs are very much lower than those of the RO. UF/MF has slightly higher power and somewhat lower chemical cost (due to reduced dosage of coagulant) than conventional pretreatment. UF/MF will save some of the RO chemicals cost, used in the process. Since this is significant, typically 5 cents/m^3, a modest saving will potentially justify the additional UF/MF capex in a whole life evaluation. This would need to be confirmed in a real life case study, but the payback of the additional capital could well be in the region of 5–10 years.

There is a replacement cost for UF/MF membranes, but at approximately 1 cent/m^3, it is almost exactly the same as the saving in RO replacement cost by having a 33% saving in RO replacement rate, e.g. 20% for conventional pretreatment being reduced to 13% for UF/MF, as explained above.

A saving not reflected in the capex comparison above is the smaller footprint of UF/MF. This will probably claw back some of the $1m cost differential, but the impact will be determined by local conditions. The other factor which is difficult to evaluate, is the security of supply issue, since on-stream time will be improved by UF/MF, but the monetary value assigned to this benefit is hard to quantify.

The final factor which will have a major influence on the complete system is the influence of UF/MF pretreatment on RO flux. A modest increase, if allowed by the feed salinity as discussed in the previous section, would easily pay for the additional capex of UF/MF. This would need to be evaluated on a case by case basis for a given feed TDS, but it is likely that this factor alone would justify UF/MF pretreatment for any feed below 35,000 ppm.

Membranes for UF/MF. Commercial UF/MF membrane systems span the range from fully hydrophilic, e.g. Cellulose Acetate (CA), to fully hydrophobic, e.g. Polypropylene (PP). Between the two extremes, there is the Polysulfone (PS)/Polyethersulfone (PES) family, Polyacrylontrile (PAN), Polyvinylidene Fluoride (PVDF), and a few other less common polymers. The most widely used polymers for UF/MF both in terms of the number of companies offering products, and the installed product base, are PES and PVDF.

Although both PES and PVDF are moderately hydrophobic in themselves, the membrane making process may utilize additional hydrophilic polymer components as pore formers or as modifiers, resulting in a final characteristic which is to some extent hydrophilic. The degree of hydrophilicity varies between products.

In water treatment, a hydrophilic membrane has some obvious advantages. Firstly, the membrane is easily wetted, and this results in high permeabilities relative to the pore size. Secondly, the organic fouling constituents often present in surface water sources would readily attach to a hydrophobic surface. A hydrophilic surface tends to resist absorptive attachment by organics, and such a surface is referred to as a low fouling surface.

In water treatment applications, both UF and MF membranes utilize a low energy process, often known as ‘direct flow,’ in which a dead end filtration cycle is followed by an intermittent backwash. The principle of direct flow is that particulates accumulated in the feed channel during the filtration cycle, are expelled by the periodic backwash. Either the backwash is carried out at a higher flux than the filtration cycle, e.g. by 2.5 times, or air is introduced into the backwash at a similar flux to the filtration cycle. The difference in energy between filtration and backwash cycles will tend to displace particles loosely lodged near the entrance of the pores. For a successful stable operation, it is important that the filtration cycle transmembrane pressure (TMP) does not increase to too high a value. A high TMP will result in particles penetrating too far into the pores, i.e., plugging the pores, and the backwash cycle will then not be fully effective at recovering permeability. The critical pressure above which the lodging of particles in pores becomes plugging is dependent upon pore size, and varies significantly between UF and MF.

If MF is operated at a low TMP, of say <0.1 bar (< 1.5 psi), pore plugging can be avoided, since the forces are too low to force the fine particles into the depth of the pore structure. However, MF tends to be operated in typical water treatment application at significantly higher TMP's than this, in an effort to reduce membrane area requirements. The preponderance of fine particles to plug the MF membrane pores results in a classic saw-tooth TMP profile, with TMP rising inexorably despite gas assisted backwash, necessitating aggressive chemical recovery cleans. The problem with this type of operation is that the feed pumping system has to be designed for the worst case TMP, leading to higher power costs.

MF tends to do better in applications where there are larger solids present in significant concentrations, especially if the solids can form a pre-coat membrane. Robust flocculated solids are particularly suitable. Under these conditions, stable performance may be possible at pressures well above the 0.1 bar (1.5 psi) quoted above. For example, wastewater applications operate well with MF membranes, giving stable performance up to at least 0.4 bar (6 psi).

Typical UF performance is characterized by stable permeability, with preventative strategies to prevent the rise of TMP as the filtration cycle proceeds. These strategies include backwashing and air or chemically enhanced backwash (AEB or CEB). Clean In Place (CIP) is available as a recovery strategy if the preventative strategies fall short, or cannot cope with an upset in feed quality or operation.

The ideal TMP for UF is normally <0.6 bar (<9 psi) to avoid fouling problems, with a top limit under normal conditions of <1.2 bar (17 psi). With TMP <0.6 bar (< 9psi), it should be possible to maintain stable performance just by use of a maintenance cleaning strategy with CEB's. As pressures rise above 0.6 bar (9 psi), there will be a progressive need for recovery cleaning with CIP's, or more frequent CEB's, with the possibility of backflush TMP's rising too high.

Modules for UF/MF. RO and NF product offerings are dominated by spiral wound elements. In contrast, UF/MF is dominated by hollow fiber/capillary modules. The term hollow fiber tends to be used for internal diameters <0.5 mm (<0.02″); capillary is used for internal diameters >1.0 mm (>0.04″). Both terms are used for the middle ground of 0.5–1.0 mm (0.02–0.04″).

The main reasons that UF/MF has developed differently from RO/NF are as follows:

- Hydrodynamic efficiency

The high permeability of UF/MF makes the spiral construction inefficient due to excessive pressure drop. Hollow fiber modules can avoid these limitations.

- Backwashability

The fact that UF membranes are made from a single polymer or polymer solution, enables the membranes to be backwashed without the potential for delamination that would occur in a multi layer flat sheet membrane, such as those used for RO and NF. The hydrodynamics of the backwashing operation, which normally takes place at a higher flow than the filtration cycle flow, provides an important advantage to the hollow fiber configuration.

- Integrity

The hollow fiber has higher seal integrity at the potting interface than the multi layer membrane used in a spiral. A small degree of leakage is inevitable in a multi layer membrane. This is due in part to the nature of the coating process used to make multi layer membranes, which gives some small defects. In addition, it is difficult to seal the ends of the substrate layer. This gives the spiral a lower degree of intrinsic integrity. More importantly, the spiral cannot be integrity tested in situ (other than by monitoring filtrate quality), or repaired. Thus for municipal water applications, where integrity has to be demonstrated both at the outset and in on-going operation, the spiral configuration is therefore unsuitable.

- Pretreatment

Spirals need good quality feed in order to prevent the feed spacer from becoming clogged with particulates. Hollow fiber and capillary modules can tolerate higher feed solids loading, and require less pre-treatment.

The feed to the hollow fiber module can either be introduced into the inside of the fiber lumen, with the permeate being withdrawn from the shell, or it can be fed to the shell, with permeate being taken from the fiber lumen. The first of these two options is the inside feed option, the second, the outside feed option.

For many commercial fibers, the surface area of the outside surface is at least twice as great as the inside surface. This means that for the same flux, the outside feed system should produce at least twice the output of an inside feed system. However, in reality, there are other constraints, which significantly limit this advantage.

The inside feed configuration has feed entering the lumen during filtration, with the flow direction being reversed during backwash. The high flow of the backwash supply to the shell arrives along the entire length of the outside surface, the main constraint to efficiency being the removal of the backwash effluent from the fiber lumen. For system designs, which operate single modules in parallel, backwash effluent, can be removed from either end of the module, ensuring an efficient removal of accumulated debris. Multiple element systems remove debris from just one end, employing by-pass tubes to help regain the loss of efficiency.

The outside feed configuration cannot be operated with acceptable efficiency using a permeate backwash due to the high pressure drop along the length of the fiber at the high flow rates. In consequence, configurations based on outside feed utilize air during the backwashing process. Either the air passes back through the fiber in the reverse of the flow direction (air backwash), or air scours the outside of the fiber whilst a low flow rate permeate backwash passes through from the fiber lumen. A true air backwash requires a fully hydrophobic fiber. It provides an effective way of removing the debris accumulated during the filtration cycle, but is mechanically aggressive to the fiber itself, and can result in stress, fatigue, and, ultimately, fiber failure.

The outside air scour option is also aggressive, causing particular stress at the tubesheet interface. The tubesheet is the area in which the fiber is bonded to the shell or manifold, normally with an epoxy resin. A strong fiber is required to withstand the stress. Both air backwash and air scour utilize an appreciable quantity of air, resulting in a significant running cost, but reduce the cost and improve the recovery associated with the permeate backwash. Due to the aggression of the air assisted backwash process, outside feed systems employ fibers known for their strength and flexibility, such as PVDF, whilst the milder conditions of the inside feed configuration allows the use of the high permeability tighter cut off PES membranes.

All UF/MF systems utilizing an inside feed configuration use a shell or housing to contain the membranes. A pressurized feed is used to supply feed to the modules, hence this type of system is known as a pressurized system. The shell has a pressure rating, and allows the system to be operated against a back pressure if required. Some pressurized systems use an outside feed configuration, but these are mainly MF systems, in which gas is used to assist the backwash process.

The other type of UF/MF system is known as a submerged membrane system. The submerged system operates with the membranes unencapsulated, and uses out an outside feed configuration. Normally, the membranes would simply be immersed in an open tank, and a vacuum applied to the permeate side to draw the filtrate through the membrane.

The main advantage of the submerged system is the saving on the cost of the pressure vessel. In addition, submerged systems can tolerate high solids without suffering the problem of lumen plugging that constrains inside feed pressurized systems. However, a major disadvantage is that the average concentration of feed near the membrane is higher than for pressurized systems, unless there is a full tank drain down at backwash or a zoned design. This is because the backwash alone does not completely remove the effluent as it does in the pressurized system. For a typical system operating with 95% recovery, the average feed concentration would be approximately three times as high for the submerged system than for the pressurized. Accordingly, submerged systems normally operate at lower flux, though the disadvantage is much less than threefold, and in fact is more typically about half that. This is because at the lower fluxes and operating pressures of submerged systems, the higher solids form an open pre-coat at the membrane surface, enabling the system to tolerate the higher solids burden.

Another aspect of the comparison is that pressurized systems use chemical soaks in a chemically enhanced backwash (CEB), and recirculatory cleaning with a Clean In Place (CIP). The ability to use two completely different methods is an advantage in improving cleaning efficiency, and provides flexibility. The submerged system also uses a frequent CEB, but the CIP is different to the pressurized system. For the submerged CIP, chemicals are introduced to the feed side either by draining the tank, or removing the membranes from the feed tank and soaking in a separate tank. This is therefore a lengthy, and potentially inconvenient process, and does not allow the straightforward efficiency advantage of the recirculating chemicals in situ.

Operating pressure however is normally a factor that favors submerged systems, since pressures at the lower fluxes are often lower, and the vacuum pressure may be provided from a natural head difference at the site.

In the economic comparison between pressurized and submerged systems, several factors therefore need to be taken into account. The submerged system saves on pressure vessels, but requires a tank. Since operating fluxes are lower, more membrane area may be required. For larger plants, and particularly in a retrofit situation, submerged systems can gain a significant capex advantage. In addition, the operating cost advantages of submerged systems are more likely to be significant for larger plants. However, small to medium plants are likely to be more suitable for pressurized technology, especially if feed quality is good, and for these plants operating costs are less important. Between these two limits, there is a large gray area, where local factors, and details of the application will determine which is the most suitable technology.

Applications. In 2004, the total UF/MF market for water and wastewater was approximately \$450m, of which 40% was in RO pre-treatment. The use of UF/MF pre-treatment prior to RO is often characterized as an Integrated Membrane System (IMS). The drive towards IMS is partly due to the improved feed water quality to the RO, but also due to the reduced risk for the RO warranty resulting from employing a barrier process in the pre-treatment stage. The main IMS applications can be categorized as follows:

- Wastewater
- Boiler feed water
- Municipal surface water
- Seawater
- General process

The use of UF/MF and membrane bioreactors (MBR's) in wastewater pre RO has already become well established, and is more widely used than conventional treatment. UF/MF simplifies the pre-treatment system, and produces better filtrate quality, since the membrane barrier achieves better removals of fine colloidal particles than several stages of a conventional system (20, 21, 22). Submerged technology is more widely used for this application than pressurized due to the high feed solids loading.

The second major application is boiler feed water treatment. Many IMS systems have already been installed in China in recent years, due to the rapid development of the power industry and a desire by the Chinese to apply the latest technology. Most of these installations, which are small to medium size, use pressurized technology even though some of the feed water quality is poor. Industrialization in other parts of the World is likely to create a similar market.

Municipal surface water applications form an important sector of IMS applications. Even if conventional treatment could be used for the RO, membranes are often favoured due to the integrity requirements of the overall treatment system. As discussed previously, it is difficult to monitor RO satisfactorily for integrity, and even if monitored, the individual elements at fault cannot be repaired.

The fourth application area is seawater. This is a newly emerging area, with the first large size IMS seawater system (26,500 m^3/day, 7 MGD) currently being commissioned in Saudi Arabia (23), and some further large projects under contract. Both pressurized and submerged technologies are likely to feature in future installations dependent upon project size (24, 25).

Submerged Membrane Process Design. Figure 8.14 shows a schematic flow diagram of a seawater RO system with membrane pretreatment. In this diagram, the membrane pretreatment is represented by submerged vacuum driven technology, which is expected to be future technology of choice for large seawater desalination systems.

In the submerged system, fibers are potted together in the form of bundles or cassettes, collected together to form a module (Fig. 8.15). The module has two water collecting manifolds, enabling filtrate to be withdrawn from and backwash flow to be provided to both ends of the bundles. The dual manifold enables lower pressure drop inside the lumen during filtration and backwash

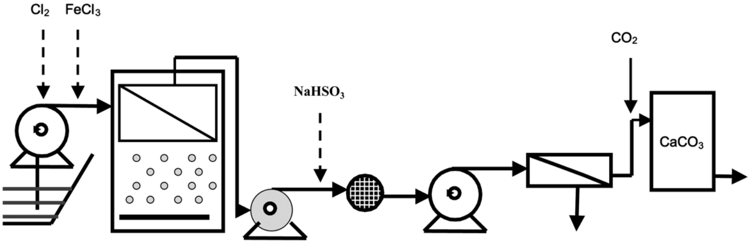
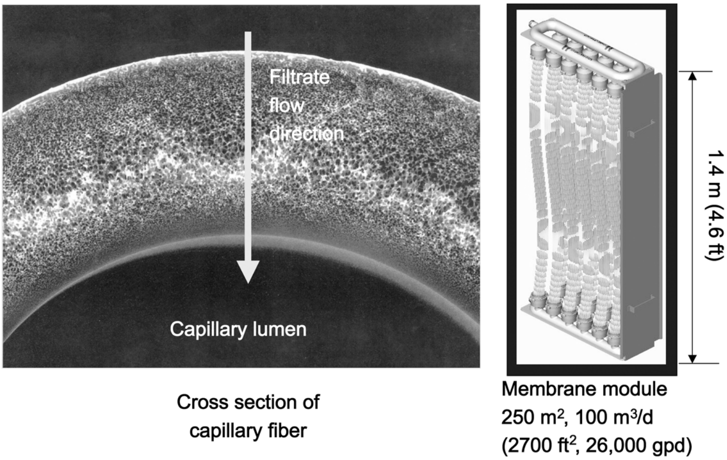


FIG. 8.14 Configuration of RO seawater system with membrane (UF/MF) pretreatment



Cross section of capillary fiber

Membrane module
250 m², 100 m³/d
(2700 ft², 26,000 gpd)

FIG. 8.15 Submersible capillary technology.

steps. A number of modules connected in parallel form a stack or block (Fig. 8.16). The membrane blocks are immersed in feed water tanks and connected to the vacuum-backwash manifold (Fig. 8.17). Feed water is continuously fed to the tank with about 95% of being removed as filtrate. The remaining 5% is bled as a waste stream from the bottom, loaded with suspended solids. Additional piping provides air for scouring of fibers.

FIG. 8.16 Submersible capillary module assembly (Courtesy Zenon Corporation).

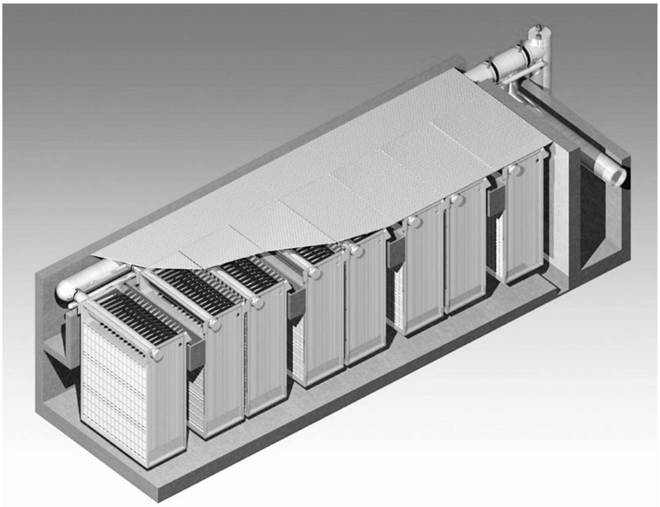


FIG. 8.17 Design concept of submersible system (Courtesy Zenon Corporation).

The operational sequence consists of filtration step followed by a filtrate backwash. The filtration step comprising suction of filtrate under vacuum of 0.07–0.5 bar (1–7 psi), lasts for 15–30 min. The filtration rates applied are 20–70 $l/m^2/h$ (12–41 gfd). During the filtration step, the surface of the fibers is scoured with air, released at the bottom of the stack to shake fibers and remove foulants. Air is provided in either a continuous or an intermittent mode. At the end of filtration interval, filtrate flow is reversed for a period of about 1 min. During the backwash the filtrate flows under pressure, applied from the lumen of the fiber at a flux rate somewhat higher than the filtration flux rate. This backwash flow of filtrate unblocks the pores and lifts deposits from membrane surface, restoring permeability.

Once to several times per day, cleaning and/or disinfecting chemicals are added to filtrate during the backwash. The duration of this chemical enhanced backwash (CEB) is longer than regular backwash as it usually includes a soaking period of 5–20 min. The chemicals added during CEB may include chlorine (or hydrogen peroxide), caustic and/or acid. Usually these chemicals are added separately. However, sometimes chlorine is added together with the caustic. If the water treated has high hardness, after CEB with hypochlorite and caustic, a backwash with acid is required to dissolve the scale formed. The CEB is applied in the same direction as backwash, from lumen (inside of the fiber) out to the tank. Periodically, when the TMP increases above 0.5 bar (7 psi) cleaning in place (CIP) is conducted. During CIP, the cleaning solution is applied from the same side as the feed water. The tank is drained and filled with cleaning solution prepared with soft water. The cleaning chemicals are the same as used during CEB, but potentially at higher concentration, and occasionally with addition of detergents. The CIP may include a few hours of soaking, preferably with warm (30–40°C) cleaning solution.

The sequence of operation of the submerged unit is listed in Table 8.2 and shown schematically in Fig. 8.18.

As indicated in Table 8.2, the submersible system has to be designed to produce a gross capacity of 10–25% higher than the nominal filtrate output. The additional capacity is necessary to compensate for off line time (backwash, cleaning and other maintenance) and the volume of filtrate required for backwash. The operating parameters that strongly affect the process economics are design filtrate flux rate, operating intervals between backwash, and frequency of cleanings. The design filtrate flux rate determines the membrane area (number of membrane modules) required to produce sufficient filtrate capacity. The length of operating intervals between backwash and the frequency of cleaning affects the on-line

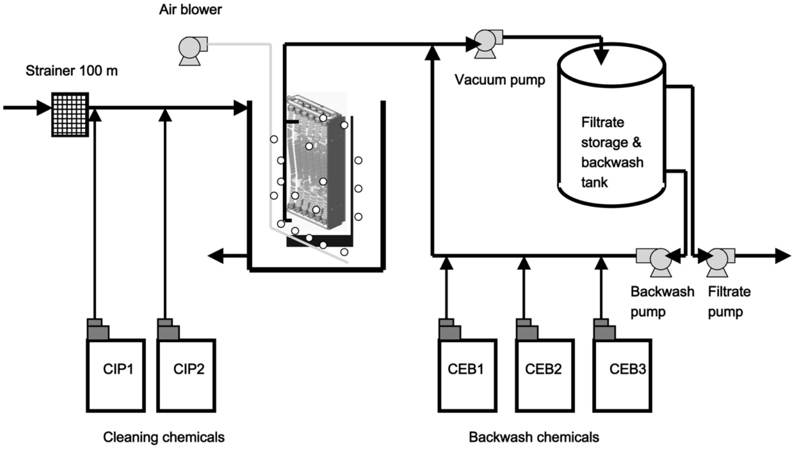


FIG. 8.18 Flow diagram of submersible capillary membrane plant.

TABLE 8.2
Operation sequence of submersible system

Process step	Objective	Duration	Frequency
Vacuum filtration	Filtrate production	15–60 min	Continuous
Pressure backwash	Foulants removal—maintaining permeability	30–60 s + 30 s for valves adjustment	Every 15–60 min
Chemical enhanced backwash (CEB)	Foulants removal—permeability restoration	1–20 min	Once—few times per day
Cleaning in place (CIP)	Foulants removal—permeability restoration	2–4 h	Every 1–6 months
Integrity test	Verification of capillary membranes integrity	20 min	Every 1–30 days
Off line time	60–180 min (4–12%)	Additional capacity required for off line time + backwash	10–22%

time factor of the plant. However, the major economic effect of these operating parameters is related to cost of chemicals, required for CEBs and CIPs.

Example 15

Vacuum driven capillary system
Filtrate output capacity: 37,850 m^3/d (10 mgd)
Feed water turbidity: average 1 NTU, maximum 5 NTU
Feed water temperature: 16–28°C (61–82°F)

Design filtrate flux:	27 l/m^2/h (16 gfd)
Backwash flux	82 l/m^2/h (48 gfd)
Backwash frequency	30 min
Backwash duration	1.0 min
Off line time due to backwash	1.5 min
Integrity test frequency	every 7 d
Integrity test duration	20 min
CEB frequency per day	1
CEB duration	20 min
Cleaning frequency	every 30 d
Cleaning duration	4 h (8 min/d)
Membrane area per module	200 m^2 (2170 ft^2)
Number of modules per block	10

Total backwash duration, min/day
1.5(1440 – 20/7 – 20)/(30 + 1.5) = 1.5 × 45 = 68
On line time, min/day
1400 – 20/7 – 20 – 68 – 8 = 1341 (93%)
Filtrate capacity required
Backwash capacity
(45/1440)(82/27) = 0.10 (10%)
Total filtrate capacity
1/0.93 + 0.10 + 0.05 (contingency) = 1.22
3785 0 × 1.22 = 46,177 m^3/d (12.2 mgd)
Membrane area required
46,177 × 1,000/(24 × 27) = 71,300 m^2
12.2 × (1,000,000/16) = 762,500 ft^2
Number of modules, each 200 m^2 (2170 ft^2) = 360
Number of membrane blocks (10 modules) = 36

The pretreatment of feed water to a UF/MF submerged system consists of equipment that provides screening of particles that could otherwise damage polymeric fibers. This is usually accomplished with self cleaning strainer, rated at about 200 µ. In some systems, after the screening equipment, coagulant is added to the feed water. Then water flows into tanks where capillary modules are submerged. In large submerged systems, the tanks are of a similar configuration to gravity sand filters. However, the submerged membranes process produces a much higher output per tank unit volume compared to gravity sand filters. The footprint required for submersible membrane system depends very much on membrane module packing density, configuration and filtration flux. However, in most cases it is about 50–80% of conventional sand filtration system footprint for the same filtrate capacity (25).

The major equipment items in a submerged system include the filtrate pumps. The pumps draw water through the membranes and discharge it to the clear well. The filtrate pumps used are of a conventional type, that are able to operate under negative suction pressure of up to 1 bar (15 psi). During operation at negative suction pressure, some air dissolved in the water will be released. This air is accumulated in vertical tanks (air receivers) attached to the filtrate pumps and periodically purged from the piping manifold. Additional major equipment includes backwash pumps, sized for a pressure of about 3 bars (45 psi). The system also incorporates a significant number of automatic valves for isolation of membrane trains and providing water flow direction according to the operation sequences. System operation is managed through a central PLC that receives input from flow, pressure, temperature and water quality (turbidity and/or particle counters) sensors, and controls operation of pumps and valves.

Pressurized membrane technology. In pressure driven capillary membranes, feed pressure of up to 2 bar (29 psi) is applied to the feed water to create water flow through the MF/UF membranes. An example of one module configuration is shown in Figs. 8.19 and 8.20. The module diameter is about 225 mm (9″) and the length is 1000 mm or 1500 mm. The membrane area of the 1000 mm long module is 30 m^2 (325 ft^2) and 46 m^2 (500 ft^2) for the 1500 mm long model. In field applications for RO pretreatment the module operates in a flux range typically between 60 – 100 $l/m^2/h$ (35–60 gfd), though higher fluxes are achievable for low solids feeds.

The module has three ports: feed, concentrate and filtrate. The mode of operation is very similar to the submerged system. There is alternate sequence of filtration followed by backwash steps. The backwash step is conducted at a pre-

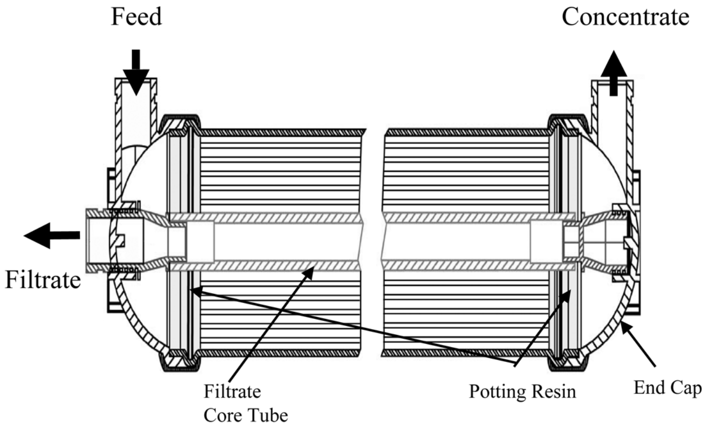


FIG. 8.19 Configuration of pressure driven UF/MF membrane module.

FIG. 8.20 Pressure driven capillary UF module.

determined frequency, sometimes with the presence of disinfectant. Periodically or whenever trans-membrane pressure (TMP) increases above the preset limit (usually 1.1 bar, 16 psi), a CIP is conducted. During the filtration step, which lasts for 15–60 min., the feed port is opened and the concentrate port closed, so that the module operates in a direct filtration mode (100% recovery). Feed flows under pressure to the lumen of capillaries, permeates through the capillary walls and is collected in central filtrate tube. The filtrate leaves the module through the permeate port. In the backwash step, the filtrate flow direction is reversed, flowing under pressure from the filtrate tank, through the filtrate port to inside

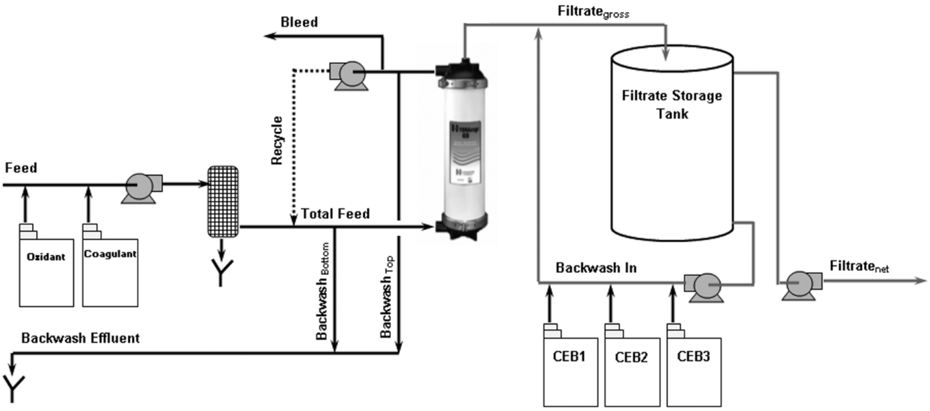


FIG. 8.21 Flow diagram of pressure driven capillary membrane unit.

of the module and then through the capillary walls into the lumen. The backwash water leaves the module through either feed or concentrate port. The flow diagram illustrating unit operation is shown in Fig. 8.21.

The pressurized units are configured as a horizontal or vertical parallel array of modules, connected to a common manifold. An example of pressure driven ultrafiltration unit is shown in Fig. 8.22. At present, most pressurized systems are UF, with the main application being potable water production.

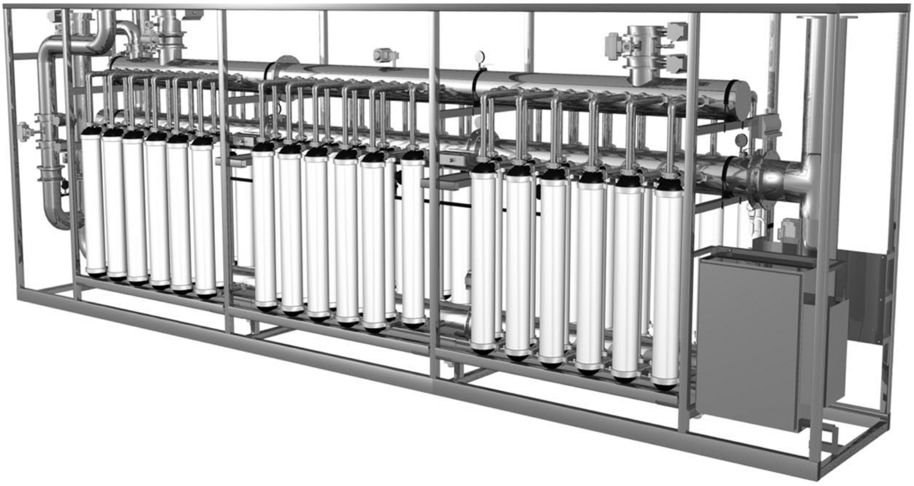


FIG. 8.22 Pressure driven capillary unit. Hydrablock with 66 (expandable to 72). Hydracap with UF modules. Filtrate capacity 29,000 m^3/d (7.7 MGD).

Membrane integrity verification. The unique premise of UF and MF technology is that it provides a membrane barrier and, potentially, an almost absolute removal of particles down to a submicron range. It is not surprising therefore that verification of membrane integrity is an important issue for these two technologies. Integrity verification methods for membrane systems are described in ASTM method (40). Some of integrity verification tests can be applied only when the system is off line, whilst others can be conducted during system operation.

1. Off line tests
   Bubble point test
   Pressure hold test
   Diffusive air flow test
   Vacuum hold test

2. Continuous (on line) tests
   Particle passage counting/monitoring
   Marked particle passage
   Turbidity measurements
   Acoustic sensing

The most frequently used integrity verification method is the pressure hold test. In this method water is purged from the system using clean, oil free air. Then air pressure at 0.3–1 bar (5–15 psi) is applied to the feed side of the membrane. After the feed and concentrate ports are closed, the decay of air pressure is monitored. The decay of applied air pressure is a result of air diffusion through the membrane and air flow through the defects. The integrity of the membrane filtration device is adequate if the pressure decay is typically less than 10% of the initial pressure within a period of 5 min (the decay is dependent upon membrane, module and pipework hold up). The ability to differentiate between diffusive flow and air flow through the broken fibers is a function of number of modules or membrane area in the system segment that is being tested for integrity. The number will vary for different membrane modules but the limit corresponds to a membrane area of about 1000–2000 m^2 (11,000–22,000 ft^2), depending on the level of integrity being established. If the pressure decay is faster than expected for diffusive flow through the capillary walls, modules with defective fibers are located by pressurizing individual modules and looking for air bubbles. The pressure at which air will pass through membrane defects is related to diameter of the defects according to equation:

$$P = 4k\gamma\cos\theta/d \qquad (27)$$

where P = bubble point pressure
k = correction factor for the pore irregularity
λ = surface tension of the fluid (water)
θ = contact angle
d = pore diameter

For hydrophilic membranes (contact angle = zero), round pores and a surface tension of water of 72 dynes/cm, the pore diameter (d) determination equation can be simplified to:

$$d = 288/\Delta P_{max} \tag{28}$$

In actual application, the bubble shows at much lower pressure then calculated by Eq. 27. This may be due to the irregular shape of membrane defects and incomplete wetting of pores (indicating a partial hydrophobic character of membrane material). Nevertheless, the pressure decay test and bubble test is probably the most effective method of membrane integrity determination. Since the diameter of capillary fibers is very low, a small number of broken fibers will have only moderate effect on membrane integrity, as shown in the example 16. This example shows calculations of water flow through a fiber that was cut in the middle. This flow is compared with the total filtrate flow. The number of capillary fibers in one module is quite large: 10,000–20,000 fibers. Therefore, the flow through one cut fiber is very low compared to overall filtrate flow from the module. In addition, because the internal diameter of the capillary is small, it is very likely that the cut fiber will be quickly plugged by particulate matter in the water. For this reason, a few membrane defects will have only a marginal effect on the quality of RO feed water produced in the membrane pretreatment system. Still, conducting an integrity test periodically is important since it provides an indication of the overall condition of the capillary membranes in the system.

Example 16

Integrity evaluation
Module membrane area: 46m^2, fiber length 1.25m, ID 0.8 mm
Number of fibers in one module: $46/(1.20 \times 3.14 \times 0.8 \times 10^{-3}) = 15{,}200$
Calculation of bypass flow through one broken fiber at TMP = 50kPa.
Q = pd4TMP/(128Lm)
$Q = 3.14(0.8 \times 10^{-3})4 \times 50/(128 \times 1.62 \times 10^{-3})\ (1/0.65 + 1/0.65) \times 1000 \times 3600$

$Q = 3.4310^{-3}$ l/h

Filtrate flow of element @ 85 $l/m^2/h = 46m^2 \times 85\ l/m^2/h = 3910$ l/h

Log of flow ratio = $\log(3910/3.4310^{-3}) = 6.06$

TABLE 8.3

Pore diameter and corresponding bubble point pressure.

Pore diameter, micron	Bubble pressure, bar (psi)
0.05	57.6 (835)
0.10	28.8 (418)
1.00	2.88 (42)

TABLE 8.4

Effectiveness of particles rejection vs. number of broken fibers

No. of broken fibers in element	1	2	5	10	100
Log removal	6.1	5.8	5.4	5.0	4.0

9

Chemical stabilization of permeate

The permeate produced in brackish and seawater RO systems contains low concentration of alkalinity and hardness. Its buffering capacity is low and CO_2 presence results in decreased pH. Such water if sent directly, without treatment, to the distribution piping would have tendency to dissolve the protective calcium carbonate deposit from piping walls and would initiate corrosion. Therefore, RO permeate has to be stabilized by increasing concentration of alkalinity and hardness at the exit from the desalination plant. If RO permeate originates from anaerobic feed water source, then residual H_2S present in the permeate water has to be removed as well, prior to entrance of water to the distribution network. The common method of H_2S removal is aeration, conducted at permeate water pH lower than 6.5. Effectiveness of H_2S removal by air stripping is very high. Any residual H_2S after aeration can be converted to sulfuric acid by oxidation using chlorine gas or hypochlorite solution, according to reactions:

$$H_2S + 4Cl_2 + 4H_2O = H_2SO_4 + 8HCl \tag{29a}$$
$$H_2S + 4NaOCl = H_2SO_4 + 4\ NaCl \tag{29b}$$

Oxidation with hydrogen peroxide yields predominantly elemental sulfur according to equation:

$$H_2S + H_2O_2 = S + 2H_2O \tag{29c}$$

In brackish RO systems, in absence of H_2S, the permeate stabilization process includes mainly removal or neutralization of CO_2. Removal of CO_2 is effectively accomplished by aeration using either forced draft or induced draft degasifiers. Both degasifier types have capability to effectively reduce CO_2 concentration below 10 ppm.

If increase of alkalinity concentration is required for water stabilization, it could be achieved by addition of calcium hydroxide, calcium carbonate or sodium hydroxide to permeate water containing sufficient concentration of CO_2 (64).

$$2CO_2 + Ca(OH)_2 = Ca(HCO_3)_2 \tag{30}$$
$$CO_2 + CaCO_3 + H_2O = Ca(HCO_3)_2 \tag{31}$$
$$CO_2 + NaOH = NaHCO_3 \tag{32}$$
$$CO_2 + Na_2CO_3 + H_2O = 2NaHCO_3 \tag{33}$$

Seawater feed has low alkalinity (about 150 ppm). Therefore, permeate from seawater RO system has both low alkalinity and low CO_2 concentration. Due to very high rejection rate of seawater membranes, concentration of calcium in permeate stream is very low as well, in the range of 1–3 ppm. The objective of permeate treatment is to stabilize water composition by increasing concentration of calcium and alkalinity and increasing pH to achieve positive value of Langelier Saturation Index (LSI). In majority of seawater systems concentration of CO_2 in permeate is too low to form sufficient concentration of alkalinity to achieve positive LSI. Therefore, concentration of CO_2 has to be increased by injecting CO_2 gas into the permeate water prior to addition of caustic soda or lime. The chemicals added for permeate stabilization result in TDS concentration increase. Table 9.1 summarizes concentration changes due to addition of common chemicals used for permeate stabilization.

As shown in Table 9.1 the chemical system most effective in increasing hardness and alkalinity is one that involves reaction of CO_2 with $CaCO_3$. The decision which chemicals to use is mainly driven by the local availability and cost. In cases when only increase of hardness is required the simplest approach

TABLE 9.1
Concentration changes per unit of CO_2 neutralized, (gram/gram or lb/lb)

Reagent	Usage	Compound formed	Hardness increase	Alkalinity increase	TDS increase, ppm
$Ca(OH)_2$	0.84	$Ca(HCO_3)_2$	1.13	1.13	1.84
$CaCO_3$	2.27	$Ca(HCO_3)_2$	2.27	2.27	3.27
NaOH	0.91	$NaHCO_3$	0.0	1.13	1.91
$Na2CO_3$	2.40	$NaHCO_3$	0.0	2.27	3.40

is to react calcium carbonate with sulfuric acid, produce $CaSO_4$ solution and add it to the permeate to achieve sufficient calcium concentration. During process design, the increase of permeate TDS due to permeate post-treatment, has to be taken into consideration in calculation of resulting final products salinity.

In majority of locations potable water in distribution systems has to contain residual of a disinfectant: free chlorine, chloramines or chlorine dioxide. The concentration level of disinfectant is determined by the Contact Time (CT), expressed in mg/l-min. For example CT = 100 could be concentration of free chlorine of 5 mg/l and residence time prior to point of use of 20 min. The CT required for a given deactivation credit for viruses and pathogens is significantly higher for chloramines then for free chlorine. The relations for CT and deactivation credits could be found in USEPA Guidance Manual (30) or other relevant local regulations.

10

RO/NF system design parameters

10.1. Feed water types

The composition and quality of water, considered for processing by reverse osmosis is influenced by its origin. For reverse osmosis application water of interest is the one with ions composition that exceeds potable water limits. In brackish water RO applications the ions that commonly exceed potable water limits are calcium, magnesium, sulfates and chlorides. Less common dissolved constituents of brackish water, that may require reduction of concentration, are fluoride and nitrate. Some waters may contain also excessive concentrations of iron, manganese, organic matter, color, hydrogen sulfite and sometimes even radioactive isotopes.

Potable water limits are specified by World Health Organization (26) and regional health authorities. Accordingly, one may define brackish water as any water of composition of soluble species exceeding potable water limits. The potable water limits, or acceptance of water composition for potable application can vary from country to country, according to local affordability of treatment methods. However, it is commonly accepted that water of salinity exceeding 1000 ppm is considered as brackish and requires treatment for salinity reduction.

The upper limit of water salinity that can be effectively treated with brackish RO membranes, in a single pass configuration, is about 10,000 ppm. On the low end of salinity spectrum there are some water sources that have salinity in the potable range but still require membrane treatment. This is usually due to presence of excessive concentration of hardness, iron, organics and/or color. The low salinity water sources are usually treated with loose RO membranes, commonly called as softening or nanofiltration membranes. More recently, presence of pesticides at very low concentration in otherwise potable quality waters, is case of concern and could require membrane treatment. The ideal membrane for such application should have high rejection of organic pesticide and preferably very high passage of all dissolved ions. This also would be a nanofiltration application.

The composition of brackish water can vary widely. The composition is usually specific for the aquifer it originates from. If brackish water aquifer is very large and/or water is pumped at the rate it is being replenish by natural infiltration, then the composition remains stable. In case of excessive pumping the composition may change. In case of utilization of coastal aquifer there is possibility of seawater intrusion and salinity increase. For other locations there will be influence on composition from adjacent underground bodies of water due to hydrostatic pressure difference.

Seawater sources are characterized by high salinity, in the range of 30,000 ppm TDS to 47,000 ppm TDS. The ion composition includes mainly sodium and chloride, about 85% combined. The remaining fraction consists of sulfate (~8%), magnesium (~4%), calcium, potassium (~1.2% each) and bicarbonate (~0.6%). Boron is one of low concentration constituents. It is present in seawater at concentration of about 3–5.5 ppm. However, boron concentration is becoming increasingly important parameter of the process design as its concentration is being specified in RO permeate. Due to relatively low rejection by RO membranes of boron species existing in seawater, stringent boron specifications have significant effect on process design and product water cost. At majority of locations the ions composition of seawater is quite consistent and fluctuates in narrow range. Temperature of seawater usually reflects the seasonal fluctuations of ambient temperatures but could be affected by temperature of local water currents. At some locations, where rivers discharge or rain surface run off is present, the fluctuations of salinity and concentration of suspended matter could vary in a wide range and require careful consideration during the process design stage.

As a part of the design process of RO plant the feed water sources should be evaluated to determine the following:

1. How feed water ions composition and temperature will affect quality of permeate and the required feed pressure?
2. Does water source contain sparingly dissolved species at concentration that could result in membrane scaling at the design recovery rate?
3. Does water source contain particulate matter that could plug feed channels of membrane elements?

4. Does water source contain organic matter at concentration that could adsorb on membrane surfaces and result in significant permeability decline?
5. What is the level of biological activity? Can it result in biofouling of membranes?

Preliminary determination of the suitability of given water source for RO processing and requirements of the pretreatment process, is conducted based on results of analysis of water samples from the water source under consideration. A listing of typical water parameters being analyzed is provided in Fig. 6.1.

During the initial evaluation of water analysis it is important to check if the analysis report contains values of important water quality parameters and concentration of major ions. The primary group of water composition data includes: pH, temperature, turbidity, electric conductivity and concentrations of Ca, Mg, Na, K, HCO_3, SO_4, Cl and SiO_2. Additionally, concentration of any species that their maximum concentration had been defined in permeate, should be also determined in the feed water source, for example concentration of NO_3 or B. In some cases, mainly in specifications of nanofiltration projects, minimum concentrations of hardness in permeate is being listed as system performance requirements. The analysis should be balanced,, i.e., sum of miliequivqlents of positively charged ions (cations) should be similar to the sum of equivalents of negatively charged ions (anions). Data entry screens of majority of membrane performance projection computer programs converts concentration in ppm to miliequivalents and display the results automatically. If the difference of corresponding sums of positive and negative miliequivqlents is within ±5%, the analysis is being considered as balanced. Some computer programs include option for automatic balance of negative or positive ions concentrations as necessary.

10.2. Feed water composition

Based on the feed water salinity range a suitable membrane element type is selected for performance calculations (i.e., NF, RO or seawater). The feed water composition and temperature, together with recovery rate and average flux rate, are used as an input parameters in calculations of projected permeate quality and feed pressure. At given operating parameters and membrane type the permeate concentration is direct function of feed salinity (Table 10.1 shows representative values of ion passage in RO systems treating low salinity and high salinity brackish feed and medium salinity range of seawater feed). However, careful examination of experimental data for brackish composite polyamide membranes shows that there is anomaly of salt passage at very low and very high feed salinities (27). It is shown in Fig. 10.1 that for brackish RO membranes there is a minimum of salt passage at feed salinity of about 500 ppm. In the narrow range below this salinity, a steep increase of salt passage is observed. There is also increase of salt passage at higher feed salinities, however it is at a lower rate as compared to the very low salinity range. Eventually, at high feed salinities of about 5,500 TDS ppm the salt passage levels off. This effect of salt passage increase should be consider in calculations of permeate salinity. It is quite important in the low salinity range for ultrapure water applications such as production of high pressure boilers make up and process water for semiconductor industry.

The salt passage increase with increased salinity occurs in almost all RO brackish water systems. Even in systems operating with low feed salinity, due to the effect of recovery, the elements positioned close to the concentrate end, are exposed to high salinity water. For seawater membranes, in the common feed salinity range, the effect of salinity on salt passage is not as significant as for brackish systems and is usually neglected in permeate salinity calculations.

In brackish applications salt passage is also affected by the ion composition of the feed water. In the process of membrane elements manufacturing, salt rejection is determine using single salt solution: NaCl. In field application feed water is composed of variety of ions, which diffuse through the membrane at

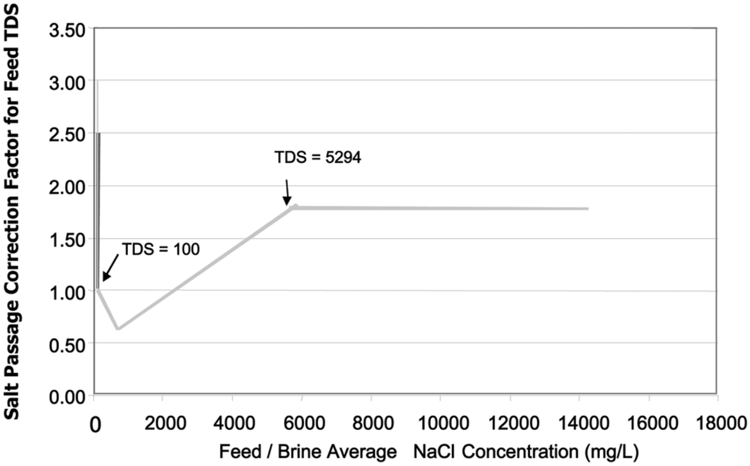


FIG. 10.1 Effect of feed salinity on osmotic pressure and required feed pressure in RO unit.

TABLE 10.1

Illustration of potential feed concentration reduction in an RO systems

(1) For the purpose of illustration of potential concentration reduction, passage is being calculated relative to the concentration in the feed stream. These values of ions passage are higher than the usual values calculated relatively to the average feed (average of feed and concentrate)

	Low salinity brackish			High salinity brackish			Medium salinity seawater		
Feed constituent	Feed, ppm	Permeate, ppm	Passage, % (1)	Feed, ppm	Permeate, ppm	Passage, % (1)	Feed, ppm	Permeate, ppm	Passage, % (1)
Ca	83	0.6	0.7	254	1.6	0.6	449	0.8	0.2
Mg	66	0.5	0.8	120	0.8	0.7	1387	2.4	0.2
Na	350	11.5	3.3	1214	36.6	3.1	12179	103.5	0.8
K	4	0.2	5.0	12	0.4	3.3	418	4.4	1.0
HCO_3	340	10.6	3.2	100	5.0	5.0	158	2.2	1.4
SO_4	74	0.3	0.4	778	5.5	0.7	3200	6.2	0.2
Cl	600	10.5	1.8	2034	56.9	2.8	21555	166.5	0.8
F	0.8	0.03	0.4	2.4	0.1	0.4	1	0.01	1.0
NO_3	46	5.6	12						
B							5	0.9	18
SiO_2	29	0.5	1.7						
TDS	1563.8	39.83	2.5	4514.4	106.9	2.4	39347	286.	0.7

different rates. In addition membrane surface of polyamide membranes is negatively charged which contributes to salt rejection through repulsion of negatively charged ions. When treating a solution of mixed ions composition the ion passage is affected by counter ions charge balance (Donnan effect). Therefore presence of highly mobile ions (like HCO_3^-) will increase passage of positively charged ions, contributing to increasing salt passage. Conversely, significant concentration of large, slowly diffusing ions (like SO_4^-) will retard passage of associated cations and resulted in lower permeate salinity. In addition, high concentration of ions with high positive charges, such as calcium and magnesium, will tend to neutralize negative membrane surface charge, reducing ions rejection that is due to surface-anions charge repulsion. Due to above effects, salt passage of brackish membranes in field conditions could differ significantly from the nominal values, obtained during testing with solution of NaCl. The effect of feed water composition on salt passage is most pronounced in softening (nanofiltration) membranes. The salt rejection of nanofiltration membranes is low and most of them have high surface charges. Therefore, exposure to mixed ion solution with high fraction of concentration of divalent cations will result in significant salt passage increase. In seawater applications salt rejection of seawater

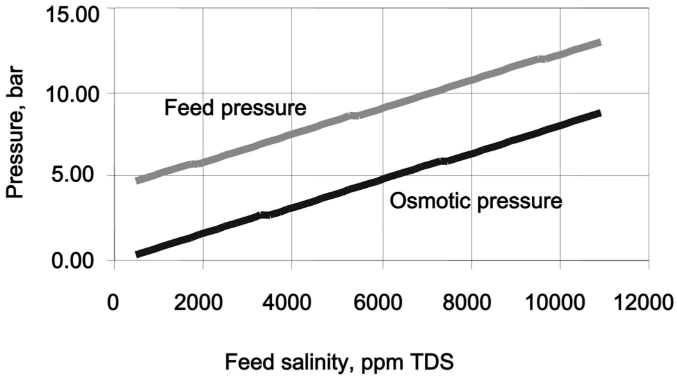


FIG. 10.2 Effect of feed water temperature on permeate salinity in a brackish RO unit. Permeate average flux 28l/m^2/h.

elements correlates well with the nominal values. This is most likely because the feed water is composed mainly of NaCl (~85%) and the membrane barrier of seawater membranes have high level of polyamide polymer crosslinking (lower density of surface charges compared to NF membranes). Feed salinity affects the required feed pressure through the value of average feed osmotic pressure (Eq. 8). In multistage brackish water systems increased osmotic pressure and pressure drop reduces NDP in each subsequent element. Consequently, permeate flux in elements in tail positions (see Figs. 84, 85) could be quite low. Therefore, feed pressure requirements for a given permeate capacity has to be adjusted according to the feed salinity. The effect of salinity on required feed pressure demonstrates itself both in brackish and seawater systems (Fig. 10.2).

10.3. Feed water temperature

Feed water temperature affects both permeate salinity and the required operating pressure. Both water and salt transport follow a similar trend (Eq. 16, Fig. 2.5). The commercial RO systems are designed to operate at constant permeate capacity (constant permeate flux). Therefore, increase of feed water temperature will result in increased permeate salinity (higher quantity of salt will pass the membrane and will be dissolved in a constant volume of permeate). This increase is about 3% per degree C. The effect is similar for brackish and seawater systems (Figs. 10.3 and 2.6). The changes of water permeability with temperature affects the net driving pressure required (Eq. 18). However, in par-

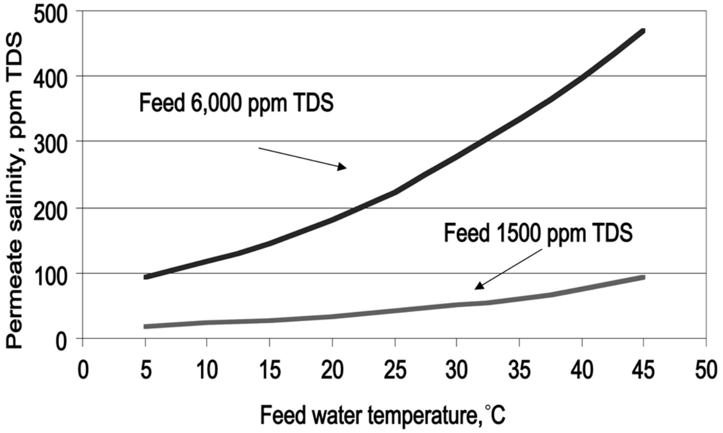


FIG. 10.3 Effect of feed water temperature on required feed pressure in a brackish RO unit. Average permeate flux 28 l/m^2/h.

allel, temperature also affects osmotic pressure of the feed water (Eq. 2). With temperature changes net driving pressure and osmotic pressure have opposite effect on feed pressure (Eq. 8). An increase of feed water temperature at low temperature range (~up to about 30°C) enables production of a given permeate flow at reduced feed pressure both in seawater and brackish water systems. At higher temperatures the reduction of feed pressure in seawater systems levels off, mainly due to increase of osmotic pressure of the average feed (Figs, 10.4 and 10.5).

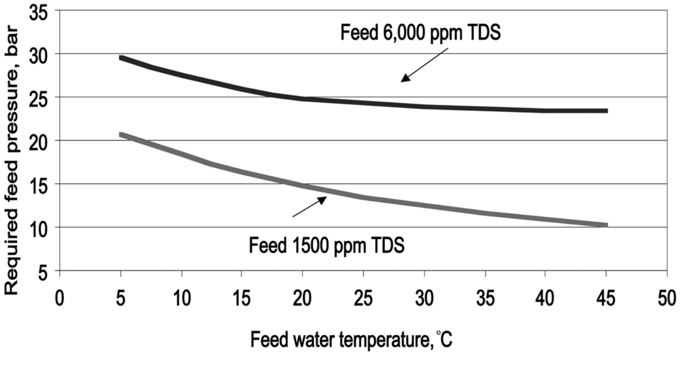


FIG. 10.4 Effect of feed water temperature on required feed pressure in a brackish RO unit. Average permeate flux 28 l/m^2/h.

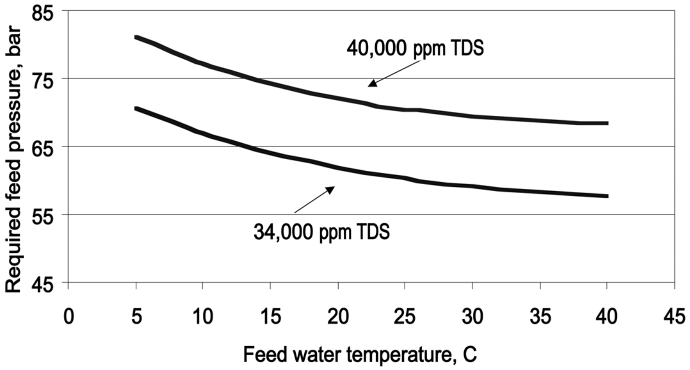


FIG. 10.5 Effect of feed water temperature on required feed pressure in a seawater RO unit. Recovery rate 50%, average permeate flux 14 l/m^2/h.

10.4. Sparingly soluble constituents

During the RO process concentration of all constituents increases due to reduction of the feed water volume. This increase of concentration is function of permeate recovery (Eqs. 6, 7). Some of the constituents presented in natural waters can precipitate if the concentration product of salt forming ions exceeds its solubility product: K_{sp}. For a given salt of composition C_mA_n in equilibrium of solid phase salt (S) with dissolved ions, the K_{sp} is defined as:

$$C_mA_n(\mathrm{S}) = mC^{+n} + nA^{-m} \qquad (34)$$

$$\mathrm{K}_{sp} = [C^{+n}]^m\,[A^{-m}]^n \qquad (35)$$

$$\mathrm{SI} = [C]^m\,[A]^n/\mathrm{K}_{sp} \qquad (36)$$

Where C stands for cation and A for anion, m and n are valency coefficients. Brackets [ ] indicate molar concentration of a given ion in solution. SI is the saturation index, indicating excess concentration of a given salt in comparison to its saturation value.

The K_{sp} is determined through measurement of ions concentrations in solution at saturation conditions (in equilibrium with solid phase). K_{sp} value is specific for a given salt and it is function of temperature and ionic strength of the solution. In brackish systems, treating natural waters, the salt of concern is mainly calcium carbonate. Less frequently calcium sulfate and silica are at concentrations that my result in scale formation. In very rare cases barium sulfate, ferrous sulfide, and ferrous carbonate could be present at concentrations that may form scale at high recoveries. In RO systems treating municipal effluents calcium phosphate sometimes forms in the tail elements. Calcium carbonate is the most common scaling constituent in brackish waters but also the easiest to control either with pH adjustment or use of scale inhibitor. In solution, calcium ions are in equilibrium with bicarbonate and carbonate species as shown in the following equations:

$$OH + H = H_2O \tag{37}$$

$$H_2CO_3 = H^+ + HCO_3^- \tag{38}$$

$$HCO_3^- = H^+ + CO_3^{-2} \tag{39}$$

$$Ca^{+2} + CO_3^{-2} = CaCO_3 \tag{40}$$

The calcium carbonate system is quite complex. Saturation conditions are not just function of concentrations of Ca, CO_3 and HCO_3 ions but also influenced by concentration of hydrogen ion (pH). Attempts to define relations for saturation conditions in potable water networks lead to development of number of saturation indexes. The calcium carbonate saturation index developed by Langelier (31) for potable water networks has been adopted by RO industry as an indicator of saturation conditions in concentrate stream of brackish water RO systems. The Langelier Saturation Index (LSI) is calculated according to relations:

$$LSI = pH - pH_s \tag{41}$$

Where pH is the actual pH of the water and pH_s is pH that corresponds to saturation concentrations of ions forming calcium carbonate.

$$K_2 = [H^+]\ [CO_3^{-2}]/[HCO_3^-] \tag{42}$$

$$K_{sp} = [Ca^{+2}]\ [CO_3^{-2}] = [Ca^{+2}]\ [HCO_3^-]\ K_2/H_s^+ \tag{43}$$

$$H_s^+ = [Ca^{+2}]\ [HCO_3^-]\ K_2/K_{sp} \tag{44}$$

$$-\log[H_s^+] = pH_s = -\log[Ca^{+2}] - \log[HCO_3^-] + \log\ [K_{sp}/K_2] \tag{45}$$

$$LSI = pH - pH_s = pH - pCa - pAlk + pK \tag{46}$$

Where K_2 is second dissociation constant to carbonic acid (H_2CO_3), K_{sp}–solubility constant of calcium carbonate at given pH and temperature.

Other parameters represent molar concentrations of relevant species in the solution.

Few years after introduction by Langelier the saturation index was modified by Larson and Buswell (32) to account for ionic strength in a form of salinity correction factor:

$$pH_s = (9.3 + A + B) - (C + D) \tag{47}$$

$$\text{where } A = (\text{Log}_{10}\,[\text{TDS}] - 1)/10 \tag{48}$$

$$B = -13.12 \times \text{Log}_{10}(°\text{C} + 273) + 34.55 \tag{49}$$

$$C = \text{Log}_{10}[\text{Ca}^{+2} \text{ as } \text{CaCO}_3] - 0.4 \tag{50}$$

$$D = \text{Log}_{10}[\text{alkalinity as } \text{CaCO}_3] \tag{51}$$

The parameter A is related to ionic strength of the solution. The value of A increases with increase of salinity. Parameter B reflects the changes of calcium carbonate solubility and changes of equilibrium of carbonic acid dissociation with temperature. The value of B decreases with temperature increase. In practical applications the LSI is either calculated using computer programs or monograms based on pH and composition of concentrate stream. Water solution has potential for $CaCO_3$ scaling at LSI > 0 and it is assumed that saturation prediction using LSI is reliable up to salinity of about 5000 ppm TDS.

For high salinity and seawater applications the LSI was modified to account for increased ionic strength by Stiff and Davis (33). The Stiff and Davis saturation index (SDSI) introduces empirical constant K in calculations pH_s to account for high ionic strength of seawater concentrates. Stiff and Davis determined value of K experimentally in the high range of ionic strength, that covers salinities encountered in RO seawater applications.

$$\text{SDSI} = \text{pH} - (9.3 + \text{K} - \text{pCa} - \text{pAlk}) \tag{52}$$

where pH has the same meaning as in LSI equation and K is a constant found in monograms.

For high salinity solutions (seawater concentrates) the SDSI value is about 1–1.3 units lower then calculated according to LSI relations.

Example 17

Calculation of Langelier saturation index

Brackish water system is design to operate at recovery ratio of 80%.

Feed water feed has TDS = 1700 ppm, ionic strength 0.04 and the following concentrations of the relevant ions:
$Ca = 300$ ppm, $HCO_3 = 250$, pH = 7.3, temp = 22°C.
After acidification and pH adjustment to 6.5, $HCO_3 = 166$ ppm, $CO_2 = 78$ ppm
Concentration factor = $1/(1 - R_p) = 5$
Approximate concentrations in the concentrate:
TDS = 8500 ppm, ionic strength 0.18, Ca = 1500 ppm, $HCO_3 = 830$ ppm, $CO_2 = 78$ ppm

Calculation of concentrate pH.
$pH = pK_1 + \log([HCO_3]/[CO_2]$
K_1 is the equilibrium constant of the carbonic acid dissociation reaction.
At 25°C its value is 4.2×10^{-7}, $pK_1 = 6.37$ (neglecting effect of water ionic strength on K_1).
Concentrate pH = $6.37 + \log((830/78)(44/61)) = 6.37 + 0.89 = 7.26$

Calculation of pH_s and LSI
Following Eqs. 41 and 47–51:
pH_s (feed) = $9.3 + 0.22 + 2.15 - 2.48 - 2.13 = 7.06$
LSI (feed) = $6.5 - 7.06 = -0.56$
pH_s (concentrate) = $9.3 + 0.29 + 2.15 - 3.17 - 2.83 = 5.74$
LSI (concentrate) = $7.26 - 5.74 = 1.52$

Calculation of SDSI
K value for ionic strength of 0.04 and 0.18 and, from monogram (41) is 2.4 and 2.8 for feed and concentrate respectively.
According to Eq. 50
SDSI (feed) = $6.50 - (9.3 + 2.4 - 2.48 - 2.13) = -0.59$
SDSI (concentrate)= $7.26 - (9.3 + 2.8 - 3.17 - 2.83) = 1.16$

Example 18

Seawater water system is design to operate at recovery ratio of 50%.
Feed water feed has TDS = 40,000 ppm, ionic strength 0.69 and the following concentrations of the relevant ions:
$Ca = 450$ ppm, $HCO_3 = 150$, pH = 8.1, temp = 22°C.
After acidification and pH adjustment to 6.5, $HCO_3 = 115$ ppm, $CO_2 = 42$ ppm.
Concentration factor = 2
Approximate concentrations in the concentrate.
TDS = 80,000 ppm, ionic strength 1.39, Ca = 900 ppm, $HCO_3 = 230$, $CO_2 = 42$ ppm, pH = 7.2

Calculation of concentrate pH
Concentrate pH = 6.37 + log((900/42)(44/61)) = 6.37 + 0.6 = 6.97

Calculation of pHs and LSI
Following Eqs. 41 and 47–51:
pH_s (feed) = 9.3 + 0.36 + 2.15 – 2.65 – 1.97 = 7.19
LSI (feed) = 6.5 – 7.19 = – 0.69
pH_s (concentrate) = 9.3 +0.39 + 2.15 – 2.98 – 2..28 = 6.58
LSI (concentrate) = 6.97–6.58 = 0.39

Calculation of SDSI
K value for ionic strength of 0.69 and 1.39, from monogram (41) is 3.3 and 3.5 for feed and concentrate respectively.
According to Eq. 50
SDSI (feed) = 6.50 – (9.3+ 3.3 – 2.65 – 1.97) = –1.48
SDSI (concentrate) = 6.97 – (9.3 + 3.5 – 2.98 – 2.28) = –0.57

As indicated by results included in examples 16 and 17, LSI and SDSI have similar values at low ionic strength. At high ionic strength of solution, the values of SDSI are becoming lower then LSI, due to the difference of corresponding values of K (Stiff and Davis) and A and B (Langelier). So far field results from seawater RO systems confirm lower tendency of calcium carbonate scaling as compared to brackish RO units.

The saturation relation of calcium sulfate is simpler to calculate than the one for calcium carbonate as its solubility depends only on concentrations of calcium and sulfate ions, temperature and ionic strength. The effect of ionic strength on solubility of calcium sulfate is quite significant as shown in the following example of calculations of saturation indexes (SI) for brackish and seawater RO systems:

Example 19

For $CaSO_4$; $K_{sp} = (\text{temp}/25)^{0.152} \times 1.8 \times 10^{-3} \times IS^{0.75}$

Brackish water system
R = 75%, TDS of concentrate = 4000 ppm, IS = 0.07
Ca = 900 ppm, SO_4 = 2400 (concentrations in concentrate)
$K_{sp} = 2.5 \times 10^{-4}$
SI = ([900/40000][2400/96000])/2.5×10^{-4} = 2.25

Seawater system
R = 50%, TDS of concentrate = 80,000 ppm, IS = 1.60

Ca = 900 ppm, SO_4 = 6000 ppm (concentrations in concentrate)
$K_{sp} = 2.6 \times 10^{-3}$
$SI = ([900/40000][6000/96000])/2.6 \times 10^{-3} = 0.54$

In the example above the product of calcium and sulfate ions concentrations in solution is much higher in concentrate stream of the seawater system as compared to the brackish system. However, due to the differences of ionic strength, the calcium sulfate is above saturation level in the brackish system and below saturation in the seawater system.

Another constituent that sometimes can present problem in RO applications is silica. In seawater sources the silica concentration is very low, just few ppm. However, in brackish water silica sometimes is present in significant concentrations. Silica can be present both in colloidal and reactive (soluble) form (61). In the past, the safe limit of silica in concentrate was considered as being about 140–170 ppm (as SiO_2). In the last decade a new scale inhibitors were introduced that are effective in maintaining much higher concentration of silica in solution. Some suppliers of these specialty inhibitors claim safe limits for silica concentration as high as 300 ppm. When treating brackish sources with significant silica concentration an extreme caution should be exercised with maintaining the designed recovery rate as silica scale is very difficult to remove.

Another potential scaling constituents, barium, although has very low solubility as barium sulfate, is much less common then silica. Events of barium sulfate scaling in RO systems were seldom reported in the past. Another scaling component of concern is calcium phosphate. Its presence is almost uniquely associated with municipal or industrial effluents at various concentrations. In majority of RO system treating municipal effluents concentration of phosphates is low and does not result in membrane scaling. At locations where phosphates concentration is high it still can be managed by using scale inhibitors or acidification. In extreme cases high concentration of phosphates can negatively affect economic feasibility of wastewater reclamation process due to high cost of acid required for pH adjustment.

The K_{sp} values (at 25°C) of common salts that could form scale in brackish RO systems are listed in Table 10.2 according to reference 58.

The potential for formation of calcium sulfate scale and blocking of feed channels in the spiral wound element is demonstrated in the following example:

Example 20

[Ca] = 1000 ppm = 0.025 mol
$[SO_4]$ = 2400 ppm = 0.025 mol
$K_{sp} = 2.25 \times 10^{-4}$

Saturation [Ca] = $[SO_4] = (2.25 \times 10^{-4})$^1/2 = 0.015 mol

Excess [Ca] = $[SO_4]$ = 0.010 mol

MW $CaSO_4$ = 40,000 + 96,000 = 136,000mg/l

Excess $[CaSO_4]$ = 0.010 mol = 0.010 × 136,000 = 1,360 mg/l

RO system operates at 80% recovery rate. At average flux of 26l/m^2/h, 7 elements per vessel the concentrate flow in the last element is: 26 × 36.8 × 7 × (1 – 0.80)/0.80 = 1674 l/h

Potential $CaSO_4$ deposit = 1674 × 1.36 = 2276 g/h = 2.28 kg/h

Assuming specific density of $CaSO_4$ = 4g/cm^3, volume of excess $CaSO_4$ that could precipitate is RO element is (2.28 × 1000)/4 = 570 cm^3/h.

Free volume of feed channels in SW element is:

100 cm × 100 cm × 0.075 cm × 20 = 15000 cm^3 (about 50%)

The above calculations shows that at the saturation conditions scale could lead in a short time to complete blockage of tail elements.

Potential for scale formation is extremely important issue in brackish applications. In some locations it may determine the maximum recovery rate for brackish RO systems. Due to variability of water compositions in brackish applications and limited level of understanding of the relevant salt solutions systems at saturation in RO conditions, it is quite difficult to make accurate predictions about scaling. RO industry adopted limits for individual salts based literature data and some field experience (Table 10.2). Due to lack of accurate analytical models, developed for RO applications, these limits include significant safety margins.

Manufacturers and suppliers of scale inhibitors are continuously introducing new scale inhibitors that enable operation at higher levels of concentrations then those initially proposed by membrane manufacturers. So far the experience

TABLE 10.2

Ksp of scale forming common to RO

Compound	Formula	K_{sp} (ppm)
Calcium sulfate	$CaSO_4$	2.5×10^{-5}
Barium sulfate	$BaSO_4$	2.0×10^{-10}
Reactive silica	H_4SiO_4	(120 – 160)
Calcium carbonate	$CaCO_3$	LSI < 0 S&DSI < 0
Calcium phosphate	Ca_3PO_4	(see comments in Chapter 17)

with the subsequent products introduced over the years was quite positive. Seldom any problems of scale formation could be related to malfunction of scale inhibitor if applied according to manufacturer specifications. Scale inhibitors prevent scale formation by retardation of nucleation process of scale forming crystals. The mechanism of prevention of crystal grow is either through threshold effect, crystal structure distortion, dispersion or sequestration. Dosing rate of scale inhibitor is determined by supplier of chemicals based on feed water composition and recovery rate. Feed water analysis should include information on concentration of iron. High concentration of iron containing compounds in concentrate may reduce effectiveness of some scale inhibitors. Required concentration of scale inhibitor seldom exceeds 10 ppm of active ingredient in the concentrate stream.

In seawater applications, due to typical seawater compositions and high ionic strength, potential for scaling is not a recovery limiting issue. In seawater systems recovery rate is limited by required operating pressure as discussed in the following chapter.

10.5. Particulate matter

To achieve high membrane packing density the RO membrane elements are configured with very narrow feed channels. In spiral wound elements the height of the feed channels is less than 0.8 mm (0.03″), filled with feed spacer net, which separates adjacent membrane surfaces and promotes turbulence. To avoid blockage of such narrow internal passage, RO feed water entering membrane elements, should have low concentration of particulate matter. The commonly accepted quality indicators of RO feed water in this respect include:

1. Turbidity
2. Suspended solids concentration
3. Silt density index (SDI)

Turbidity determination, usually expressed as nefelometric turbidity units (NTU) is determined through measurements of intensity of light scattered by suspended particles in water samples. Suspended solids concentration is determined by filtration of measured volume of water sample and weighting of dry residue on the filter. The SDI is determined through measuring the rate of filtration of water sample through the filter (Fig. 10.6). Determination of all three

Silt Density index : SDI = $100*(1 - t_0/t_{15})/15$

t_0	t_{15}	SDI (15 min)
18	22	1.2
20	33	2.6
26	73	4.3
46	50	0.5 (?)

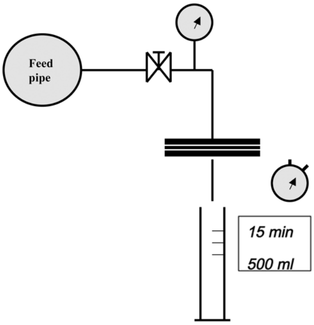


FIG. 10.6 Determination of silt density index.

above indicators is described in ASTM procedures (55, 54, 17). Out of above three indicators only turbidity can be measured continuously on line. The other two are conducted as discrete measurements on water samples taken periodically.

Another water quality parameter that is being used to monitor operation of pretreatment system is number of particles measured with particle counters. This could be applied as on line, semi-continuous measurement. Application of particle counters for RO application is still at very early stages. So far no relations had been reported as being established between particle counter measurements and fouling rate or performance stability of RO membranes.

The feed water indicator, most relied on in RO applications, is SDI. It is based on measurement of rate of declining flow, at a constant pressure, of a water sample through a porous filter membrane of nominal 0.45 μ porosity. The filter is placed in a simple circular holder (Fig. 10.6) and connected to feed water line at applied pressure of 2 bar (30 psi). The time measured for filtration of a constant volume (500 ml) at the beginning of the test (t_1) and after 15 minutes (t_2) is used to calculate the SDI according to the following equation:

$$\mathrm{SDI} = 100\%(1 - t_1/t_2)/15 \tag{53}$$

If the filter plugs to fast for meaningful determination of the filtration time, the volume of filtrate being collected or time between measurements can be decreased. As shown in Fig. 10.6 (last entry in the table) it is possible to have long filtration times and still calculate low SDI values. If the reading for filtration time t_1 significant exceeds 30 seconds then most likely there is a problem with equipment or test conditions. The SDI method is very sensitive to concentration of foulants but it is not very accurate. No meaningful correlation has been established between values of SDI and turbidity.

Attempts to improve accuracy of this method led to introduction of Modified Fouling Index (MFI). The test for MFI is based on measurement of pressure increase required for maintaining of constant filtration rate through well defined membrane filter (28). The MFI results are more reproducible than SDI but the test is difficult to perform manually and automatic equipment is necessary (18) for a routine determination in the field conditions.

According to the Eq. 53, the maximum value of SDI (for 15 min measuring interval) can be only 6.67. The majority of membrane manufacturers specify upper limit for feed water SDI as 5. However, field results show that for stable, long term performance of RO elements, the SDI of feed water should be consistently below the value of 4. Some limited research works (34) demonstrated that in respect of solids concentration the SDI scale is a geometric one. Therefore, for water having SDI = 3 and another water having SDI = 5 the corresponding suspended solids concentration difference is about four times higher.

The following Figs. shows the SEM pictures of clean filter (Fig. 10.7) and two filters (Figs. 10.8, 10.9) after being used for SDI determination of seawater feed. Fig. 10.6 shows SEM picture of filter pad after SDI determination that resulted in SDI value of 2.2. Fig. 10.9 shows corresponding SEM picture for SDI value of 4.8. The tick deposit on the filter, which was used to filtrate water sample with higher SDI, is clearly visible.

Field results have demonstrated that in majority of cases water from deep wells has very low SDI, usually less than 1. RO systems, operating with good quality well water feed, practically do not show any pressure drop increase across the membranes or flux decline. Surface water, after a conventional pretreatment, usually has SDI in the 2–4 range. RO system processing feed water with SDI in the 2–3 range shows stable membrane performance. Membrane cleaning frequency for such feed water does not exceed 1–2 per year. RO systems processing feed water of higher SDI, in the 3–4 range, usually suffer from some degree of membrane fouling and somewhat higher membrane cleaning frequency may be required. Long term operation of RO system with feed water having SDI above 4 is not recommended. As mentioned already, past attempts to correlate turbidity with SDI values were not successful. These two feed water quality indicators correlates to the number and size of suspended particles in a different way. However, usually the feed water with SDI in the 2–3 range has corresponding turbidity below 0.1 NTU, usually at 0.05 NTU range.

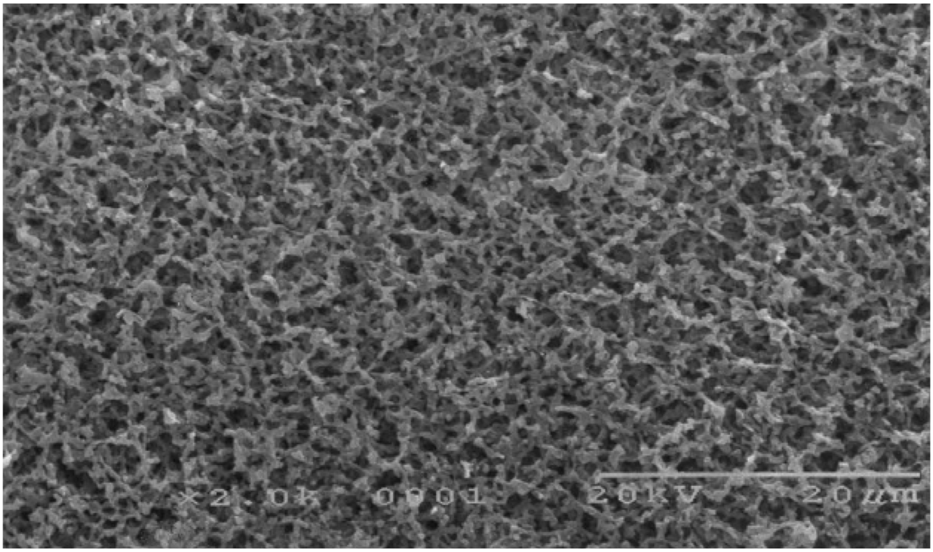


FIG. 10.7 SEM picture of a clean SDI filter pad. Magnification ×2000.

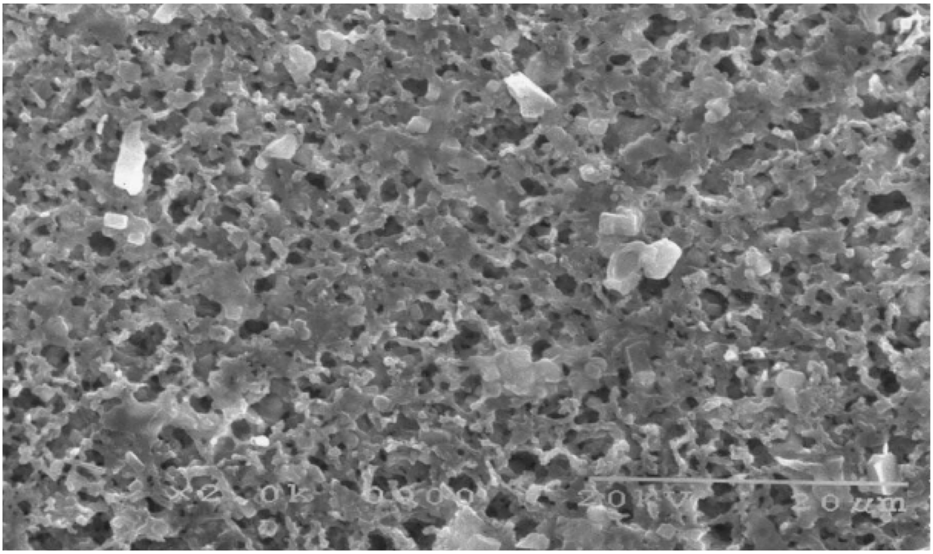


FIG. 10.8 SEM picture of a filter pad after SDI determination. SDI = 2.2. Magnification ×2000.

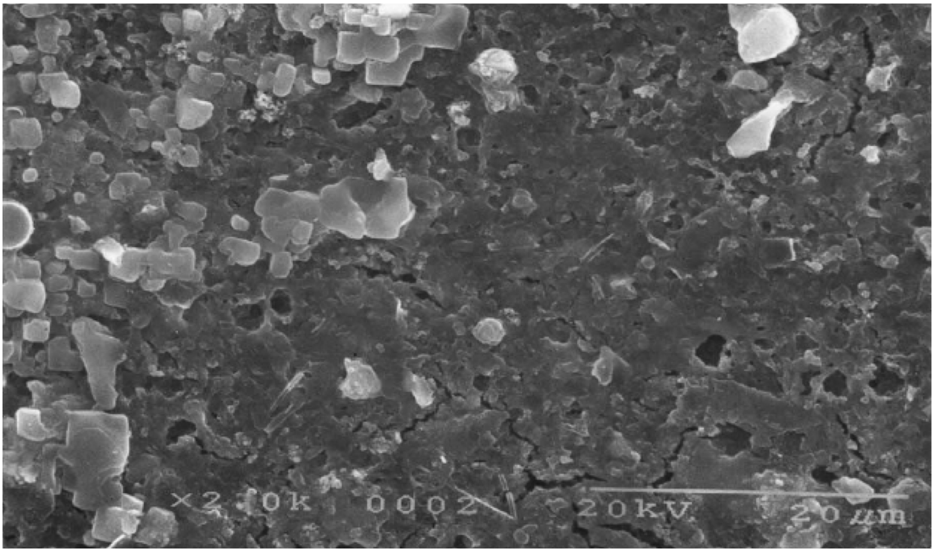


FIG. 10.9 SEM picture of a filter pad after SDI determination. SDI = 4.8. Magnification ×2000.

10.6. Organic matter

Organic matter in RO feed is customary expressed as a total organic carbon (TOC). Surface water, water from shallow wells and municipal secondary effluent usually contains some concentration of dissolved organics. In surface waters the organic matter originates from decomposition of humic substances. In shallow well water the presence of organics could be result of water infiltration through strata containing natural organic matter. In both cases the TOC concentration is usually below 3 ppm. This low concentration of TOC in the fed water has little effect on membrane permeability. In RO feed originated from secondary effluent, the TOC concentration fluctuates in the wide range between 2–10 ppm. The presence of high concentration of organic matter results in flux decline due to adsorption on the membrane surface. The initial rate of flux decline, due to organic adsorption is rapid, about 10–30% decrease from the initial permeability value. Then the permeability stabilizes, and the decline levels off. However, if colloidal matter is also present in the feed water, the permeability decline is usually more severe, due to formation of thick, mixed foulants layer of low water permeability.

Some potential water sources for RO processing are contaminated by oil and grease. Such conditions could exist in industrial wastewater streams or in seawater in the areas of heavy maritime traffic. Oil and grease have to be completely removed from the feed water prior entering the RO membrane elements. Low concentration of oil and grease will not result in structural damage of the membrane element but will cause severe decline of water permeability. Conventional method of removing low concentration of oil and grease includes air flotation and/or media filtration with flocculation using metal based flocculants.

10.7. Biofouling

The assumption is that every body of water contains microorganisms in equilibrium with the local nutrients supply. The water born microorganisms easily attached themselves to the surfaces in the RO systems and form colonies. The attachment to the surfaces is through excretion of extra cellular polymeric substance (EPS), composed mainly of polysaccharides. On wetted surfaces, the microbiological cells and surrounding EPS form biofilm, that can grow at rapid rate if sufficient nutrients and energy are available. The structures of biofilms are not uniform, depending on type of microorganisms and environmental conditions: pH, temperature, flow velocity, age and variety of other parameters. The general knowledge on biofilms mechanism of formation and their control is still developing. More recent research approach, under umbrella of "biofilm ecology" utilizes genetic tests to monitor biofilms growth and study microorganisms diversity (37).

The phenomena of biofilms formation in RO systems (biofouling) has been recognized early and is well documented. The methods of effective prevention are continuously being developed. Majority of RO systems processing well water experience little or no biofouling. Mitigation of biofilm formations can be achieved through proper system configuration and operating conditions:

1. Elimination of stagnant areas in the pipe-network.
2. Direct hydraulic connection between well head(s) and RO system, without feed water storage tank.
3. Total prevention of exposure of feed water to light. All FRP piping and other plastic components of the system have to be completely opaque.
4. Avoiding addition of water treatment chemicals that can be metabolized by microorganisms.
5. Avoiding continuous chlorination of the feed water.

RO systems treating surface water (mainly seawater), are more susceptible to biofouling than brackish systems. In the past it was common procedure to chlorinate raw seawater through the pretreatment system to control bacterial population, and reduce free chlorine (using sodium bisulfite) prior to RO membranes. More recently it was established (35, 36) that chlorination/dechlorination approach actually has a potential of increasing bifoluling in the membrane unit. The current understanding is that large size molecules of humic substances, that are present in the surface waters are not digestible by microorganisms. However, during chlorination of the feed water, free chlorine breaks them into smaller fragments of assimilable organic carbon (AOC), that can be metabolized by microorganisms. The microorganisms that survived chlorination will grow rapidly in subsequent sections of the system, past the dechlorination point. As the polyamide membrane elements can not be exposed to free chlorine, past the point of addition of chlorine reducing agent, no biocide is present in the feed water. In systems where chlorination/dechlorination is practiced, membrane surfaces provide good conditions for development of biofilm.

Following the above understanding of conditions leading biofilm formation, the current approach is to avoid all together chlorination of the feed water or practice intermittent chlorination/dechlorination only (dosing rate range: less than 1–5 ppm, duration 0.5–few hours per week). This approach restores the balance between the microorganisms population and availability of nutrients and usually results in low level of biological activity. One of important conditions, contributing to lower fouling rate is selection of intake location that will provide clean water, with low concentration of foulants and low level of biological activity. Once biofilm has been formed in the RO system it is extremely difficult to restore it to clean conditions.

The general approach to mitigate microorganisms grow in RO system is to periodically change drastically the environmental conditions (extreme pH, wide change of salinity) and conduct extensive cleaning procedures. Biofouling demonstrates itself through increase of pressure drop in the RO unit, followed by general decline of membrane permeability (need to apply higher feed pressure). In initial stages the biofouling it is localized in the feed piping and lead elements.

RO treatment of municipal effluents represents unique situation of high biofouling potential of RO feed water and development of effective way of controlling microorganisms population. In the past, municipal effluents were processed using cellulose acetate membranes. Therefore, it was a common practice to chlorinate RO feed water to prevent microorganisms grow. Due to presence of ammonia and high concentration of organics in the effluent the free chlorine was converted to chloramines. Today, almost all RO systems treating municipal effluents use composite polyamide membranes. Based on field experience it was established that in wastewater applications salt rejection of composite membranes, in presence of chloramines concentration of 2–4 ppm, is sufficiently stable to provide 3–5 years effective membrane life. The above range of chloramines concentration is also sufficient to effectively mitigate membrane biofouling. The specifics of design and operating conditions of membrane based wastewater reclamation systems are discussed in Chapter 17, Wastewater treatment and reclamation by RO and NF process.

10.8. Permeate recovery ratio

Permeate recovery ratio, which is the rate of conversion of feed water to permeate affects both the economics and process of the RO/NF systems. The volume of raw water required for a given permeate capacity of the RO system is directly determined by the design recovery ratio. Therefore, the size of the raw water supply system, capacity of the pretreatment system, size of the high pressure pump, feed and concentrate manifold are all functions of the recovery ratio as well. On the operating side, permeate recovery rate affects average osmotic pressure and pressure drop. Therefore, recovery rate affects feed pressure and permeate salinity. Higher recovery results in lower feed and concentrate flow rates. For this reason with increased recovery the pressure drop decreases (Fig. 10.10) according to Eq. 54.

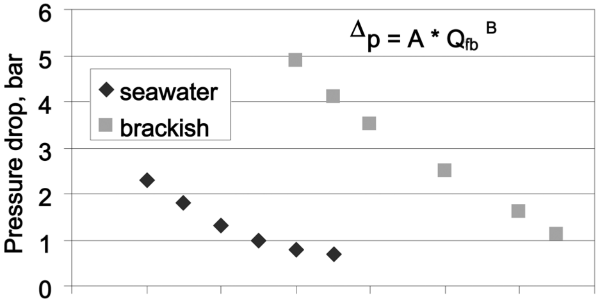


FIG. 10.10 Example of pressure drop in RO unit vs. recovery rate.

$$\Delta P = A \cdot Q_{fb}^{B} \tag{54}$$

Where ΔP is pressure drop, A and B are constants specific for element configuration, Q_{fb} is average of feed–concentrate flow

Because seawater RO systems operate in lower recovery range then brackish plants, and in seawater systems overall pressure drop is lower due to lower flow rate per vessel, the effect of recovery on pressure drop is more pronounced in brackish units. Recovery rate affects directly the increase of feed concentration along the RO unit, which in turn affects the average feed osmotic pressure (Fig. 2.4). Therefore, higher recovery will result in need of higher operating pressure. In brackish systems, depending on feed salinity, there could be a minimum of feed pressure with increased recovery (Fig. 10.11). The minimum of feed pressure value is result of system pressure drop decrease, which, at sufficiently low feed salinity could be more pronounced that the parallel increase of the osmotic pressure. However, as recovery increases, eventually the osmotic pressure increase will be more significant then the decrease of pressure drop.

In seawater systems the conditions are somewhat different. There, due to high feed salinity, increase of recovery rate is translated directly to increase of required feed pressure (Fig. 10.12). The increase of average feed salinity with recovery rate affects directly permeate salinity both in brackish and seawater systems (Figs. 10.13, 10.14). Due to process economics and limited availability of brackish feed, there is tendency to design brackish water systems for a highest possible recovery rate. Concentrate disposal is usually less expensive at higher recovery. However, in some brackish RO systems, where limitations on concentrate salinity are imposed, the recovery rate has to be adjusted accordingly.

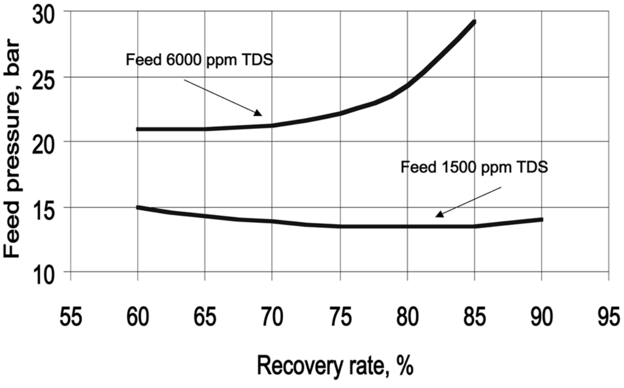


FIG. 10.11 Required feed pressure vs. recovery rate in a brackish RO unit.

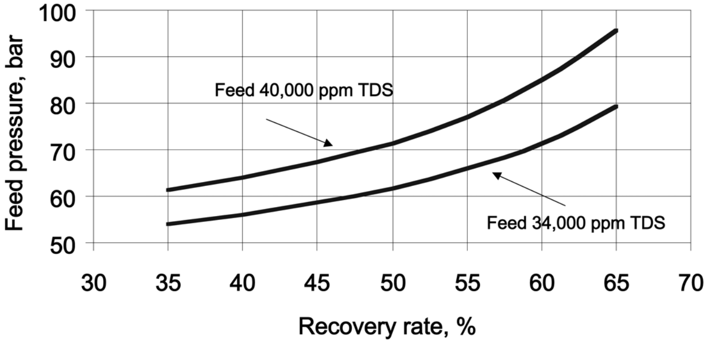


FIG. 10.12 Required feed pressure in vs. recovery rate in a seawater RO unit. Average permeate flux 14 l/m^2/h.

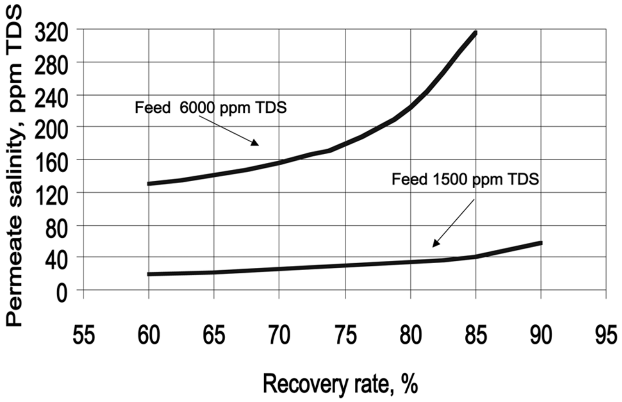


FIG. 10.13 Permeate salinity vs. recovery rate in a brackish RO unit. Average permeate flux 28 l/m^2/h.

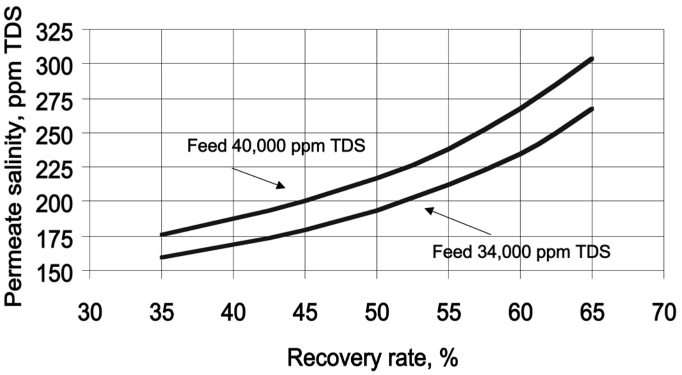


FIG. 10.14 Permeate salinity vs. recovery rate in a seawater RO unit. Average permeate flux 14 l/m^2/h.

In majority of cases, in brackish water systems the maximum recovery is limited by solubility of sparingly soluble salts. The scaling constituents that most frequently affect the recovery limits are calcium sulfate and silica. Calcium carbonate, although has also limited solubility, its precipitation can be easily controlled by pH adjustment and/or use of scale inhibitors. Common range of brackish system recovery rate is 75–85%. RO systems operating with very low salinity feed are sometimes designed to operate at 90% recovery.

In seawater RO systems the recovery rate is determined by considerations of maximum feed pressure, operating cost and required permeate salinity. The common range of recovery rate is 35% for feed water of high salinity (≧45,000 ppm TDS) and up to 50% recovery at lower feed salinities (~35,000 ppm TDS). The general tendency is to design RO seawater systems for recovery rate that require feed pressure would not exceed 70 bar (~1000 psi). In the recent efforts to reduce operating cost through minimizing power consumption the recovery rate is being optimized according to feed salinity and temperature. The optimization of operating parameters will be described in the process design section (Chapter 11, RO/NF system design).

The array of pressure vessels in RO unit is affected by the total recovery rate. In brackish RO units the recovery rate is almost always above 60%, most likely in the 70–90% range. The common configuration of brackish unit is of two stages system for the system recovery rate up to 85%. Above 85% recovery rate a three stage system configurations are more likely. To correct the uneven permeate flux distribution along the three stage systems, inclusion of a feed booster pump between stage two and three is quite common. As mentioned already in chapter 5, it is customary to design RO units for a 2:1 ration of pressure vessels in subsequent stages in order to maintain similar average feed flow rate per vessel. As in seawater systems maximum recovery rate seldom exceeds 50%, therefore seawater units at present are mostly design in a single stage configuration. The number of elements per vessel, in a single stage design, is being recently increased from 7 to 8. The 8 elements per vessel configuration has an advantage of lower capital cost of RO unit, smaller footprint and better memebrane performance due to lower concentration polarization (153, 154).

10.9. Permeate flux rate

Permeate flux represents the water flow rate through a specified area of membrane surface. That is, part of the feed water passes the membrane leaving on the feed side rejected dissolved constituents and water born particles. The value of the average flux rate (the flux rates averaged over the entire system) is a very important consideration in designing a system. Depending on the feed water source type and how "clean" the water is, if the selected flux rate is too high, membrane surface fouling from the retained constituents becomes more likely. The design average permeate flux rate of the RO system uniquely defines number of membrane elements and pressure vessels required for a given system capacity. Value of the average flux rate also affects the operating parameters. Higher permeate flux rate results in lower permeate salinity and requires higher feed pressure.

Relations between permeate flux, permeate salinity and feed pressure are defined by Eqs. 2–6. As mentioned above, as water flows through the system, rejected dissolved constituents and water borne particles are retained in the concentrate stream. The concentration of the retained constituents near the membrane surface is higher than in the bulk of the feed stream. With increased concentration, formation of a fouling layer on the membrane surface may occur. This excess of concentration at the membrane surface depends on the flux rate and the concentration of the rejected constituents in the feed water.

The constituents, which are of special concern in respect of potential membrane fouling, are particles, biofragments and dissolved organics. The recommended range of average flux rate for an RO system is defined according to the quality of the feed water. Feed water quality is defined in terms of the Silt Density Index, turbidity, TOC and suspended solids. Because in most cases, prior to construction of the desalination system feed water quality parameters are not available, feed water is qualified in terms of water source. From the conventional surface sources, one can expect that the raw water could have a high concentration of potential foulants and that their concentration may experience significant seasonal variations.

Feed water from deep wells is usually of good and consistent quality. It can be assumed that RO permeate, used as a feed in two pass systems, does not has any significant fouling tendency. The corresponding design range of the average flux rate for low salinity applications is 17–23 $l/m^2/h$ (10–14 GFD) for surface water, 23–30 $l/m^2/h$ (14–18 GFD) for well water and 31–43 $l/m^2/h$ (18–25 GFD) for treating RO permeate. When designing second pass systems, treating RO permeate, for operation at a very high flux rate, attention should be paid not to exceed maximum feed flow recommended for a given element type. Feed water to seawater systems originates either from an open intake or shallow beach wells. The range of design flux rate for seawater systems is 11.0–15.0 $l/m^2/h$ (6.5–9.0 GFD) for feed water from an open intake and 13.6–19.0 l/m^2/h (8–11 gfd) for a beach well source. In applications of RO membranes treating municipal effluents following membrane pretreatment, stable performance can be maintained at flux rate of 17–21 l/m^2/h (10–14 gfd).

The above values represent an average permeate flux rates, used for RO system design. In the actual system the permeate flux varies considerably along the system. The permeate flux rate is high in the lead elements and decreases gradually in the direction of feed-concentrate flow. This is due to increase of osmotic pressure and pressure drop and corresponding decrease of NDP (see Eq. 8). An example of feed and osmotic pressures profile in a two stage brackish RO unit is shown in Fig. 10.15. The values of pressures for the feed and concentrate points were calculated for an RO unit treating feed water of 2000 ppm TDS salinity at recovery rate of 85%. For illustration simplicity the change of osmotic pressure and decrease of feed pressure is depicted as straight lines. As shown in Fig. 10.15, feed pressure at the entrance to the RO membranes has to be sufficiently high to compensate for osmotic pressure of the concentrate and pressure drop along the RO unit. Approximately, only about 50% of the applied pressure is available as a driving force of the desalination process.

The specific shape of pressures profile in the RO unit will depend on feed salinity, temperature and recovery rate and permeate flux rate. Corresponding conditions for a single stage seawater unit, processing 40,000 ppm TDS feed at 50% recovery, are shown in Fig. 10.16. It is evident that elements in the lead position, being exposed to higher NDP, will operate at higher flux rate as compared to elements that follows. The difference of the average permeate flux,

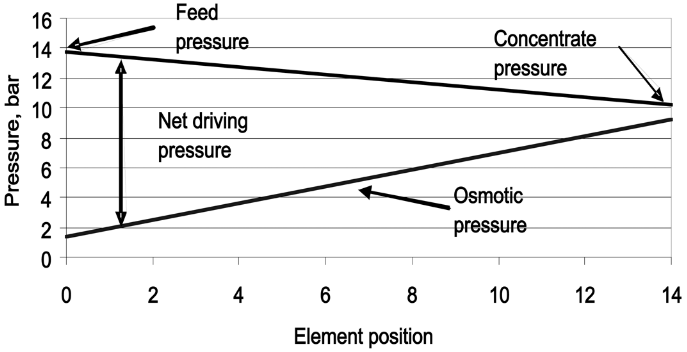


FIG. 10.15 Approximate osmotic pressures and feed pressure distribution along a two stage brackish RO unit. Feed salinity 2000 ppm TDS, 85% recovery rate.

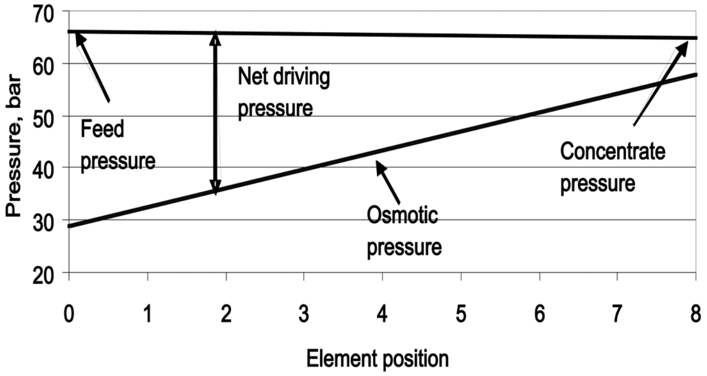


FIG. 10.16 Approximate osmotic pressures and feed pressure distribution along a single stage seawater RO unit. Feed salinity 40,000 ppm TDS, recovery rate 50%.

from the first and last element, could be quite significant. This flux rate difference will be higher for membrane elements with higher specific permeability. Common solutions to equilibrate flux rate along the RO unit are utilization of interstage booster (Fig. 5.3) or combination of low and high permeability membrane elements on subsequent concentrate stages (see information on hybrid configuration in chapter 16).

Table 10.3 lists nominal performance parameters of three common types of commercial brackish RO membranes: regular, high permeability and ultra high permeability. As examples, representative performance of CPA3, ESPA2 and ESPA4 are used. The above elements performance were used to calculate feed pressure, flux distribution and power consumption for a RO unit treating 2000 ppm TDS feed at recovery rate of 85%. The results are summarized in Table

TABLE 10.3
Representative brackish RO membranes

Type	Model	Nominal permeate flow, m^3/d (gpd)	Nominal salt rejection, %	Element membrane area, m^2 (ft^2)	Permeability, $l/m^2/h/bar$ (gfd/psi)
Regular	CPA3	41.6 (11000)	99.7	36.8 (400)	3.31 (0.133)
High permeability	ESPA2	9000	99.6	36.8 (400)	4.47 (0.18)
Ultra high permeability	ESPA4	12000	99.2	36.8 (400)	8.28 (0.33)

10.4. Comparison of results, calculated for system configurations with single type of membranes of increased permeability, indicate that use of higher permeability membranes will result in lower operating pressure and power requirements. However, at the same time the flux disparity between membrane elements in lead and tail positions will increase and the average flux of elements in the lead position will be increasingly higher.

High water permeability of membrane is usually associated with high rate of salt transport. Example of such conditions is shown in example of ESPA4/ESPA4 configuration.

Use of interstage booster pump, as shown in the ESPA4/ESPA4 + (B) example, results in more even flux distribution and lower permeate salinity. The interesting aspect of this particular configuration is that use of interstage booster with very high permeability membranes results in energy consumption somewhat lower than for configuration with a single process pump. For the cases listed in Table 10.4 of a very high permeability ESPA4 membranes this difference is 0.11 kWhr/m^3 (0.42 kWhr/kgal) or about 7%. For system utilizing lower permeability membranes and therefore requiring higher feed pressure this small

TABLE 10.4

Performance of brackish membranes of various permeability

Unit permeate capacity 10,000 m^3/d (2.64 mgd), recovery rate 85%, two stage unit, pressure vessel array 40:18 (7M PV), average permeate flux 27.6 l/m^2/h (16.2 gfd), feed salinity 2000 ppm TDS, temperature 25°C

Element type: 1st/ 2nd stage	Feed pressure, bar (psi)	1st stage concentrate, bar (psi)	2nd stage concentrate, bar (psi)	Permeate salinity ppm TDS	Lead element flux, l/m^2/h (gfd)	Tail element flux, l/m^2/h (gfd)	Energy consumption, kwhr/m^3 (kwhr/kgal)
CPA3/ CPA3	16.1 (233)	14.2 (206)	12.7 (184)	56	38.1 (22.4)	9.5 (5.6)	0.67 (2.54)
ESPA2/ ESPA2	13.9 (202)	12.1 (175)	10.8 (157)	63	41.6 (24.5)	5.9 (3.5)	0.58 (2.19)
ESPA4/ ESPA4	10.6 (154)	9.1 (132)	8.1 (117)	277	56.2 (33.1)	0.3 (0.2)	0.44 (1.66)
CPA3/ ESPA2	15.4 (223)	13.5 (196)	12.0 (174)	55	36.5 (21.5)	10.0 (5.9)	0.64 (2.42)
CPA3/ ESPA4	13.9 (202)	11.8 (171)	10.0 (145)	130	32.3 (19.0)	9.2 (5.4)	0.58 (2.19)
ESPA2/ ESPA4	12.8 (186)	10.9 (158)	9.4 (136)	148	37.9 (27.3)	9.4 (5.5)	0.53 (2.00)
ESPA4/ ESPA4+(B)	8.1 (117)	6.2 (90)	9.6 (139)	188	40.4 (23.8)	7.5 (4.4)	0.41 (1.55)

(B)–Interstage booster pump–5 bar (73 psi) energy benefit associated with interstage booster in a multistage system configuration declines. In seawater systems the single stage configuration is always more energy efficient than corresponding configuration of two stages or two stages with interstage booster. This is due to a lower pressure drop across the single stage as compared to two stage system operating at the same average permeate flux rate.

Results included in Table 10.4 shows that hybrid design, combination of two membrane types of different permeability, provides more even flux distribution, good permeate salinity and intermediate power consumption requirement. Availability of high permeability membranes with acceptable salt rejection makes this solution a convenient and cost effective design alterative.

Apart from considerations of energy requirement and permeate salinity, it is not clear at this point if conditions of large difference of permeate flux between load and tail elements and high permeate flux of elements in lead position are detrimental to long term membrane performance stability. It has been documented in field operation that RO system designed at high permeate flux rate have high fouling rates. This is especially the case for system treating water containing high concentration of dissolved organics. However, if RO system is designed for an average flux rate, considered adequate for a given type of feed water, excessive flux of the lead element, may not necessary result in accelerated membrane fouling. The lead elements that operate at high permeate flux rate, treat water with relatively low concentration of fouling constituents compared to the elements in tail position. Low concentration of foulant and high feed flow rate in lead elements could mitigate fouling process even at high flux rates. It can be also assumed that the RO system is self regulating in respect of surface absorption fouling. If as a result of fouling the permeability of elements in lead position declines, the system is converting itself to a hybrid configuration of low permeability elements followed by higher permeability and flux distribution should improve, effectively reducing fouling rate. So far little information from field operation is available to confirm the above assumptions and probably will be safer for a designer to follow the current approach and configure brackish RO system for a more uniform permeate flux distribution.

10.10. Membrane age

Performance of membrane elements changes with operating time. Usually permeability declines and salty passage increases. The membrane barrier, made of aromatic polyamide is very robust. However, formation of fouling layer on the membrane surface, abrasive effect of particles in the feed water and exposure to extreme pH cleaning chemicals will eventually change property of membrane surface and result in performance deterioration. These expected changes of performance are accounted for in calculation of projected performance of RO system by assuming annual increase of salt passage and decrease of permeability. The approach varies among different membrane manufacturers but it is generally assumed that salt passage could increase up to 10% per year and permeability irreversibly decline by 7%. Fig. 10.17 shows changes of membrane performance according to Eqs. 55 and 56. The assumptions are that the salt passage increase is linear and permeability decline is a compound function of operating time.

$$SP_y = SP_0\,(1 + SPI \times Y) \tag{55}$$

$$Perm_y = Perm_0\,(1 - Decl)^Y \tag{56}$$

where SP_y is salt passage at the year Y, SP_0 is the initial salt passage and SPI is salt passage increase factor, expressed as a decimal fraction. For the permeability (Perm) the same designation applies.

Accordingly for 3rd year of operation at 10% and 7% increase:

$$SP_3 = SP_0\,(1 + 0.1 \times 3) = SP_0 \times 1.30$$

$$Perm_3 = Perm_0\,(1 - 0.07)^3 = Perm_0 \times 0.80$$

The useful membrane life is determined by changes of membrane performance and required quality of product water and operating feed pressure. The

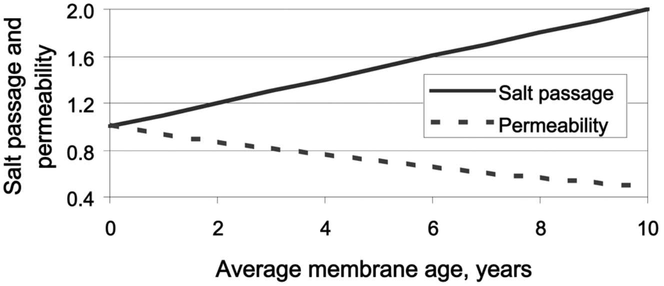


FIG. 10.17 Projected changes of membrane performance (salt passage and water permeability) with operating time.

usual approach is to budget in the operating cost an annual replacement of some fraction of the number of membrane elements in the system. However, for practical reasons the actual replacement is conducted by replacing all membrane elements in a train or desalting stage when system performance deteriorates below the acceptable level in respect of permeate salinity or feed pressure. Membrane replacement is usually conducted after attempts to restore membrane performance through cleaning did not produce satisfactory results. In order to reduce membrane replacement component of the operating cost, the old elements removed from the system are tested individually. Those with satisfactory performance are used in the next replacement operation.

For projections of long term performance an estimation of average membrane life is necessary. The estimation is based on assumed (or warranted) membrane replacement rate. An example of calculation of average membrane age is given in Table 10.5.

In the above table the average age is calculated based on setting the membrane replacement event at the beginning of the given year of plant operation. If, for example, the replacement would be conducted at the year end, starting at the end of first year, then the average membrane age at the end of 5th year would be higher, about 3.75 years. As will be discussed in the following chapters, the average membrane age is used as a parameter for calculation of membrane performance, which eventually is the basis for the terms of system performance warranty.

If the attempts to restore system performance by applying membrane cleaning are not sufficiently successful, the only other alternative is to correct performance through membrane replacement. The fraction of membrane elements

TABLE 10.5
Average membrane age table for replacement rate of 15% per year

Year of operation	0	1	2	3	4	5	Average age
0	100%						0
1	15%	85%					0.85
2	15%	15%	70%				1.55
3	15%	15%	15%	55%			2.10
4	15%	15%	15%	15%	40%		2.50
5	15%	15%	15%	15%	15%	25%	2.75

that have to be replaced will depend on membrane conditions and required level of system performance.

Table 10.6 demonstrates effect of element replacement on permeate salinity. In this example the system with new elements produced initially permeate salinity of 240 ppm TDS. The upper limit of allowed permeate salinity is 400 ppm TDS and at the current conditions of the elements in the system the permeate salinity is 500 ppm TDS. The fraction of the elements to be replaced depends on the target permeate salinity after replacement. Under assumption that new elements will have the same nominal salt rejection as the original load, to restore system to the ability to produce permeate of the original salinity (240 ppm TDS) all elements in the system will have to be replaced. To bring system in compliance with required permeate salinity of 400 ppm TDS a 37% of elements will have to be replaced.

Table 10.6 demonstrates two options for productivity restoration. In this example the system was design to maintain rated capacity at 80% of nominal elements productivity. Due to excessive fouling the current productivity is only 70%, example A (or 60%, example B). To recover lost productivity two alternatives are possible: some of the old elements can be replaced with new ones or new elements can be added to the system.

The results listed in Table 10.7 demonstrate that it is more cost effective, for productivity increase, to add new elements to the system rather then replace some of the old elements. However, with the addition of elements, the mem-

TABLE 10.6
Membrane replacement for permeate salinity correction.

RO system, Mediterranean seawater 40,000 ppm TDS, temp 29°C, recovery 45%, flux 13.8 lmh, average feed salinity 56,350 ppm TDS, RO elements with nominal 99.7% salt rejection. Initial permeate salinity: 240 ppm. Design salinity limit: 400 ppm. Current permeate salinity: 500 ppm .

Target permeate salinity, ppm TDS	Actual salt passage, %	Nominal salt passage, %	Elements to be replaced,%
240	0.43	0.30	100
300	0.52	0.38	80
350	0.62	0.44	58
400	0.71	0.50	37
500	0.88	0.63	—

TABLE 10.7
Permeate flow restoration through membrane replacement or addition

Flux restoration mode	Current capacity, %, examples A & B	Target capacity, %	Elements to be replaced or added,%
Elements replacement (A)	70	80	33
Elements replacement (B)	60	80	50
Elements addition (A)	70	80	10
Elements addition (B)	60	80	20

brane area in operation will increase. This will results in increased permeate salinity. The salinity increase will be in proportion to increase of membrane area. Therefore, the option of adding elements can only be implemented if a sufficient permeate quality margin exists.

11

RO/NF system design

11.1. System design guidelines

The RO system configuration and operating parameters have to conform to good engineering practice of RO system design. These parameters include: system average permeate flux rate, flow range per vessel of feed and concentrate, saturation values of scaling constituents in the concentrate stream.

The value of design system average permeate flux depends on the type of feed water to be processed by the system. The representative design average flux ranges and related parameters are listed in Table 11.1.

TABLE 11.1
Representative design parameters.

Feed water source	Representative SDI	Average system flux range, l/m^2/h (gfd)	Flux decline coefficient,%
RO permeate	<1	31–42 (18–25)	<5
Well water, brackish	<2	24–30 (14–18)	5–7
Surface water, brackish (1)	2–4	17–21 (10–12)	7–10
Seawater well (1)	1–3	13.6–19 (8–11)	7–10
Seawater surface intake (1)	2–4	12–15 (7–9)	7–10
Seawater surface intake (2)	1–3	13.6–19 (8–11)	7–10
Secondary effluent (2)	2–4	17–24 (10–14)	10–15

(1) Conventional pretreatment

(2) Membrane pretreatment

The design average flux rate defines number of membrane elements in the system, which in turn (Eq. 17), together with the recovery rate determines array of pressure vessels. In seawater systems the recovery rate is determined based on optimum feed pressure. In brackish systems the tendency is to maximize recovery rate according to saturation level of scaling compounds in the concentrate stream. In presence of generic scale formation inhibitors the following saturation values in the concentrate are considered to provide safe operating conditions:

TABLE 11.2
Conservative values of saturation limits used in RO industry

Compound	Saturation level,%
$CaSO_4$	230
$BaSO_4$	60,000
$SrSO_4$	800
SiO_2	100% (or 140–170 ppm)
$CaCO_3$	LSI <1.8, SDSI <1.0

As mentioned already, some specialty scale inhibitors enable to maintain much higher levels of saturation, therefore providing possibility of operation at increased recovery rate. Another two important design parameters that have to be considered are maximum feed flow and minimum concentrate flow rate per vessel. The limit on maximum feed flow is related to limit on pressure drop, increasing with increasing flow rate. Excessive pressure drop may result in telescoping of membrane elements that will lead to mechanical damage. At average operating conditions the pressure drop is 0.15–0.25 bar (2–4 psi) per element. 1–2 bar (15–30 psi) per stage.

The requirement of maintaining minimum concentrate flow is related to prevention of hydraulic conditions that would lead to high concentration polarization at low flow rates in the tail element. Avoidance of low concentrate flow rate is important in brackish system, where composition of concentrate could be close to saturation. In seawater systems low concentrate flow will result in higher permeate salinity and lower apparent permeability. The limiting values of fed and concentrate flow will depend on element configuration and thickness of the feed channel spacer. The common ranges of limiting flow values for 200 mm (8″) diameter, 1 m (40″) long elements are:

Maximum feed flow, m^3/h-vessel (gpm/vessel): 14–18 (60–80)

Minimum concentrate flow, m^3/h-vessel (gpm/vessel): 2.5–3.0 (11–13)

11.2. The design process of RO/NF system

The approach to system design will vary according to the role in the project of the party involved. There will be different design approach by project developer, engineering firm, construction company (OEM), membrane manufacturer or the end user. For large project it is quite common that a consortium is formed to compete for the desalination project. If more than one group is directly involved in the design process then the design responsibilities are defined by conceptual division of the system into a separate process areas. For example "pretreatment island" segment could include design work related to intake and all pretreatment process. The "desalination island" segment could include process starting from high pressure pumps and end up at permeate storage. The project design process will depend on the final objectives and on the project stage. For example, different level of details will be required during the bidding stage and during system construction. The following section provides outline of the design process as it is usually approached by an OEM at the bidding stages of the project.

Every real project starts with system specifications, which include general information about the desalination plant and criteria for bids evaluations. The evaluation criteria may vary from project to project. The evaluation can be based mainly on system cost in a "turn key" type project.

In a "built, own and operate" (BOO) or "built, own, operate and transfer" (BOOT) the operating cost is usually the predominant factor, affecting strongly the net present value. Project specifications include specific requirements regarding system permeate capacity and permeate quality. They will also specify conditions of payments, system acceptance conditions, duration of acceptance test and remedies in case of equipment or performance deficiencies.

The design process is outlined schematically in Fig. 11.1. The input information for system design includes project requirements, water source data and cost related information. The information, related to process economics, may include cost of major equipment and consumables, interest rate, cost of permitting and others. For some projects a pilot unit operation may be necessary to establish or confirm some of the process parameters. The design work will include developing of system process parameters based on project requirements, raw water information and membrane performance projections. Next, the process and instrumentation diagram (P&ID) or process flow diagram (PFD) will be developed. Based on P&ID or PFD a bill of materials (BOM), including major equipment is prepared. Bill of materials (BOM) is used to make preliminary

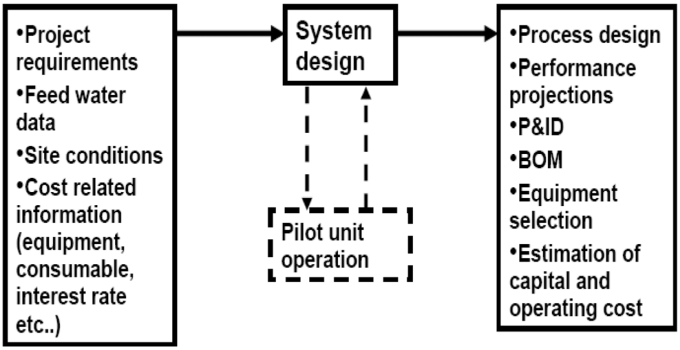


FIG. 11.1 RO system design process.

selection of major equipment and estimate equipment cost. Project cost will include direct cost (equipment) and indirect cost (engineering, permitting etc.). Operating cost is calculated based on process design, cost of consumables and labor. Detailed discussion of RO desalination project budgeting is included in chapter 18.

11.2.1.Site and feed water supply

In projects involving brackish water systems, development of raw water supply is usually conducted by specialized contractor and water analysis is provided to OEM‘s biding for the RO project. It is quite common however, that for the new sites the composition of raw water, listed in project specifications, represents estimation only, usually based on drilling and sampling of observation wells or limited pumping of the production wells, with a safety margin added. If the RO system is to be built on an existing site, historical information on composition of designated feed water usually exists. In projects involving NF systems, low salinity water, designated as a NF feed, is usually being already used for some other purpose and its composition is known.

Disposing of concentrate from brackish desalination plants is always a problem. However, in majority of cases, permitting for concentrate discharge is responsibility of the end user. In seawater projects it is quite common that scope of the project will include requirement, from parties competing for the project, for arranging of a site for the desalination plant and development of seawater delivery system, including concentrate discharge.

For small to medium size RO systems beach wells or Raney collector (also called sub-surface intake) could be evaluated as a viable option of clean seawater supply. Possibility of using beach wells or Raney collectors depends on water transmission of the local soil and possibility for obtaining permit for constructing such a structure adjacent to the beach area. At favorable conditions, sub-surface intake could provide seawater of much better quality than available from surface intake (42). For large capacity system seawater delivery project frequently consists of design and construction of an open intake. As discussed already in chapter 8.2.1, the intake structure should be built at location of a consistently good water quality. Usually this requires depth of seawater of at least 10–15 m (32–49 ft). At some locations, of extended shallow water areas, this may translate to a very long raw water supply piping from intake to the on shore pump house. Logistics and cost of delivery such pipe segments to the plant site have to be considered during the design process.

For obvious reasons the desalination plant should be located very close to the shore line. With the continuous population growth, ocean front sites are scare commodity and obtaining permit for such a site could be quite difficult. One solution is location of desalination plant at an existing industrial site, which is close to the water line. Power plants are good candidates for desalting plant collocation (see Fig. 8.8). If power plant uses seawater cooled steam condenser, its water demand far exceeds any feed water requirement of an average size RO seawater plant. This cooling water is eventually discharged back to the ocean, at the same salinity but at temperature 3–10°C (5–18°F) higher then the water temperature at power plant inlet.

The concerns and benefits of using seawater discharged by power plant are discussed in chapter 8.2.1. One of the more important issues is screening. Power plant applies screens in seawater intake, mainly in response to local environmental regulations. Screening requirements of power plant could be significantly different than required by the RO equipment. In RO process, the specific requirements of screening of seawater influent are related to prevention entrance of particulate matters that could result in upsets of operation of the RO pretreatment system. Therefore, special screening may be required in addition to one employed by power plant. Overall, the economic and logistic benefits of using seawater, available on site as a power plant condenser cooling water outlet, usually out weights concerns about quality of raw water from such a source.

Land area requirements of desalination plants includes area for feed water reservoir, pretreatment system, RO building, post-treatment and product water storage reservoir, general storage warehouse and repair workshop, electric transformation station, laboratory, administration and utility areas. Stand alone RO seawater system usually will have an additional separate area for housing of the intake clear well and pumps, which, for practical reasons has to be located very short distance from the water line. This location also includes step down transformer and motor control center (MCC) for the intake pumps. To convey seawater from the intake pump house to the RO plant, an easement access has to be arranged.

In case of RO system treating surface water and using conventional pretreatment, the combined pretreatment system and RO building area requirements are about 35–40 m^2/1000 m^3/d (1400–1600 ft^2/mgd) permeate capacity. In case of membrane pretreatment the required area for a pretreatment unit can be reduced by about 25%. The over all area of the plant site is usually 3–5 times the combined area of pretreatment and RO building.

Figure 11.2 shows aerial picture of 330,000 m^3/d (87 mgd) RO seawater plant at Ashkelon, Israel, with indication of major plant components. The plant consists of two identical systems of 50% permeate capacity each. Seawater is provided from an open intake and the plant utilizes conventional (single stage media filtration) pretreatment. The site area, without product water reservoir, is about 70,000 m^2 (17 acres). The site includes, in addition to RO equipment, a small power generation facility to supply all electricity required for operation of the RO plant.

FIG. 11.2 Aerial view of a 330,000 m3/d (88 MGD) seawater RO plant at Ashkelon, Israel (courtesy of Israeli Desalination Engineering).

11.2.2.Selection of pretreatment process

Numerous examples of RO systems of all sizes, types and locations have demonstrated unique importance of good design and operation of the pretreatment system for successful, long term, performance of membrane elements. Selection of suitable pretreatment process is mainly driven (in addition to process economics) by the type and quality of raw water. Basic rules of pretreatment system design have been developed over time. However, proper design of pretreatment system requires good understanding of chemistry, biology and hydraulics of the water supply systems. Experience has shown that the minimum pretreatment, still providing required feed water quality, is the preferred solution. Excessive number of pretreatment steps may, in some cases, result in poorer feed water quality and accelerated membrane fouling. This last rule especially applies to brackish water systems treating well water. There is significant number of cases where continuous chlorination contributed to bacterial after-growth in RO unit or sand leaking from media filters, passed cartridge filters and accumulated in lead RO elements.

The pretreatment process is most effective when tailored to raw water quality. Table 11.3 summarized expected values of water quality parameters of raw water from brackish wells, surface sources and secondary effluent.

The pretreatment required for raw water of the quality listed above, will be a "standard one", i.e., accepted as a good engineering practice. Components of pretreatment systems for each three water types are listed in Table 11.4. At locations, where raw water is of very stable and better than average quality, designing a simplified pretreatment system may be considered. The simplified configuration will reduce system cost but will increase risk of higher fouling rate. On the other hand if, there are periods of poor water quality, capability of

TABLE 11.3
Expected water quality from well and surface intake

Quality parameter	Well water	Surface water	Secondary effluent
Turbidity, NTU	<1	<2	<2
SDI	<1	<5	not measurable
Suspended solids, ppm/ml	<5	<5	<20
TOC, ppm	<3	<5	<20
SiO_2	<25	n.a.	n.a.

TABLE 11.4
Summary of pretreatment options.

Raw water source	Simplified pretreatment	Good engineering practice	Potential future technology
Well water (dedicated wells)	Acid or scale inhibitor (SI), cartridge filters	Sand screen, scale inhibitor, cartridge filters	Cartridge filters replaced by backwashable screening devices
Surface water, good quality.		Intermittent chlorination (optional), coagulation, media filtration, acid or SI, cartridge filtration	Cartridge filters replaced by backwashable screening devices
Surface water, poor quality	Intermittent chlorination (optional), coagulation, media filtration, acid or SI, cartridge filtration	Intermittent chlorination (optional), coagulation, two stage media filtration or clarification + media filtration, acid or SI, cartridge filtration	Intermittent chlorination (optional), clarification, coagulation, membrane (MF or UF) filtration, acid or SI.
Wastewater	Secondary effluent, chloramination, coagulation, media filtration, acid or SI, cartridge filtration	Secondary effluent, chloramination, backwashable screens, coagulation, membrane (MF or UF) filtration, acid or SI, cartridge filtration	Primary effluent, MBR, chloramination, acid or SI, backwashable screens.

pretreatment system has to be enhanced accordingly. The pretreatment system has to be able to cope with raw water quality all the time the pretreatment effluent is processed by the RO membranes. Even short excursion of feed water quality my result in severe fouling of RO elements. Pretreatment of well water is the simplest one, consisting only of scale or acid dosing followed by cartridge filters (Fig. 11.3). This configuration has been proven in field operation as very adequate. Brackish systems operating on well water from dedicated wells, with this type of pretreatment have history of operating the same complement of original membrane load, without replacement, for over 10 years.

The cleaning frequency is low, not exceeding one cleaning every 2–5 years.

Some brackish wells have tendency to release small quantity of silt during startup. In such systems addition of screens or other sand removal devices to the

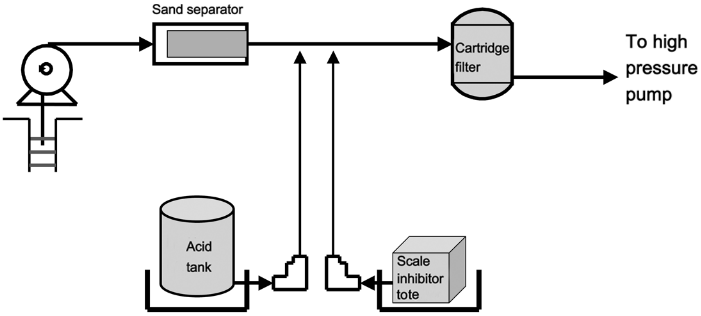


FIG. 11.3 Configuration of a conventional RO pretreatment system treating well water source.

pretreatment process is recommended. This above described pretreatment configuration is very reliable and cost effective. The future improvement pretreatment technology of well water feed may include replacing current cartridge micron filters with devices that are backwashable and provide more reliable separation barrier. The objective is to reduce, or even eliminate, cartridge replacement cost and provide barrier that is more resistant to abrasion of sharp particles. Such a device would also act as a sand separator.

Selection and quantity of chemicals used in brackish water systems is function of water composition and system recovery rate. Acid is used to reduce concentration of bicarbonate (converting bicarbonate to CO_2) and therefore reducing tendency of calcium carbonate scale formation. Without presence of scale inhibitor the target feed water pH is such that will result in slightly negative LSI in the concentrate. Calculation of required rate of acid dosing is a standard feature of computer projection programs provided by all membrane manufacturers. Quantity of acid, necessary to reduce feed pH to the required value can be calculated easily if raw water temperature, pH and bicarbonate concentration are known.

In presence of acid, bicarbonate is converted to CO_2 according to the following reaction:

$$HCO_3^- + H^+ = H_2O + CO_2 \tag{54}$$

$$K_1 = [H^+]\,[HCO_3^-]/[CO_2] \tag{55}$$

where K_1 is the equilibrium constant of the carbonic acid dissociation reaction. At 25°C the K_1 value is 4.2×10^{-7} (close approximation for low salinity solution. With increased salinity a correction for K_1 should be applied).

$$pK_1 = -\log(K_1) = 6.37$$

Accordingly:

$pH - pK_1 = \log([HCO_3^-]/[CO_2])$ (56)

$R = [HCO_3^-]/[CO_2]$ (57)

$R = 10^{\wedge(pH-6.37)}$ (58)

$CO_{2f} = HCO_{3f}/R_f$ (59)

For acidified feed $R_a = 10^{\wedge(pHa-6.37)}$ (60)

$HCO_{3a} = R_a\,(CO_{2f} + 0.88\ HCO_{3f})/(1 + 0.88R_a)$ (61)

$CO_{2a} = CO_{2f} + 0.88\,(HCO_{3f}-HCO_{3a})$ (62)

Acid quantity required for pH reduction from pH_f to pH_a

H_2SO_4 (ppm 93%) = $1.06\,(HCO_{3f}-HCO_{3a})$ (63)

HCl (ppm 30%) = $2.4\,(HCO_{3f}-HCO_{3a})$ (64)

where subscript "f" designates feed before acidification and subscript "a" designates values of acidified feed.

Example 21

Calculation of acid required for pH adjustment and concentration of CO_2 in acidified feed.

Feed pH prior to acidification $pH_f = 7.4$

Bicarbonate concentration HCO_3 (as $CaCO_3$) = 320 ppm

Required $pH_a = 6.5$

$R_f = 10^{\wedge(7.4-6.37)} = 10^{\wedge1.03} = 10.7$

$CO_{2f} = 320/10.7 = 29.9$ ppm

$R_a = 10^{\wedge(6.5-6.37)} = 10^{\wedge0.13} = 1.35$

$HCO_{3a} = 1.35(29.9 + 0.88 \times 320)/(1 + 0.88 \times 1.35) = 192.2$ ppm

$CO_{2a} = 29.9 + 0.88(320 - 192.2) = 142.3$ ppm

Dosing rates of H_2SO_4 or HCl

H_2SO_4 dosing rate = $1.06(320 - 192.2) = 135.5$ ppm

HCl dosing rate = $2.4(320 - 192.2) = 306.7$ ppm

In majority of systems sulfuric acid is used for pH reduction. It is used at commercial concentration (about 93%) without dilution. Hydrochloric acid, being more expensive, is applied less frequently. When scale inhibitor is used, LSI in concentrate can be higher, at 1.8–2.2 range (specific value provided by scale inhibitor supplier). Dosing rate of scale inhibitor depends on recovery rate. The target value is 8–10 ppm in the concentrate stream. Dosing rate of acid can be easily controlled by measuring pH of feed downstream of acid addition or pH of a concentrate. Dosing rate of scale inhibitor is more difficult to monitor. Some methods include installing "hot wire" sensors in scale inhibitor dosing line, level measurement in flow through container or tracing fluorescent additive included in scale inhibitor formulation. The micron cartridge filters used in RO pretreatment should be rated at 5–15μ nominal pore size. The preffered porosity rating is 5μ, which provides good protection against small paticles reaching membrane elements and entering feed channels. The design filtration flow is about 1 m^3/h per 25 cm long cartridge (4.5 gpm per 10″).

In the past, surface water RO system was synonymous with seawater RO. Presently, growing number of nanofiltration systems, treating low salinity surface water or well water under influence of surface water, are being built and operated, producing potable water. Regardless the big difference in water salinity between seawater and surface brackish water, the configurations of pretreatment systems are quite similar. The configurations of seawater and nanofiltration membrane elements are practically identical therefore they required feed water of a similar quality.

Representative configuration of pretreatment system treating surface water is shown in Fig. 11.4. Treatment of surface water starts with coarse screening with about 5–20 mm (0.2–0.8″) screen opening to prevent entrance of large objects that could block flow in the pretreatment system. Coarse screens are usually

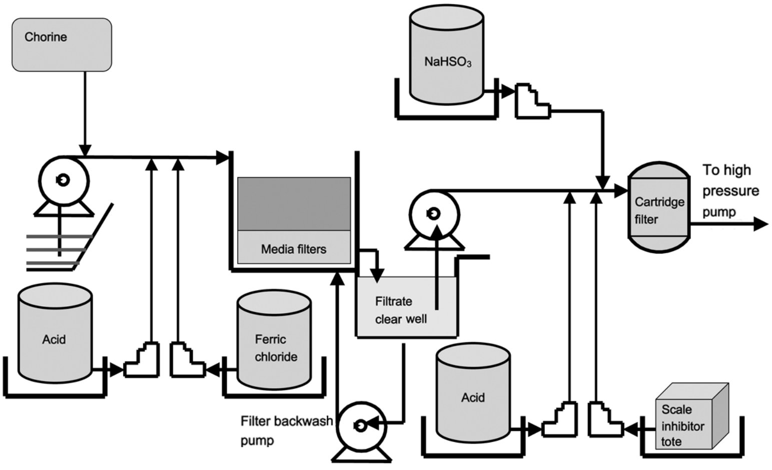


FIG. 11.4 Configuration of a conventional RO pretreatment system treating surface water source.

located at the intake inlet. The flow through screen opening should not exceed 0.2 m/s (0.6 ft/s).

At some locations coarse screens are followed by fine screens with openings of 2–4 mm (0.08–0.15″). The objective of fine screens operation is to block juvenile marine organisms from entering the pretreatment system. Intake structure is connected with on shore shaft through system of piping. Flow velocity in the intake piping should be sufficiently high, in the range of 1m/s (3ft/s), to prevent sediments to settle. Water from on shore shaft is pumped to the pretreatment system by intake pumps. The design value of hydrostatic head, to be developed by the intake pumps, is tailored to the distance of intake structure from the plant, elevation difference and type of pretreatment to be used. In large plants the tendency is to configure pretreatment system in such a way that water passage through media filter will be based on gravity flow from the filters inlet to the feed water clear well. In practice, this conditions can be accomplished only in a system that utilizes gravity filtration.

There are three basic alternatives of surface water pretreatment:

1. Conventional: single stage media filtration
2. Conventional: two stages media filtration or clarification followed by media filtration
3. Membrane pretreatment: stand alone MF/UF or including clarification or prefiltration step

Each option consists of similar functionality of treatment steps employed, as it is summarized in tables 11.5 through 11.7. The conventional, single media filtration step, includes chlorination as an optional feature for disinfection of water supply lines when necessary. The equipment used for chlorine dosing may include chlorine (gas) storage tanks and chlorine injecting equipment. In some systems, where addition of sodium hypochlorite solution is practiced, storage tanks and dosing pumps are used. The dosing rate 1–5 ppm is targeted to receive low residual 0.2–0.5 ppm of free chlorine at the entrance to cartridge filters. The coagulation process that follows, should start at such distance from the media filters that would provide sufficient time for flocculation, up to 20 min.

Coagulant dosing system consists of storage tank for ferric chloride or ferric sulfate solution, dosing pumps and in line mixing device (static mixer). The commercial strength of ferric salt solution is around 35%. Ferric coagulant could be also purchased and delivered in a solid form and solution prepared on site. In some RO plants, a polymer dosing unit is also included in the pretreatment system. Usually polymer is added, after coagulant flock is already formed, as a filter aid at low concentration of less than 1 ppm. Use of cationic organic polymer as a primary coagulants at high dosing rates, as it is sometimes practiced in potable water systems, is not recommended. In RO systems, if excess of filtration polymer is not retained by media filters, it can react with scale inhibitor forming fouling deposits or adsorb on membrane surface affecting its performance.

Media filtration system consists of pressure vessels or gravity filtration tanks, media filters, connecting piping, valves, filtrated water storage, filtrate backwash pumps and air blowers for air scouring. Filter backwash water can not be discharged directly, especially if metal coagulant has been used. Usually, backwash water (about 2–3% of filter effluent flow) is sent to clarifier to settle solids. Clarified effluent could be discharged or sent back to the head of the filtration system. Sludge from the clarifier is send to thickener and dewatering unit (using filter press) and eventually disposed as a landfill with the solids concentration of 20–50%.

Following media filtration, pH of feed water is adjusted using acid or scale inhibitor is added. This is to prevent scale formation in membrane elements. Chemicals dosing system consists of chemicals storage tanks, placed in containments, and positive displacement dosing pumps. Acid is usually delivered by tankers and transferred to on site storage tank(s). Scale inhibitor could be delivered in totes, which can be connected directly to suction manifold of dosing pumps. Alkalinity of brackish water is usually quite high. Therefore, dosing rate of acid is usually significantly higher than the dosing rate of scale inhibitor. Seawater has low alkalinity, less than 200 ppm and dosing rate of acid (if used at all) is only somewhat higher than the dosing rate of scale inhibitor.

If pretreatment system includes a chlorination unit, sodium bisulfite dosing unit for feed water dechlorination has to be included in system configuration as well. This dosing unit consists of equipment for preparation of dosing solution for dechlorination by dissolving solid sodium bisulfite in water, solution storage tank and dosing pumps. In some large systems, sodium bisulfite solution is prepared off site by a vendor and delivered to the plant.

Usually, the last treatment step of feed water prior to membrane elements, is micron filtration. In commercial RO systems micron cartridge filters (sometimes also called safety filters) are stainless steel vessels in horizontal or vertical configuration containing cartridge filters. Each cartridges housing vessel is equipped with large lid for loading and unloading of filtration cartridges. For the convenience of lid opening, the large capacity filter housings are usually of horizontal configuration. The filtration capacity of single filter housing could be up to a thousand m^3/h (4400 gpm) range and correspondingly contains hundreds of cartridges.

In RO systems that utilize pressure sand filters, the exit pressure of media filter effluent is often sufficient to pass water through cartridge filters. In systems that utilize clear well for feed water storage after media filtration, a transfer pump is used to create sufficient water head at the entrance to cartridge filters (Fig. 11.4).

The number of cartridge filter housings is in proportion to the number of high pressure pumps or number of RO trains in the system. The usual configuration is single cartridge filter housing per high pressure pump and membrane train. However, in some large systems cartridge filters and high pressure pumps form so called "pumping center" (56) and there the number of cartridge filters is determined based on capacity and cost considerations and not according to membrane trains–high pressure pumps symmetry.

Pretreatment with two stage media filtration or clarification followed by media filtration is only applied if raw water has high concentration of suspended solids. Locations that may require this more extensive treatment are area exposed to frequent and significant events of surface run off and/or periodic development of algae blooms. The two stage filtration process is more expensive and requires additional footprint. Therefore, it should be compared to another more expensive pretreatment alternative that includes membrane filtration in place of media filters.

The initial approach to application of membrane filtration was to replace media filtration equipment in a single stage filtration system with pressure driven or submersible membrane technology. It was realized recently that, in case of difficult water source, membrane filtration works more effectively if water is treated first with clarification, dissolved air flotation or rough media filtration. Due to ability of MF/UF membrane operation at increased filtrate flux rate, the cost and footprint of the combined membrane system could be competitive over the two stage media filtration system and usually will provide more consistent and better quality effluent.

11.2.3 Energy consumption of RO process

Energy consumption in RO unit is the sum of energy used in all segments of RO process: energy used for water supply (E_{ws}), energy used in pretreatment unit (E_{pt}), net energy spent by the high pressure feed pumping unit (E_{pu}), energy for permeate treatment and pumping (E_{pe}) and energy used by auxiliary equipment (E_{ax}) :

TABLE 11.5
Summary pretreatment configuration with single stage media filtration

Treatment step	Objectives	Equipment, materials	Process parameters
Intermittent chlorination (optional)	Mitigate bioactivity in supply lines	Chlorine injection as chlorine gas or NaOCl solution	Dosing rate 1.0–5 ppm, duration few hours per week
Coagulation	Destabilization of colloidal particles	Dosing of $FeCl_3$, $Fe_2(SO_4)_3$ or organic polymers	Ferric:1–30 ppm Polymer: 0.2–1 ppm, strong mixing, residence time 0.5–1 min
Flocculation	Agglomeration of colloidal particles	In line flocculation	Gentle mixing, residence time 1–20 min
Media filtration unit	Removal of colloidal particles through entrapment and blockage	Dual or multimedia filters, pressure or gravity	Filtration rates: 8.0– 15 $m^3/m^2/h$ (3.2–6.0 gpm/ft^2)
Media backwash unit	Removal entrapped particles from filters.	Backwash pumps and air scour blowers	Frequency every 8–24 h. Backwash rates 35–55 $m^3/m^2/h$ (15–23 gpm/ft^2) for ~10 min. Air scouring rates 55–90 m/hr (3–5 $scfm/ft^2$) for ~5 min
Acid or scale inhibitor dosing	Preventing scale formation in RO unit	Dosing system for scale inhibitor: pumps and tanks	Dosing rate, acid ~20 ppm, scale inhibitor 1–5 ppm
Sodium bisulfite dosing	Applied for reduction of residual free chlorine	Dosing system for sodium bisulfite: pumps and tanks	Dosing rate ~up to three times residual chlorine concentration
Cartridge filtration	Protecting pumps and membranes from particulate matter	50 cm (20″) long cartridges, rated at 5–15 µ nominal porosity	Filtration rate about 2 m^3/h (9.0 gpm) per 50 cm (20″) cartridge

TABLE 11.6
Summary pretreatment configuration with two stage media filtration

Treatment step	Objectives	Equipment, materials	Process parameters
Intermittent chlorination (optional)	Mitigate bioactivity in supply lines	Chlorine injection as chlorine gas or NaOCl solution	Dosing rate 1–5 ppm, duration few hours per week
Coagulation	Destabilization of colloidal particles	$FeCl_3$, $Fe_2(SO_4)_3$ or organic polymers	Ferric:1–30 ppm Polymer: 0.2–1 ppm, strong mixing, residence time 0.5–1 min
Flocculation	Agglomeration of colloidal particles	In line flocculation	Gentle mixing, residence time 10–20 min
Media filtration, Stage 1	Removal of coarse colloidal particles through entrapment and blockage	Dual or multimedia filters, pressure or gravity	Filtration rates: 10–15 $m^3/m^2/h$ (4–6 gpm/ft^2)
Media filtration, Stage 2	Removal of fine colloidal particles through entrapment and blockage	Dual or multimedia filters, pressure or gravity	Filtration rates: 8 – 12 $m^3/m^2/h$ (3.2–4.8 gpm/ft^2)
Media backwash unit	Removal entrapped particles from filters.	Backwash pumps and air scour blowers.	Frequency every 8–24 h. Backwash rates 35–55 $m^3/m^2/h$ (15–23 gpm/ft^2) for ~10 min. Air scouring rates 55–90 m/hr (3–5 scf/ft^2-min) for ~5min
Acid or scale inhibitor dosing	Preventing scale formation in RO unit	Dosing system for scale inhibitor: pumps and tanks	Dosing rate, acid ~20 ppm, scale inhibitor 1–5 ppm
Sodium bisulfite dosing	Applied for reduction of residual free chlorine	Dosing system for sodium bisulfite: pumps and tanks	Dosing rate ~three times residual chlorine concentration
Cartridge filtration	Protecting pumps and membranes from particulate matter	50 cm (20″) long cartridges, rated at 5–15μ nominal porosity	Filtration rate about 2 m^3/h (9.0 gpm) per 50 cm (20″) cartridge

TABLE 11.7
Summary pretreatment configuration with submersible membrane filtration

Treatment step	Objectives	Equipment, materials	Process parameters
Intermittent chlorination (optional)	Mitigate bioactivity in supply lines	Chlorine injection as chlorine gas or NaOCl solution	Dosing rate 1–5 ppm, duration few hours per week
Coagulation (optional)	Destabilization of colloidal particles	Dosing of $FeCl_3$ or $Fe_2(SO_4)_3$	Ferric: 0.5–30 ppm, strong mixing, residence time 0.5–1 min
Flocculation (optional)	Agglomeration of colloidal particles	In line flocculation	Gentle mixing, residence time 10–20 min
Membrane filtration	Removal of colloidal particles using membrane barrier	Submersible membrane modules, tanks, vacuum pumps	Filtration rates: 34–68 l/m^2/h (20–40 gfd)
Membranes backwash	Removal foulants from membrane surfaces.	Backwash pumps and air scour blowers	Frequency of filtrate backwash: every 15–60 min. Backwash rates: 70–130 l/m^2/h (40–80 gfd). Air scouring frequency every 10–60 s, 30–40 st m^3/min (1000–1500 scmf)
Chemical enhanced backwash (CEB)	Removal of foulants from membrane surfaces.	Backwash pumps, chemical dosing pumps for NaOCl and NaOH solution	CEB frequency once–twice per day. Conditions as during backwash
Acid or scale inhibitor dosing	Preventing scale formation in RO unit	Dosing system for scale inhibitor: pumps and tanks	Dosing rate, acid ~20 ppm, scale inhibitor 1–5 ppm
Sodium bisulfite dosing	Applied for reduction of residual free chlorine	Dosing system for sodium bisulfite: pumps and tanks	Dosing rate ~three times residual chlorine concentration
Cartridge filtration (optional)	Protecting pumps and membranes from particulate matter	50 cm (20″) long cartridges, rated at 5–15μ nominal porosity	Filtration rate about 2 m^3/h (9.0 gpm) per 50 cm (20″) cartridge length

$$E = E_{ws} + E_{pt} + E_{pu} + E_{pe} + E_{ax} \tag{65}$$

In RO plant, except for small quantity of energy that is used for lightning and to operate control system, the rest is used to power water pumping equipment. Energy of pumping (E_ρ) is direct function of flow rate (Q_f), head differential (ΔP_f) developed by the pump and specific gravity (ρ) of the fluid being pumped. Energy usage for the pumping is inversely affected by hydraulic efficiency of the pump (η_ρ) and transformation efficiency of the electric motor (μ_m).

$$E_\rho = \mathrm{k} Q_f \Delta P_f \rho/(\eta_\rho \eta_m) \tag{66}$$

where k is units conversion constant

If flow rate is expressed in m^3/h and pressure in bar then the vale of constant k in Eq. 66 is 0.02777. If the flow rate unit is gpm and pressure expressed as psi then k value is 0.000427. The results of calculations are given as kWhr. The ΔP_f term in Eq. 66 indicates that feed pump provides pressure differential between pump suction pressure and pump discharge pressure. The pump suction pressure is usually not the same as the ambient pressure.

During the RO process the feed stream is split into two streams: permeate and concentrate. The fraction of feed energy proportional to permeate flow is used directly for the RO process, i.e., ions-water separation, transport of permeate through the membrane barrier and to compensate for friction flow resistance in the permeate channel. At the outlet from RO unit the permeate pressure is practically equal to the ambient pressure (potential of making additional work close to zero). Some of the remained feed energy is only partially used to overcome friction flow resistance in the feed channel. The rest of the energy is associated with the concentrate flow, which leaves the RO unit under pressure. The flow rate and pressure of the concentrate stream depends mainly on type of application: nanofiltration, brackish or seawater. The largest fraction of energy remained in concentrate stream and available for recovery is in the seawater RO application.

The representative conditions for a seawater process are shown in Fig. 11.5. In this example Concentrate flow if 50% of the feed flow and its pressure is about 97% of the feed pressure. Therefore potentially, about 48% of the energy spent to pressurize feed stream, is available for recovery. In practice, the fraction of feed energy that is possible to recover is much smaller. On the feed side, as it is shown in Eq. 64, not all electric energy is converted to water pressure. Part of it is used up in thermal losses due to inefficiency of pumps and electric motors. Similar situation applies to the concentrate stream.

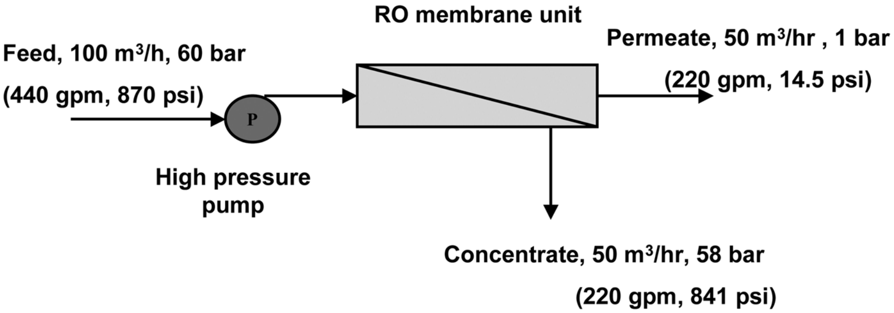


FIG. 11.5 A diagram of flows and pressures in seawater RO unit.

The energy recovered (E_ρ) in energy recovery device is function of concentrate flow rate (Q_c), head differential (ΔP_c) between inlet and outlet of energy recovery device and its efficiency (η_t) and specific gravity of the concentrate stream (ρ). It is given as:

$$E_\rho = \mathrm{k} Q_c \, \Delta P_c \, \rho \, \eta_t \tag{67}$$

Accordingly, net energy spent in the pumping unit (Eq 63) is:

$$E_{pu} = E_\rho - E_r \tag{68}$$

The Eq. 63 can be modified as:

$$E = E_{ws} + E_{\rho t} + E_{\rho e} + E_{ax} + E_\rho - E_r \tag{69}$$

Table 11.8 provides representative values of pumping energy requirements in nanofiltration, brackish and seawater systems, assuming that feed water pumping unit includes power recovery device. It is evident, based on the calculation results listed in Table 11.8, that power recovery device is not a viable option for the nanofiltration systems. Due to low operating pressure and high recovery rate there is practically no energy left to recover. In brackish water systems, feasibility of installing power recovery devices will depend on feed salinity, recovery rate and amortization period. In seawater RO systems power recovery devices make significant contribution to reduction of the operating cost.

TABLE 11.8
Representative values of pumping energy requirements in RO systems

Application	NF	RO, low TDS	RO, high TDS	SW
Permeate capacity, m^3/d (mgd)	40,000 (10.6)	40,000 (10.6)	40,000 (10.6)	40,000 (10.6)
Recovery rate, %	85	80	65	50
Feed flow, m^3/h (gpm)	1,961 (8,635)	2,083 (9,172)	2,564 (11,920)	3,333 (14, 676)
Concentrate flow, m^3/h	294.1 (1,295)	416.7 (1,835)	897.4 (3,951)	1666.7 (7,339)
Raw water pump differential head, bar (psi)	5 (72.5)	5 (72.5)	5 (72.5)	3 (43.5)
Feed pump suction head, bar (psi)	1 (14.5)	1 (14.5)	1 (14.5)	1 (14.5)
Feed pump discharge pressure, bar (psi)	7 (101.5)	15 (217.5)	25 (362.5)	65 (942.5)
Concentrate pressure, bar (psi)	2 (29)	10 (145)	20 (290)	62 (899)
Concentrate discharge pressure, bar (psi)	0.5 (7)	0.5 (7)	0.5 (7)	0.5 (7)
Permeate pumping pressure, bar (psi)	5 (72.5)	5 (72.5)	5 (72.5)	5 (72.5)
Raw water and permeate pump efficiency, %	0.78	0.78	0.78	0.78
Feed pump efficiency, %	0.84	0.84	0.84	0.84
Electric motor efficiency, %	0.94	0.94	0.94	0.94
Power recovery device efficiency, %	0.82	0.82	0.82	0.82
Specific gravity, feed	1	1	1	1.02
Specific gravity, concentrate	1	1.01	1.01	1.03
Energy of raw water pumping, kWh	364.2	387.0	476.3	379.0
Energy of feed pump, kWh	437.1	1083.6	2286.3	8084.3
Energy recovered, kWh	–9.9	–88.4	–390.9	–2289.5
Energy of permeate pumping, kWh	309.6	309.6	309.6	309.6
Total energy usage, Kwh	1101.1	1691.8	2681.3	6483.4
Specific power consumption, kWhr/m^3 (kWh/kgal)	0.66 (2.50)	1.02 (3.87)	1.61 (6.09)	3.89 (14.7)
Fraction of energy recovered, %	0.9	5.0	12.7	26.1
Power cost savings potential (@$0.06/kWh), $/year	4,683	41,816	184,911	1,083,025

11.2.4 Pumping equipment for RO applications

Feed pumps used in commercial RO systems are either centrifugal or positive displacement type. Positive displacement pumps have higher hydraulic efficiency then centrifugal pumps. However, their output capacities are limited and, on the average, they require more frequent maintenance then the centrifugal type pumps. Positive displacement pumps are used in small capacity systems, where they have distinct efficiency advantage: 90% vs. 65–75% for centrifugal type. In large capacity RO systems centrifugal type pumps are used almost exclusively. Their capacity, in respect of RO train feed flows, is practically unlimited and efficiency of high capacity pumps is being close to the efficiency of positive displacement pumps: 84–88%.

In RO applications, pressure and flow of positive displacement and centrifugal pumps, is controlled in different ways. Positive displacement pump produces output that is almost constant in the wide range of pump capacity, regardless of output pressure. Therefore, as shown in Fig. 11.6, output flow of the positive displacement pump is regulated by throttling valve located on bypass line connecting pump output and its suction. Feed pressure in the membrane unit is regulated by throttling the RO unit concentrate valve. In centrifugal pump, output head varies with flow, increasing as flow decreases.

The centrifugal pump control configuration is shown in Fig. 11.7. The pump output flow is controlled by throttling valve located on the pump dis-

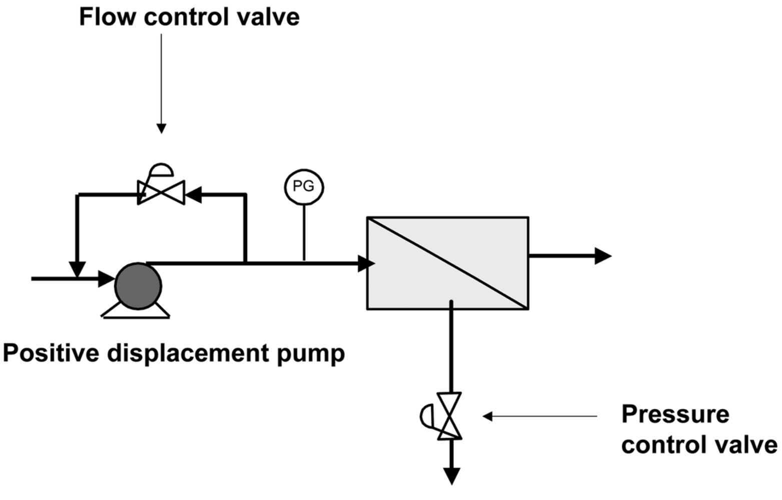


FIG. 11.6 Controlling flow and pressure in RO unit with positive displacement pump.

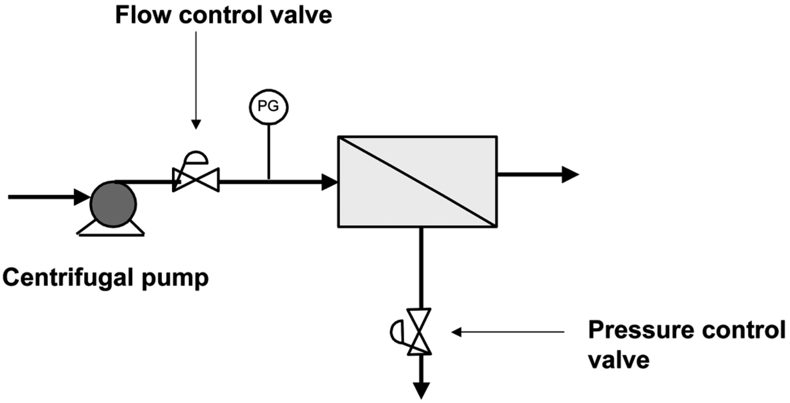


FIG. 11.7 Controlling flow and pressure in RO unit with centrifugal pump.

charge line. Feed pressure to the membrane elements is controlled by throttling the RO concentrate valve.

As discussed already in Chapter 10.10 performance of membrane elements changes with time. The expected changes are gradual decrease of permeability, in the range of 5–15% per year, and increase of salt passage, increase of about 10% per year. Usually the rate of performance change is not linear, showing higher rate initially and leveling off with operating time. These changes are more pronounced in systems processing surface water. Brackish systems processing well water experience higher membrane performance stability. In addition to time paced membrane performance changes, both water permeability and salt transport are affected by feed water temperature. Again, surface water usually has wider temperature fluctuation then the well water.

Selection of pumping equipment and unit configuration has to account for need to increase feed pressure with time and adjust feed pressure to compensate for feed water temperature fluctuations. At some locations (52) feed water salinity may fluctuate as well, providing another variable to RO process design. The simplest design solution: use of oversized feed pump and throttle it as required, results in unnecessary waste of energy. Widespread use of electric motors equipped with variable frequency drivers (VFD), provides adequate solution for feed pressure adjustment in nanofiltration and brackish systems. In seawater systems VFD could be quite expensive, if installed on large capacity electric motors. In addition, VFD introduces some additional inefficiency (about 2%), which, in energy intensive seawater systems could be significant.

From number of configuration alternatives available, the more common solution is to use a two stage feed pumping system: a booster pump and main feed pump, connected in series. The booster pump, equipped with VFD, is sized to provide sufficient pressure to cover expected fluctuation of feed pressure in the RO unit and the main feed pump operates at constant differential head. This arrangement results in power saving but introduces additional pump, motor and control equipment. A solution to simplify the pumping unit configuration in seawater systems and reduce number of pumping stages is to use the pump that transfers water from the clear well to the cartridge filters as a booster, pressure modulating pump. As a pressure modulator, the required head from this pump is in the range of 5–15 bar. (75–215 psi). To apply this pumping configuration it is necessary however, to use high pressure rated piping between discharge of the transfer pump and suction of high pressure pump. In addition, the housing of cartridge filters has to be rated for high pressure operation.

There number of pressure recovery devices types being used currently in RO applications:

1. Reverse running pump
2. Pelton wheel
3. Turbocharger
4. Pressure exchanger (isobaric device)

Reverse running pumps were very popular power recovery devices in the past due to their reliability and simplicity of operation. Their advantage is that discharged stream from this type of power recovery devices can be under pressure and therefore enable convenient transmission of the concentrate to the disposal point. One of the more significant disadvantages was that outside the design operating point the hydraulic efficiency could be significantly lower.

Pelton wheel, the device that is gaining acceptance lately, has a relatively flat flow–efficiency pump performance curve. Therefore, power recovery efficiency remains more or less constant over the range of concentrate pressure fluctuations. The major drawback is that effluent exits Pelton wheel at atmospheric pressure. Therefore, plant location should be such that the Pelton wheel effluent will reach discharge point by gravity. Otherwise, additional pumping of concentrate is necessary for disposal. Pelton wheel is an impulse turbine with high hydraulic efficiency in the 84–88% range. Pelton wheel consists of a rotor with cup shaped buckets mounted on the rotor circumference (Fig. 11.8). The concentrate jet from a nozzle impinges the buckets, turning the rotor. The Pelton

FIG. 11.8 Picture of Pelton wheel showing rotor with buckets.

wheel usually operates on a common shaft with electric motor and a high pressure pump, reducing energy requirement of the motor (Figs. 11.9, 11.10).

The turbocharger, is a close coupled two pump rotors device on a common shaft (Fig. 11.11). On rotor receives concentrate flow under pressure and the second one generates differential head of the feed water. The rotors, inlet and outlet connections are configured in a way that both rotors are turning in the same direction. The rotor processing concentrate is providing driving force for the one processing feed stream. This device is very compact. The efficiency of turbocharger is a product of individual efficiencies of two rotors. For example if

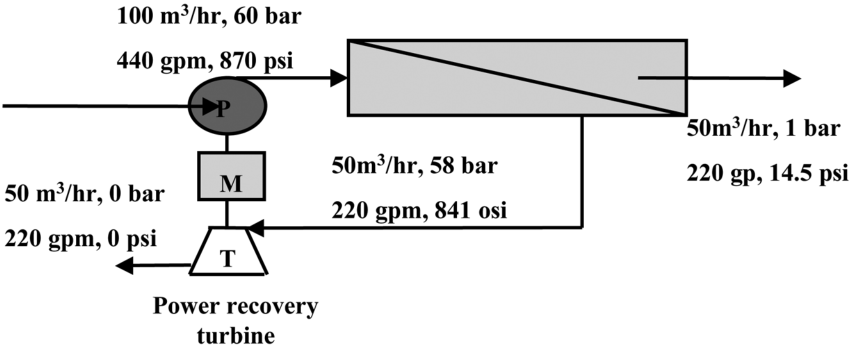


FIG. 11.9 Configuration of high pressure pump with Pelton wheel energy recovery device.

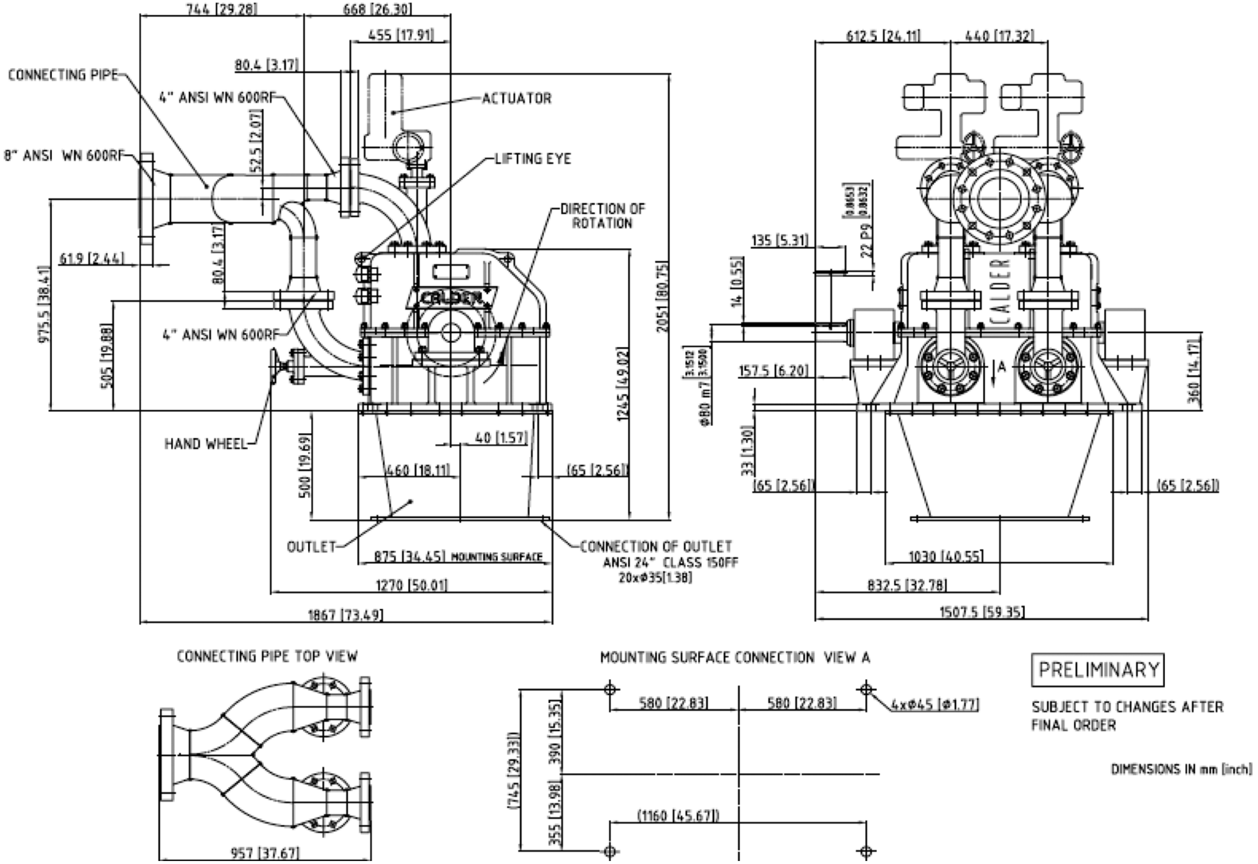


FIG. 11.10 Engineering drawing of Pelton wheel equipment (courtesy of Calder Corporation).

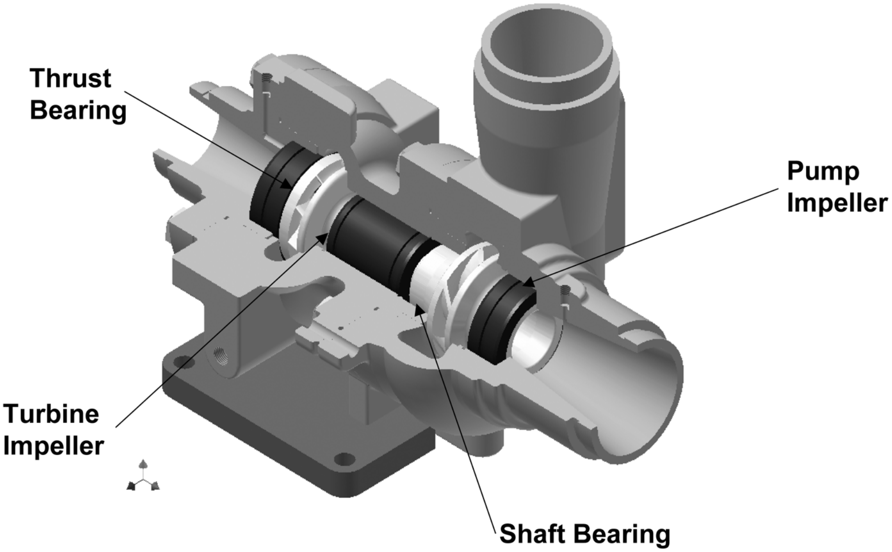


FIG. 11.11 Sectional cut off of "Turbocharger" pumping device (courtesy of Pump Engineering).

both rotors have efficiency of 90% then the overall efficiency will of the device will be 0.9 × 0.9 = 0.81. The turbocharger is usually applied to provide interstage boost pressure in a two stage RO unit or providing pressure increase of the feed stream following the high pressure pump (Fig. 11.12). Use of turbocharger reduce differential head requirement of the main high pressure pump.

The pumping devices that are gaining increased acceptance recently in RO seawater systems are pressure exchangers (also called isobaric devices) of various configurations. Two better known types are DWEER and PX (Figs. 11.13–11.15). Both are positive displacement pumping devices. The DWEER design consists of two parallel cylindrical chambers, each with piston separating feed from the concentrate. Initially, feed water fills one chamber pushing piston to extreme position. In the next step concentrate, under pressure pushes the piston in reverse direction, pressurizing the feed and displacing it from the chamber. In the parallel chamber the reverse process takes place simultaneously. The issue of concern for this type device is reliable operation of valves, regulating timely flow of feed and concentrate.

The PX device consists of ceramic rotor in a pressure vessel, closed by two end caps (Fig. 11.15). One cap provides entrance and exit ports for feed and the other has two corresponding ports for concentrate, but in reverse order. The

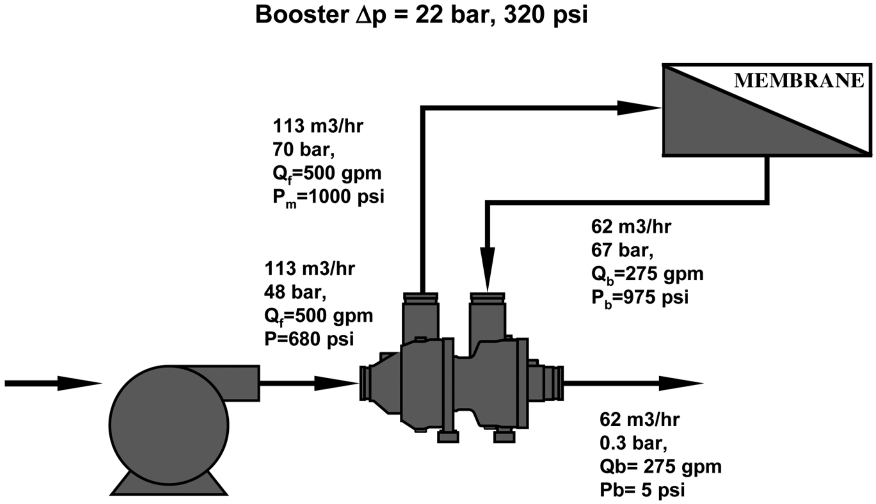


FIG. 11.12 Single stage RO unit with "Turbocharger" pressure booster (courtesy of Pump Engineering).

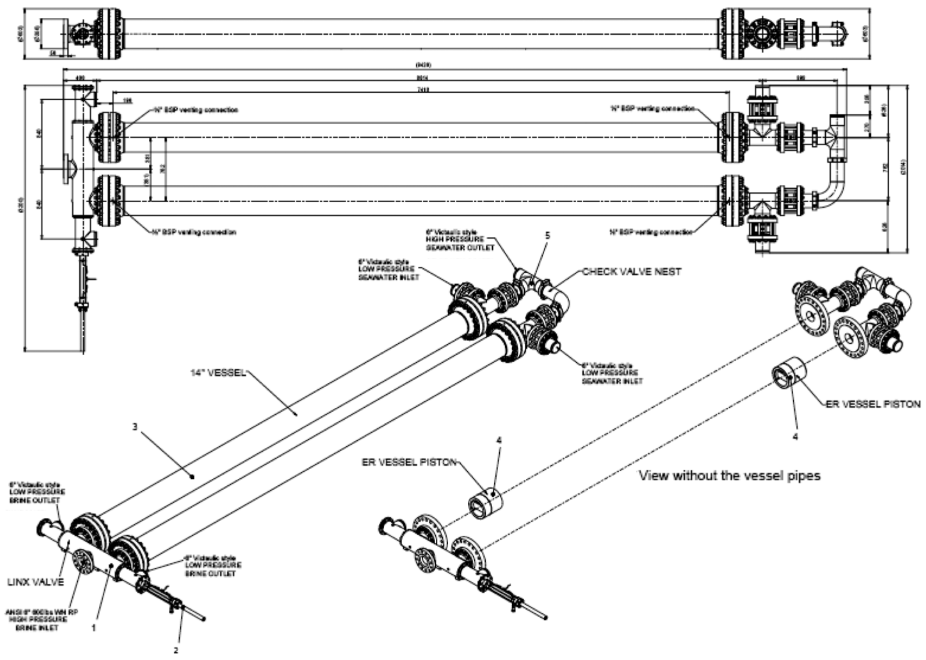


FIG. 11.13 Schematic drawing of DWEER pumping-power recovery device (courtesy of Calder Corporation).

FIG. 11.14 Picture of DWEER pumping- power recovery device at the seawater RO Ashkelon plant (courtesy of Israel Desalination Engineering).

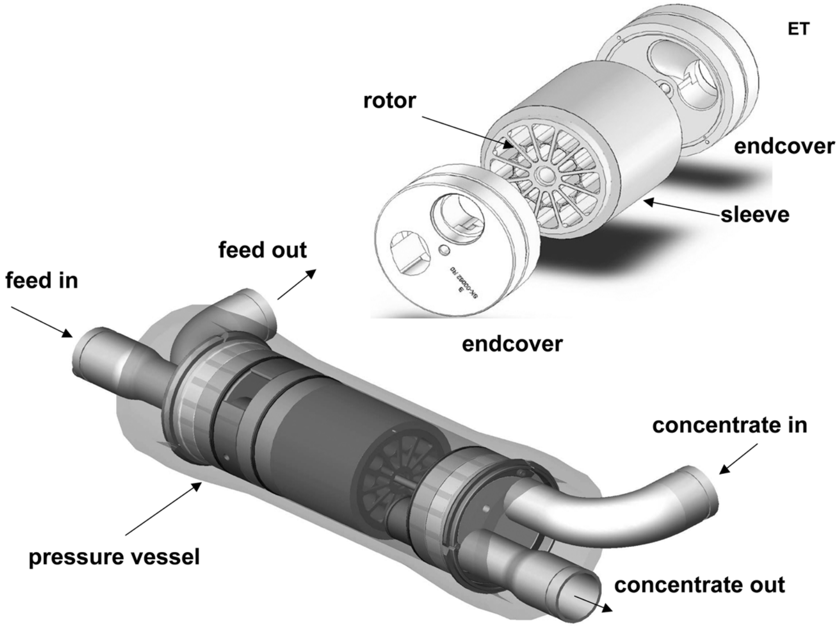


FIG. 11.15 Configuration of PX pressure exchanger (courtesy of ERI).

rotor has evenly spaced cylindrical cavities, open through the rotor and positioned radially, parallel to it axis. During operation rotors turns and its open cavities coming successively in front of feed and concentrate ports located in the end caps. In one position cavity is filled with feed water at low pressure and as rotor turns, the cavity moves to a position in front of the concentrate port. There the feed water is replaced with concentrate flowing at high pressure from the opposite end cap. As the rotor continues to turn, cavity moves to the position where is filled again with feed, replacing the concentrate. In such direct contact concentrate transfers its energy to the feed stream. During this operation some mixing (about 3–4%) between concentrate and feed occurs. The mixing phenomena creates some increase of salinity of the feed stream (about 1.5–2%), which results in somewhat higher permeate salinity and increase of required feed pressure. Corresponding mixing effect in the DWEER equipment is smaller than in the PX units, as the feeed and concentrate are separated by pitons.

The flow diagram illustrating principle of operation of pressure exchanger device is shown on Fig. 11.16.

In the current configuration, of RO units operating with pressure exchange unit, the feed stream is split into two streams: F1, which has flow rate equal to flow rate of permeate and F2 equal to the concentrate flow. Stream F1 is processed by high pressure pump up to the feed pressure required. Stream F2 is processed by pressure exchanger. Stream F2 exchanges its energy with the concentrate

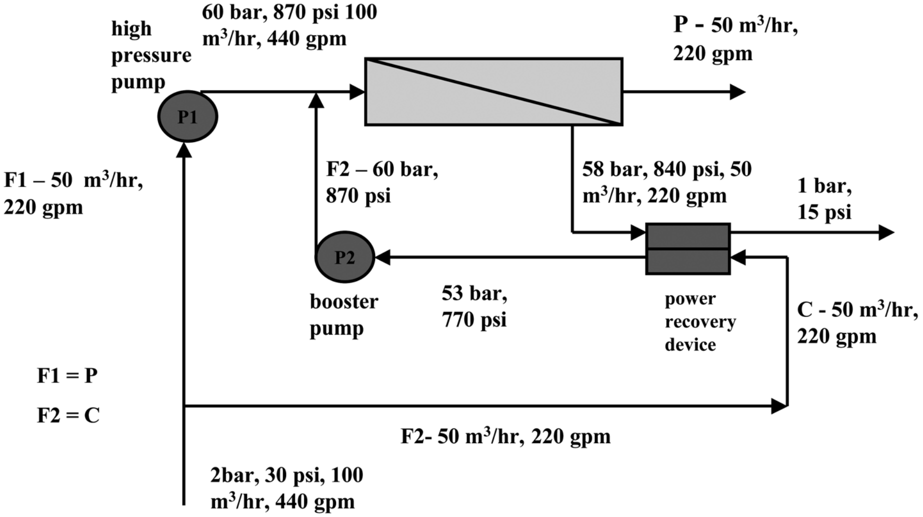


FIG. 11.16 Configuration diagram of RO unit with pressure exchanger.

stream and exits pressure exchanger at pressure slightly lower than the pressure of the concentrate and lower than the feed pressure. Therefore, after exiting from the pressure exchanger unit, pressure of the F2 stream is increased up to the feed pressure of the F1 stream by a booster pump. At the entrance to the RO membrane train both streams (F1 and F2) are combined.

The pressure exchanger is a positive displacement device and operates with very high energy exchange efficiency, in the range of 94–96%. Due to the high efficiency and direct transfer of concentrate energy to the feed water, pressure exchangers provide significant energy savings as compared to more conventional power recovery equipment. Flow capacity of these devices is still limited to 50–150 m^3/h (220–660 gpm). However, due to advantage of high efficiency, these devices are used already even in a very large systems. Currently the largest desalination system using pressure exchange equipment is the 330,000 m^3/d (87 mgd) seawater Ashkelon Plant (Fig. 11.2).

Table 11.9 includes comparison of power calculation for three configurations of pumping equipment in seawater system. The calculations were conducted for a RO system processing Mediterranean feed (40,600 ppm TDS) at recovery rate of 50%, feed water temperature of 22 C, average permeate flux rate of 14.0 l/m^2/h (8.2 gfd) utilizing Hydranautics SWC5 elements and assuming 3-year membrane age which corresponds to 20% flux decline. For the calculations corresponding to a system utilizing pressure exchanger a 3% feed salinity increase was applied. The configurations are: A) standard efficiency pump and power recovery turbine, B) high efficiency pump and Pelton wheel and C) high efficiency pump and pressure exchanger.

The results listed in Table 11.9 demonstrate a clear energy advantage of system configuration, which includes pressure exchanger. The difference of calculated energy consumption for this particular design case is about 0.4 kwhr/m^3 (1.5 kWh/kgal). The above calculations correspond to configuration of a single high pressure pump, pumping from a clear-well up to the entrance to membrane elements. In actual system, most likely there will be a transfer or booster pump that will provide variable pressure boost for high pressure pump. The VFD will be controlling only the motor of the booster pump. Therefore, in energy calculations the efficiency of VFD should be applied only to the relevant fraction of feed pressure. The pumping unit configuration, that includes pressure exchanger (Design case C), corresponds to the diagram shown on Fig. 11.16 and calculation of energy consumption was conducted accordingly.

In NF and brackish RO systems feed pump could be of horizontal or vertical configuration. Power recovery devices are usually of the same orientation as is the pump. In brackish systems Turbochargers and pressure exchangers can be

TABLE 11.9
Summary of power calculation for different configurations of pumping units in RO seawater systems.

Design case	A	B	C
Configuration	Pump + power recovery turbine	High efficiency pump + Pelton wheel	High efficiency pump + pressure exchanger
Pump efficiency, %	82	88	88
Pelton wheel/pressure exchanger efficiency, %	82	88	94
Electric motor efficiency, %	94	96	96
VFD efficiency, %	98	98	98
Raw water and pretreatment pressure losses, bar (psi)	4 (58)	4 (58)	4 (58)
Feed pressure, bar (psi)	64(928)	64(928)	64.5 (935)
Concentrate pressure, bar (psi)	62 (899)	62 (899)	63.5 (921)
Concentrate discharge pressure, bar (psi)	0.5 (7)	0.0	0.5 (7)
Permeate pumping, bar (psi)	10.0 (14.5)	10.0 (14.5)	10.0 (14.5)
Auxiliary equipment, kWh/m^3 (kWh/kgal)	0.05 (0.19)	0.05 (0.19)	0.05 (0.19)
Pretreatment losses, kWh/m^3 (kWh/kgal)	0.30 (1.14)	0.28 (1.06)	0.28 (1.06)
High pressure pump, kWh/m^3 (kWh/kgal)	4.34 (16.43)	4.14 (15.67)	2.08 (7.87)
Power recovery, kWh/m^3 (kWh/kgal)	–1.37 (–5.18)	–1.45 (–5.45)	+0.19 (+0.72)
Permeate pumping, kWh/m^3 (kWh/kgal)	0.36 (1.36)	0.35 (1.32)	0.35 (1.32)
Total, kWh/m^3 (kWh/kgal)	3.68 (13.94)	3.37(12.79)	2.95 (11.16)

used to reduce energy required by main feed pump or to recover some fraction of energy remained in the concentrate.

In seawater systems the more common high pressure pump type used is of split case horizontal configuration. As a power recovery device, reverse running pump and turbochargers were commonly used in the past. At present, Pelton wheel is becoming more popular due to higher efficiency. In addition, constant efficiency in the wide range of inlet head and flow of the concentrate provides additional operating flexibility with lower energy penalty, compared to use of reverse running pump.

11.2.5. Optimization of energy consumption

The approach to optimization of energy consumption will depend mainly on feed water salinity. In respect of energy consumption there are four basic categories of RO applications:

1. Nanofiltration
2. Low salinity RO
3. High salinity RO
4. Seawater

Nanofiltration systems, operate at low feed pressure of 5–8 bar (72–116 psi). Feed water is of low salinity and salt passage is relatively high. Therefore increase of system recovery rate will, in majority of cases, result in reduction of specific power consumption. Recovery rates of 80–90% are quite common in NF applications. The recovery rate is usually limited by scaling potential of the feed water. Due to high recovery rate, nanofiltration units are configured as two or three stage array systems with six to seven elements per vessel. In such systems, with large number of elements in series (14–21), a significant fraction of the feed water energy is lost to friction in the feed channels of membrane elements. Therefore, system design that results in operation at low permeate flux rate, i.e., low feed flow rate per pressure vessel, will result in smaller pressure drop per stage and therefore reduction of required feed pressure. The operation at lower feed pressure and correspondingly lower energy consumption has to be evaluated against higher capital cost, additional membrane replacement and higher permeate salinity. As shown in examples included in Table 11.8, amount of energy remained in NF concentrate is to low to be considered for recovery by application of any energy recovery devices.

Optimization of design for systems treating low and high feed salinity RO may result in different configurations in each case. Usually, low salinity RO systems (feed salinity of 1000–2000 ppm TDS) can be operated at higher recovery rate in respect of membrane scaling concerns. In system operating with low salinity feed, osmotic pressure is relatively small fraction of the feed pressure and increase of osmotic pressure with increased recovery rate is not significant. Therefore, reduction of pumping energy due to smaller pumping volume at higher recovery, usually will offset any pressure increase due to higher average osmotic pressure. In addition, at low feed salinity, an increase of permeate salinity with increased recovery is usually less of concern.

In RO systems, operating with high salinity brackish feed all the above factors have to be considered at the design stage. It is possible, that for very high salinity feed, that maximum recovery rate will be limited by overall process economics or permeate salinity rather then scaling potential of the concentrate. Feasibility of use of energy recovery devices will depend on quantity of energy in the concentrate, available for recovery. Application of energy recovery equipment will be more cost effective in large systems. There, concentrate streams from a number of trains can be combined together and processed by one energy recovery device, therefore reducing cost of equipment and providing usable energy output. In such centralized energy recovery system the concentrate flow, feeding energy recovery device, may vary with number of RO units in operation. It is important therefore to select energy recovery equipment that has efficiency relatively constant in the range of the expected flow rates.

Additional considerations in NF and RO projects are cost of raw water and concentrate disposal. At some locations economics of raw water supply and concentrate disposal could be predominant factors in the design optimization process. In seawater RO systems, cost of raw water supply and concentrate disposal, although could be significant, usually is not an important parameter to be considered in optimization of recovery rate. The predominant factor is usually osmotic pressure associated with given recovery rate and its effect on feed pressure and energy consumption.

The optimum recovery rate will depend on number of site specific parameters: feed salinity and temperature range, cost of chemicals used in the pretreatment system and cost of electricity. Cost of pretreatment and pumping equipment and efficiency of pumps, motors and energy recovery devices are also relevant parameters of the optimization process. Fig. 11.17 shows dependence of energy consumption on recovery rate for RO system equipped with high efficiency pump and Pelton wheel turbine, processing seawater of 35,000 ppm TDS salinity. The energy consumption results were calculated for feed water temperature of 14 and 28°C. As can be seen the minimum energy consumption is around 45–50% recovery rate and the position of the minimum depends on

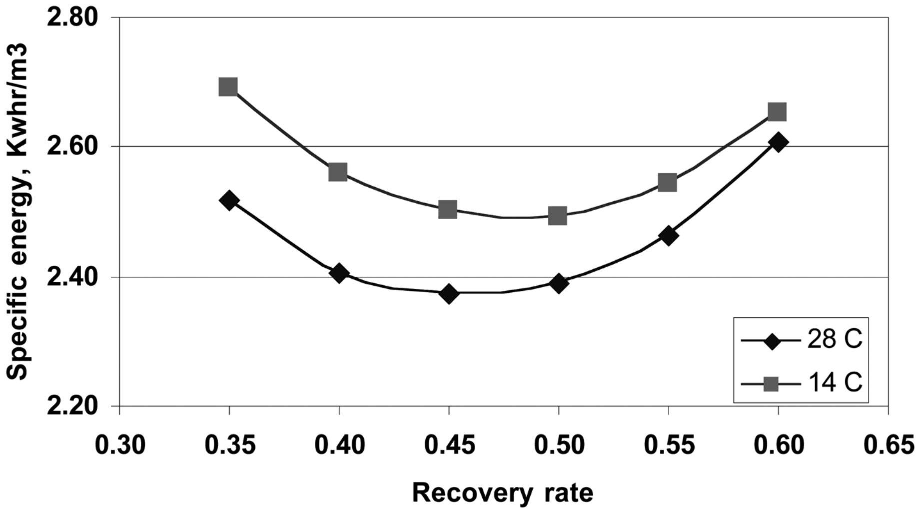


FIG. 11.17 Optimization of power consumption in RO seawater unit vs. recovery rate.

feed water temperature, shifting to lower recovery rate as feed water temperature increase.

The above figure demonstrates multitude of parameters that may affect optimization of energy consumption. The additional factor to be considered is change of membrane performance with time, mainly permeability decline, due to compaction or fouling. The above discussion points to the conclusion that the selection of pumping unit should result in system configuration that will enable operation in the optimum range of process parameters. The system should be flexible enough to enable fine tuning of process parameters during the actual operation as the local conditions such as raw water conditions, product water demand and/or economic factors may fluctuate with time.

11.2.6.Configuration of RO trains

The RO trains configuration related decisions to be considered at the design stages are:

1. Number of elements per vessel and number of pressure vessel per train.
2. Number of desalting stages per train and their configuration.
3. Capacity of RO trains and number of trains in the system.

All the above considerations are interrelated to some extend and affect both capital and operating cost. In commercial systems number of elements per pressure vessel is either 6 or 7. More recently large seawater systems are being designed and some already in operation, with 8 elements per vessel. The decision on number of elements per vessel depends on unit configuration i.e., number of desalting stages, which in turn depends on the recovery rate. Nanofiltration and brackish RO system that are designed to operate at recovery rate up to 85% are configured as a two stage trains. The RO systems, configured as a two stage units, almost universally use array of seven elements per vessel. Systems designed to operate at recovery rate of 90% are configured as a three stage trains. Some of three stage trains are configured with six elements and some with seven elements per vessel. The three stages train configuration of seven elements per vessel has lower capital cost then six elements per vessel configuration, but higher pressure drop across the system. There are 21 elements in series as opposed to 18 and the average feed flow per vessel is higher by about 16% due to lover number of pressure vessels. In three stages nanofiltration systems operating at high recovery (~90%), there is very little NDP available in the last stage. To correct this condition it is quite common to have first two stages configured as a single train unit. The last stage is configured as a separate train, with dedicated feed pump. The separate third stage train usually is feed by concentrate from number of two stage trains. Such configuration results in better performance of the membranes in the third stage (operation at higher flux rate) and lower capital cost.

Seawater systems, processing regular seawater, operate at recovery rate range of 35–50%. In the past, two stages seawater systems were quite common with five or six elements per vessel. The current design approach however, is to configure seawater systems in a single stage configuration with seven or even eight element per vessel. The advantages of eight elements versus seven elements configuration are quite significant (47), providing stable performance and better economics of the process.

With the increasing capacity of RO plants there is a tendency to increase trains size as well. There are number of considerations regarding train size. Some are related to flexibility of plant operation, maintaining sufficient output capacity during maintenance operation and so on. Recently, train size limitation was related to probability of interconnector o-rings failure and related trains down time (56). One of the major factors in determination train size is the logistic of membrane cleaning operation, i.e., required capacity of the membrane cleaning unit and convenience of preparation of cleaning solutions. During cleaning operation flow rate of cleaning solution per vessel is about 7–9 m^3/h. (30–40 gpm). The cleaning tank has to be of sufficient size to provide operating volume of about 5 min. for the cleaning pump. To clean a train segment of 100 pressure vessels a cleaning tank volume of 40 m^3 (11,000 gal) is necessary. Assuming a connecting piping volume as be similar to the tank volume, about 80 m^3 (22,000 gal) of cleaning solution has to be prepared. A single cleaning operation takes about 8–24 h. Cleaning a train segment of 100 pressure vessels is probably the most can be accomplished in one cleaning. Considering these limitations a train consisting of more that 200 pressure vessels is not very likely to provide efficient operation, as the off line time due to single cleaning would most likely exceed two days at a time. Train with 200 pressure vessels will contain 1,400–1,600 membrane elements. A single brackish water element produces at field conditions about 22 m^3/d (5800 gpd). Capacity of seawater element is about 12 m^3/d (3200 gpd). The above numbers determines the upper limit of a single train permeate output capacity.

Until recently the plant configuration usually followed the 1 to 1 symmetry of membrane trains and pumping systems (Figs. 11.18, 11.19).

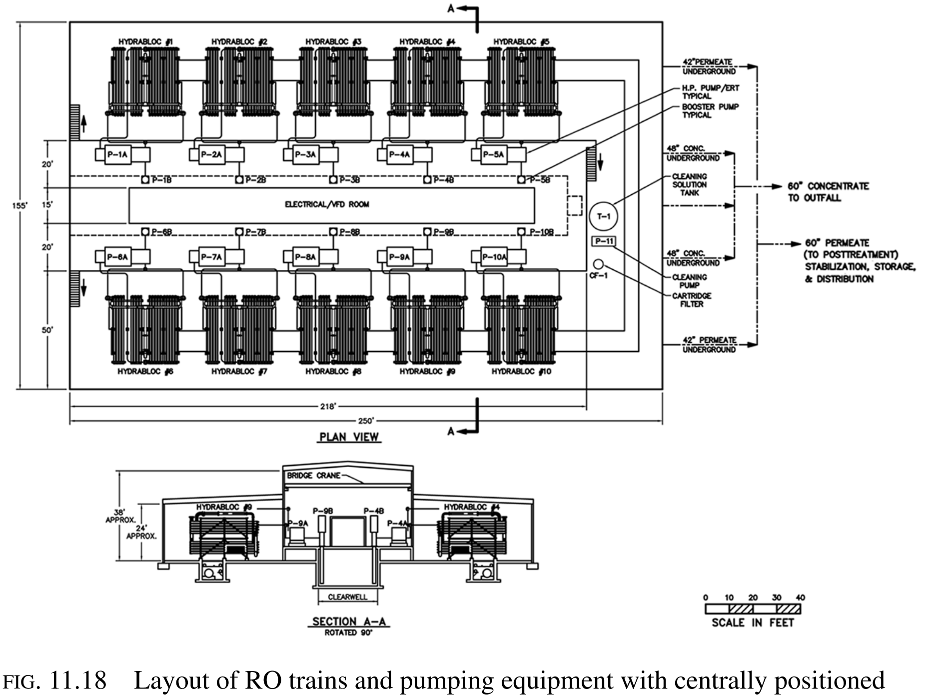


FIG. 11.18 Layout of RO trains and pumping equipment with centrally positioned clear well.

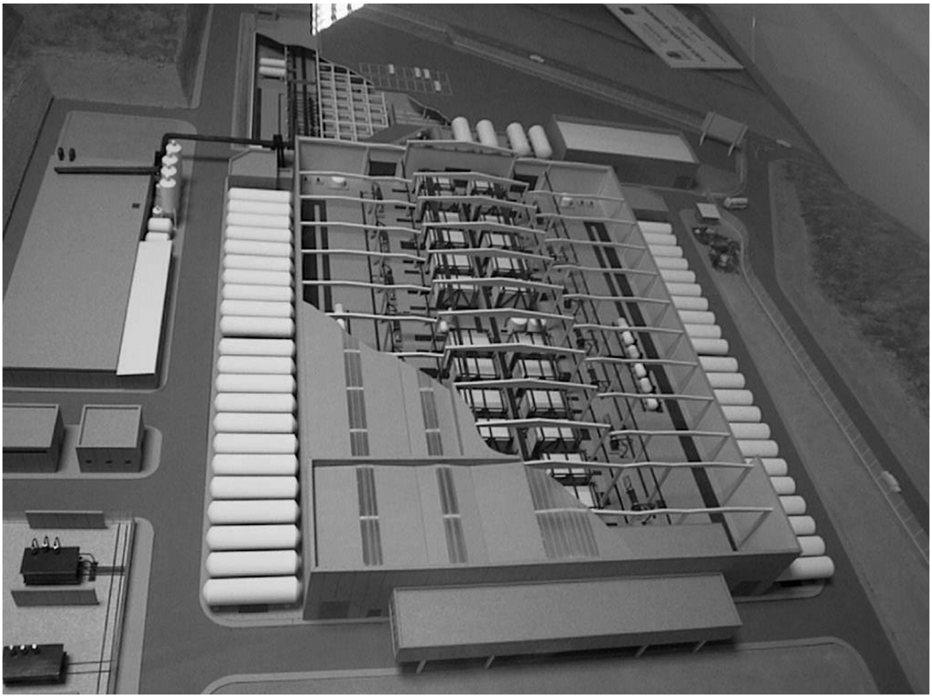


FIG. 11.19 Layout of RO trains and pumping equipment with horizontal pressure filters (courtesy of Endesa Corporation).

The configuration shown on Fig. 11.18 a is of RO plant equipped with filtration system (either gravity or membrane filtration unit) providing filtration effluent to the clear well located in the central part of the RO building, below the trains level. The transfer (booster) pumps pump filtrate from the clear well through cartridge filters to the suction of high pressure pumps. These in turn pump feed water under high pressure to the dedicated RO trains. Fig. 11.18 shows floor plan for a RO seawater system for a capacity of about 200,000 m^3/d (53 mgd). The system is divided into ten RO trains. Each train consists of 200 pressure vessels in a single stage configuration. Each vessel contains eight elements. The RO building footprint is 47m × 76 m (155′ × 250′). The maximum height is 12 m (38′). The second plant configuration shown on Fig. 11.19 is of 120,000 m^3/d (32 mgd) seawater RO plant. It is also of a system with RO trains with dedicated high pressure pumps and energy recovery turbines. Here the seawater is filtrated by horizontal pressure filters located symmetrically on both sides of the RO building. Filtrate from media filters flows directly under pressure through cartridge filters to the suction of the high pressure pumps, which pump feed water to dedicated trains. In the current design of some large seawater plants high pressure pumps, membrane trains and energy recovery turbines are separated into three functional centers (see Fig. 18.2, Chapter 18). According to the designers of such plant configuration (56) this design concept provides better flexibility in selecting equipment size. For example high pressure pumps can be selected based on best efficiency and lowest capital cost rather then to provide feed flow required by a single train. The same considerations apply to the energy recovery equipment.

11.2.7.Control and monitoring system

Process control and monitoring in commercial RO desalination systems utilizes almost exclusively computer based Supervisory Control And Data Acquisition (SCADA) system. The SCADA configuration includes central computer providing video display, data storage and reports. It is connected through data highway with a distributed network of process monitoring and controlling microprocessors–programmable logic controllers (PLC's). The local microprocessors evaluate process parameters of designated system unit and control its operation within determined limits. Local PLC's communicate with the central control unit usually through fiber optics cables that provide connection free of electric noises. Process control is achieved through evaluating the output signal from sensors, installed in the plant, and controlling operation of pumps and valves. Operation of the control system is supported by uninterrupted energy suppliers (UPS) that provides sufficient energy to maintain control system operational during the gap period between the time of failure of the main energy supply and the time that the emergency energy generator is operational. The control system usually should have sufficient redundancy to enable maintaining process control capability also in case of partial failure of control equipment. Transfer of process control to backup equipment is done automatically based on self diagnostic capability of the control equipment.

Majority of currently applied control systems provide the following functionality:

1. Protect system from operating at conditions that may result in equipment damage.

   For example: Equipment operation is started in predetermined sequence. Pumps are protected from operation at inadequate suction pressure, at extreme pH or temperature feed water is diverted to drain etc. . . .

2. Maintain equipment operation within the design process limits.
   For example: Operation is controlled to maintain design limits of feed temperature, pressure, flow etc.

3. Maintain production of the design quantity and quality of product water.
   For example: Feed pressure is adjusted to produce design output capacity. Permeate is diverted to drain if design quality is not met. Dosing pumps are controlled to maintain designed pH and hardness of the product water.

4. Stores operating data, generates reports in form of visual display and hard copy.
   For example: Historical results and performance trends reports are generated. Membrane performance results are normalized. Operating cost data are calculated. Operating data are organized to demonstrate regulatory compliance with required product water quality. Records of operating parameters are maintained to satisfy conditions of major equipment warranty terms. The control system is usually divided in to functional sections (control loops) according to logic of plant operation functions performed by individual sections of the plant.
   The functional control loops may include:
   - Intake control loop
   - Pretreatment control loop
   - Main RO control loop
   - RO trains control loop
   - Motor control center loop
   - Electrical circuits and VFD loop
   - Permeate post treatment and storage control loop
   - Wastewater neutralization and discharge control loop

Current programmable controllers are more powerful and are capable to handle large number of input and output signals, perform extensive calculation and control functions. It is possible to have single programmable controller to control the complete desalination process, even of a large plant. Decision between distributed or centralized system control configuration depends on the preferences of the system designer or end user.

The important issue in control systems is security and access level. An example of functional access levels structure is provided below:

Level 1: Anyone. View any display screen except for set point values.

Level 2: Operator level. View any display screen, except for set point values. Ability to stop any equipment. Ability to print reports and enter data.

Level 3: Plant engineer level. Access to all functions of the operator and ability to modify set points and all process parameters.

Level 4: Plant supervisor level. Access to all functions of the plant engineer and ability to add/delete users, change access level and change password.

Level 5: PLC program developer. Unrestricted access and modification authority,

The following process parameters are being monitored in RO plants:

- Raw water conductivity
- Raw water temperature
- Raw water flow
- Raw water pump suction and discharged pressure
- Raw water turbidity
- Dosing rates of pretreatment chemicals
- Raw water free (combined) chlorine
- Media filters head loss
- Filter effluent turbidity
- Filter effluent particle count
- Filter effluent SDI (MFI)
- Cartridge filters pressure drop
- High pressure pump suction and discharged pressure

- Feed water pressure
- Feed water pH
- Feed water free (combined) chlorine
- RO permeate flow
- RO permeate pressure
- RO permeate conductivity
- RO permeate temperature
- RO permeate pH
- RO concentrate flow
- RO concentrate pressure
- Dosing rate of post-treatment chemicals
- Product water turbidity
- Product water free (combined) chlorine
- RO permeate storage tank level

The monitoring activity conducted to protect plant equipment includes monitoring operating parameters of major equipment. This activity includes setting alarms and shut off switches to indicate off limit conditions of the following parameters:

- Levels in water storage tanks
- Levels in chemical storage tanks
- Flow of treatment chemicals
- Water temperature
- Water pH
- Water turbidity
- Free (combined) chlorine concentration
- Pressure drop in cartridge filters
- Pumps suction pressure

- Pumps discharged pressure
- Feed pressure
- Permeate pH
- Permeate conductivity
- Permeate temperature
- Permeate pressure
- Concentrate flow
- Concentrate pressure
- Pressure drop in RO system
- Temperature of electric motors

Both alarms and shut-down procedure will start after some delay time, length of which will be defined during the detail design process.

Some of the alarms will be designed to clear when conditions changes, others may required acknowledgement of operator or plant supervisor.

Instrumentation (sensors and transmitters) being used to monitor and control RO process are indicated by unique symbols in the process and instrumentation diagrams (P&ID). Example of P&ID is shown on Fig. 11.20.

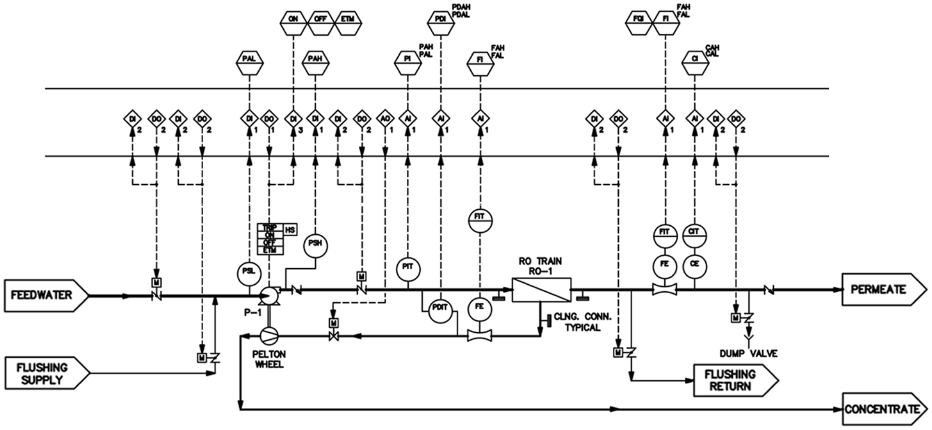


FIG. 11.20 Example of process and flow diagram (P&ID) of an RO unit.

Presently, application of process automation does not provide directly optimization of system performance that would result in achieving optimum product quality or minimum water cost. The only direct cost reduction measures potentially applied currently in RO systems is sequencing operation of desalting units according to predetermined cost parameters. For example system production capacity is utilized according to variable energy cost or priority of units operations is based on prior determined operating cost of individual units (unit required highest operating pressure will be activated last). The above cost reduction measures are applied based on evaluation, conducted off line, of the prevailing economic conditions. Optimization of RO unit (or plant performance) so far has been conducted mainly through manual adjustment of operating parameters in response to change of selective process conditions. For example, with increase of feed salinity the recovery rate is reduced. This is to prevent an increase of feed pressure and energy consumption. Recovery rate could be also reduced or feed pressure increased if lower permeate salinity is required. Algorithms for RO system optimization are seldom available in any standard form. It can be expected that with increasing number of large desalination plants an economic incentive for process optimization will increase and therefore more efforts will be directed towards operation optimization and process algorithms development.

11.2.8.Permeate processing

Product water from desalination system has to comply with local regulations of potable water quality. At some locations additional quality requirements, related to suitability for plants irrigation, are specified as well. This usually implies that all dissolved constituents have to be within a range of allowable concentrations, water can not be corrosive to the distribution piping network, turbidity has to be below defined limit and, depending on location, water should be disinfected or even contain detectable disinfectant residual at the point of distribution. The selection of suitable permeate treatment process will depend on type of membrane desalination technology applied for permeate production. Nanofiltration plants usually produce water that is within potable limits in every respect except for residual presence of a disinfectant. The properties of nanofiltration membranes and/or operating conditions are selected to produce permeate with required level of dissolved solids, alkalinity and hardness. In some instances removal of H_2S is required. Disinfection of product water is applied according to local regulations.

Permeate water produced in brackish or seawater RO systems almost always requires some type of posttreatment to reduce its aggressiveness toward concrete mortar lining of pipes and mitigate its potential to corrode pipe network materials. If permeate water contains hydrogen sulfide it has to be removed immediately at the exit of the RO unit. Availability of water of suitable salinity for blending, provides the simplest and least expensive treatment alternative for permeate water stabilization. Otherwise, the common treatment approach is to partial removal of CO_2 in degasification towers, followed caustic addition and disinfection. Usually, brackish RO permeate has enough residual hardness that conversion of residual CO_2 to bicarbonate using sodium hydroxide or other base, will result in water with positive Langelier Saturation Index.

Presence of H_2S may complicate permeate stabilization process to some extend. The complete removal of H_2S, conducted in packed towers, will result in parallel removal of CO_2. If residual CO_2, left after gas stripping, is too low to produce sufficient alkalinity, bicarbonate has to be added to permeate in form of sodium bicarbonate or soda ash. Disinfection can be conducted by chlorination, formation of chloramines, chlorine dioxide, ozone addition or UV radiation. Chlorine can be applied in form of a chlorine gas, liquid or hypochlorite solution. Sodium hypochlorite can be generated on site, but hypochlorite generation system is usually more expensive than gas or liquid chlorine installation (64). The selection of disinfecting compound or procedure will be mostly driven by local requirements and cost considerations. Permeate produced in seawater systems does not contain H_2S and CO_2 concentration is very low, at the level of few ppm (53, 214). In addition, hardness concentration is negligible.

Permeate stabilization requires increase of alkalinity and hardness (in form of Ca ion). A typical configuration of seawater RO permeate treatment system is shown in Fig. 11.21. A lime slurry is added to the permeate stream, which flows through static mixer to the lime contact vessel. Carbon dioxide is injected to react with lime to form alkalinity. At the outlet from the lime reactor small quantity of sulfuric acid could be added to dissolve residual lime suspension and reduce turbidity. At the end chlorine gas is injected or sodium hypochlorite solution is added as a disinfectant. A simplified alternative of seawater product water stabilization is adding CO_2 and pumping the permeate trough a pressure vessel contain crushed calcium carbonate (limestone). Dissolved CO_2 reacts with calcium carbonate forming calcium bicarbonate which is sufficiently soluble to increase calcium and alkalinity concentration in permeate to the required level. Out of two methods of seawater RO permeate stabilization mentioned above, in large commercial systems the first method is the preferred one, as providing more reliable results (64).

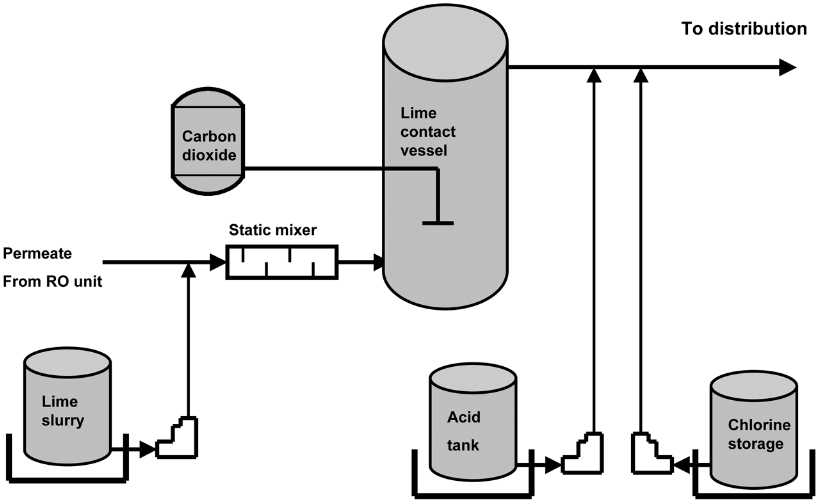


FIG. 11.21 Configuration of permeate treatment unit in RO seawater plant.

One of the usual targets of permeate treatment process is to adjust its composition to maintain positive value of Langelier Saturation Index (see chapter 10.4 Sparingly soluble constituents). Authors of one of the recent publication (124) are making arguments for use of Calcium Carbonate Precipitation Potential (CCPP) rather than LSI for prediction of permeate water stability. They are considering CCPP as a better indicator for prediction of calcium carbonate propensity for deposition in the water distribution piping network.

The CCPP is defined as:

$$\text{CCPP} = 50000\,(\text{TALK}_i - \text{TALK}_{eq}) \qquad (70)$$

where TALK_i is initial total carbonate alkalinity
TALK_{eq} is total carbonate alkalinity at equilibrium
CCPP and ALK are expressed as $CaCO_3$

The recommended range of CCPP is 4–10 as ppm of $CaCO_3$

Manual calculation of the CCPP index is quite difficult and involves complex computation. Recently released by American Water Works Association, spreadsheet based, computer program (125) enables more convenient approach to calculation of CCPP.

If one of the objectives of operation of desalination plant it to supply water for crops irrigation then the usual water quality requirements, included in project specifications, are sodium absorption ratio (SAR) and/or concentration of boron in the permeate. The SAR value is related to the effect of water composition on soil permeability.

The SAR is defined as:

$$\mathrm{SAR} = [\mathrm{Na}^+]/(0.5\ ([\mathrm{Ca}^{++}]+[\mathrm{Mg}^{++}]))^{\wedge 0.5}. \tag{71}$$

Value of SAR above 10 is being considered as having undesirable effect on soil permeability (48). Due to high rejection of calcium and magnesium by RO membranes, permeate produced in seawater RO systems will have SAR value significantly higher than 10 and its calcium concentration will have to be increased in the post treatment system to reduce it. As mentioned already, permeate stabilization results in increase of permeate salinity. This effect has to be considered in the overall system design. Boron, at concentration higher than 0.5 ppm, will affect appearance of citrus fruits. Boron concentration in permeate is usually a concern in seawater RO systems only. Producing permeate with boron concentration below 0.5 ppm has to be addressed through system configuration. The design alternatives of RO seawater systems configurations required for stringent specifications of boron in permeate are discussed in the next chapter.

11.3. Special design cases

11.3.1.Achieving low boron limits with seawater RO technology

Boron is poorly rejected by the RO membranes because boric acid molecule is relatively small, uncharged and non-polar. At elevated pH, however, ionization fraction of boric acid increases and the rejection of boron increases accordingly. Phenomenon of increased boron rejection at elevated pH has been utilized in the RO systems processing low salinity water. In particular in semiconductor industry, if additional boron reduction is required, a two pass system with pH adjustment of feed to the second pass RO is practiced. For boron reduction in low salinity waters, a HERO (57) process has been developed. This process is applicable to water that has sufficient alkalinity that enables almost absolute reduction of hardness using week acid cation ion exchange step. After hardness reduction, RO processing of feed water at relatively high pH without danger of scaling is possible.

In seawater desalination processes, where achieving relatively stringent, low target of boron concentration in permeate has been challenging, a two-pass system configuration has been often employed. In such a design, the permeate produced by a seawater RO system, which operates at low or neutral feed pH, is processed again by a brackish RO unit operating at elevated feed pH. Unfortunately, a simple two pass configuration have number of disadvantages such as low overall recovery, potential for scaling of the second pass RO unit at high feed pH and inability to produce water with constant quality in terms of low boron concentration at high ambient temperatures. Stringent boron specifications in a number of large seawater systems being built recently resulted in development of new design configurations. These options include multistage RO processing of 1st pass permeate and incorporation of boron specific ion exchange units.

Chemistry of boron. Chemistry of boron species in water solution has to be considered when evaluating process design options. It affects the cost of the desalination process, both the cost of equipment and operation.

Boric acid is a very weak acid in water solution. Its ionization equilibrium is represented as:

$$B(OH)_3 + H_2O \leftrightarrow H^+ + B(OH)_4^- \tag{71}$$

$$K_a = [H] \times [B(OH)_4^-]/[B(OH)_3] \tag{72}$$

The value of equilibrium constant, K_a, depends on temperature and ionic strength. Note that ionic strength is a function of water salinity. The value of $pK_a = (-\log K_a)$ ranges from 8.4 to 9.5 depending on ionic strength and temperature (58). Fig. 11.22 shows the values of pK_a as a function of temperature for a typical low salinity water, a seawater feed and a seawater concentrate. The

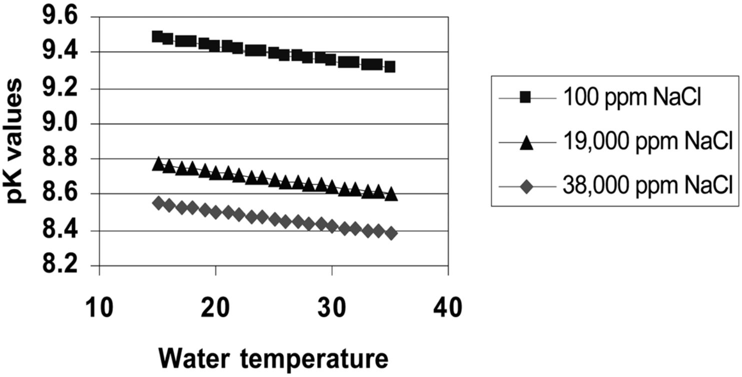


FIG. 11.22 Dissociation constant (pK) of boric acid as a function of temperature and salinity.

equilibrium between boric acid and borate ion shifts toward lower values with increasing ionic strength of solution. The practical importance of this relationship is that at given feed pH, higher fraction of boric acid will be dissociated (Fig. 11.23) in solution of a higher ionic strength.

Figure 11.24 shows nominal rejection of boron estimated based on the assumption of 99% rejection of $B(OH)_4^-$ and 90% rejection of $B(OH)_3$. Note that the boron rejection will depend not only on the overall solute permeability of the RO membrane but also on the characteristics of the membrane surface, which has not been included in this simple prediction. Furthermore, the ionization fraction also increases with temperature (Fig. 11.25). However, because the temperature effect on boron permeability through the membrane is stronger than the corresponding increase of ionization rate, boron rejection will decrease as the feed water temperature increases, as shown in Fig. 11.26.

Boron rejection by seawater and brackish RO membranes. In the past, specifications of boron concentration in the RO permeate have not been considered as a requirement of RO processes in seawater desalination projects. Therefore, limited efforts have been devoted to evaluate boron rejection by seawater RO membranes. One of the few published results includes an investigation conducted by Nitto Denko as a preparation for the operation of the 40,000 m^3/day seawater RO plant at Okinawa, Japan (43). A 80 m^3/day pilot unit, consisting of commercial size membrane elements, was operated at Sesoko Island in southern Japan (59). The results indicated that in a seawater system, in order to produce permeate with boron level consistently below 1 ppm, at least a two pass RO

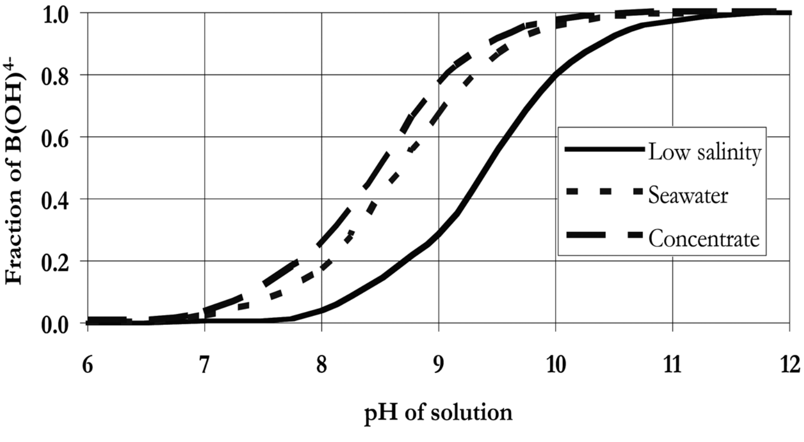


FIG. 11.23 Distribution of boron species vs. feed water pH and salinity at 25°C.

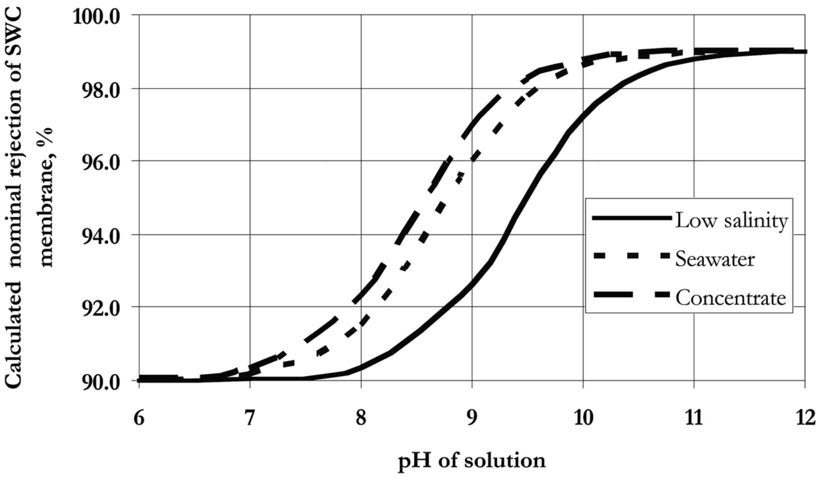


FIG. 11.24 Calculated rejection of boron species vs. feed water pH and salinity at 25°C.

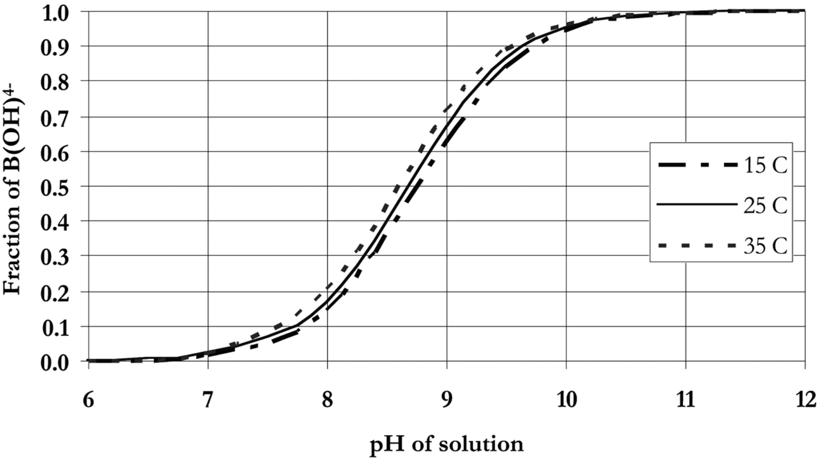


FIG. 11.25 Distribution of boron species in seawater vs. feed water pH and temperature.

configuration would be necessary. The results from a pilot testing were subsequently confirmed during operation of the commercial plant at Okinawa, where membrane elements manufactured by Nitto Denko and Toray were operating in parallel trains.

The boron rejections of previous generation, regular seawater membrane elements, tested at nominal test conditions ranged from 85–90%. This corresponded

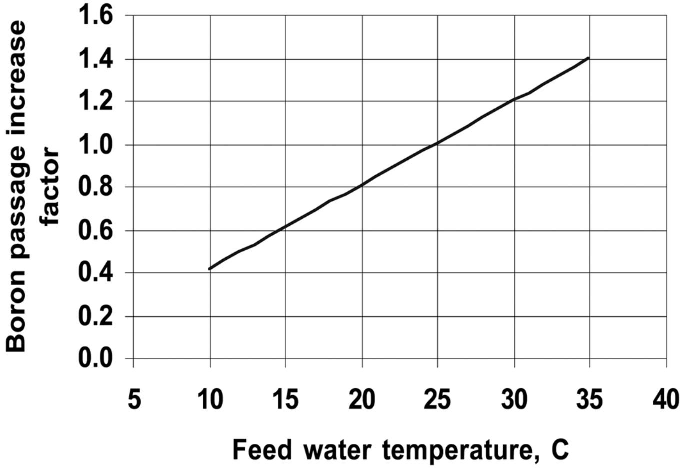


FIG. 11.26 Effect of temperature on boron passage through RO membranes.

to about 78–80% boron rejection relative to an average feed concentration in commercial seawater systems. A subsequent R&D program led to development of commercial seawater membranes with improved boron rejection of 92–94%. This level of nominal boron rejection corresponds to 88–90% boron rejection under the operating conditions of commercial seawater RO plant. At comparable operating conditions, boron passage through brackish membranes is typically higher than that through seawater membranes.

The boron rejection of brackish membranes at nominal test conditions is in the range of 65–75%. Fig. 11.27 displays boron rejection of representative brackish (membranes 1 and 2) and seawater commercial membranes (membranes 3 and 4) at nominal test conditions (pH = 7–7.5, recovery rate = 15%). The higher boron rejection brackish membranes usually have lower permeability and have to operate at higher feed pressure then the regular membranes. However, in applications with stringent boron limits, better process reliability provided by higher boron rejection of special membranes usually provides compensation over somewhat higher energy consumption.

Boron reduction in seawater RO systems. The first large-scale RO seawater project that included boron limit in permeate specifications was the RO plant in Dhekelia, Cyprus (60). It employed hollow fiber membrane elements made by DuPont. The plant capacity was 40,000 m^3/d (10.5 mgd) and commenced operation in April 1997. The boron limit in permeate was specified as less than 1 ppm. The second large seawater plant, with an objective to reduce boron, was a

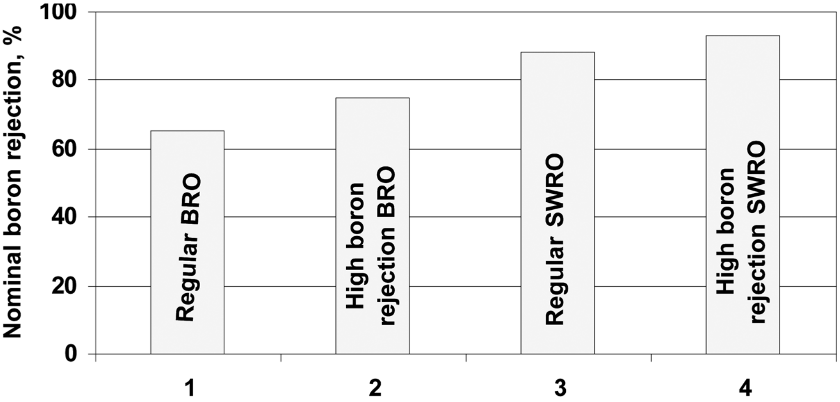


FIG. 11.27 Nominal boron rejection of commercial polyamide composite membranes.

54,000 m^3/d (14.2 mgd) plant at Larnaca, Cyprus. The permeate quality requirement was the same as for the Dhekelia plant, i.e., less than 1 ppm of boron concentration in the permeate. The Larnaca plant commenced operation in May 2001. It operates in a partial two pass configuration. The second pass processes up to 15% of the first pass permeate at feed pH 10. The Larnaca plant utilizes spiral wound membrane elements made by Hydranautics. The reported levels of boron in the seawater at the Larnaca site are in the range of 4–6.5 ppm. Following the two first RO plants with specific permeate boron requirements additional large plants came on line with even more stringent boron levels. Currently the largest plant is the one in Ashkelon, Israel. Its permeate capacity is close to 330,000 m^3/d (87 mgd) and requirement of boron concentration in permeate is 0.4 ppm. The Ashkelon plant utilizes spiral wound membrane elements made by Dow.

RO system configurations for born rejection. In order to improve economics of desalination systems with stringent limits on boron concentration in permeate, process designers have optimized the traditional configuration of RO seawater systems. The most common approaches, which are being implemented in commercial systems, are discussed below.

Single pass system. In a single pass system configuration a permeate water, after permeation through a single RO membrane barrier, leaves the RO system without additional membrane processing. For a majority of seawater sources, the RO unit would be designed to operate at recovery rate of 40–50% and permeate flux of 12–15 l/m^2/h (7–9 gfd). Feed pH in seawater RO systems ranges from acidified feed (pH = 6–7.5) to natural pH of the seawater (7.8–8.2). At this operating conditions, a RO unit equipped with commercial seawater membranes can produce permeate with salinity within potable limits (< 500 ppm TDS). However, it will not always be capable to maintain boron concentration below 1 ppm. The boron passage is a strong function of feed water temperature and condition of the membrane surface. In many recent RO seawater projects, the required permeate concentration of boron is in the range of 0.4–1.0 ppm. This level of boron can not be achieved reliably in a single pass design, especially at elevated feed water temperatures.

Two pass system. The two pass system option is shown in Fig. 11.28. It includes seawater RO unit (RO1) followed by additional RO units (RO2) equipped with brackish membranes. The RO1 operates at recovery rate of 40–50% and the RO1 permeate is processed by RO2 at recovery rate of 85–90%. During the period of low feed water temperature or low boron concentration in the seawater feed, it is sufficient to process only parts of RO1 permeates with RO2 to produce required boron level in a combined permeate. Sodium hydroxide is typically added to feed water to RO2 to increase feed pH to 9.5–10. Consequently, boron rejection by the brackish membranes in RO2 increases to the range of 80–95%. The two pass system can reliably achieve boron reduction down to the level of less than 0.5 ppm. Since the feed to the RO2 process is the permeate water from RO1, the RO2 unit can operate at very high flux rate (about 34 $l/m^2/h$, 20 gfd). Therefore, RO2 consists of relatively small number of membrane elements to process the required flow, as compared to first-pass seawater (RO1). Consequently, the additional capital cost of second pass processing is not very high.

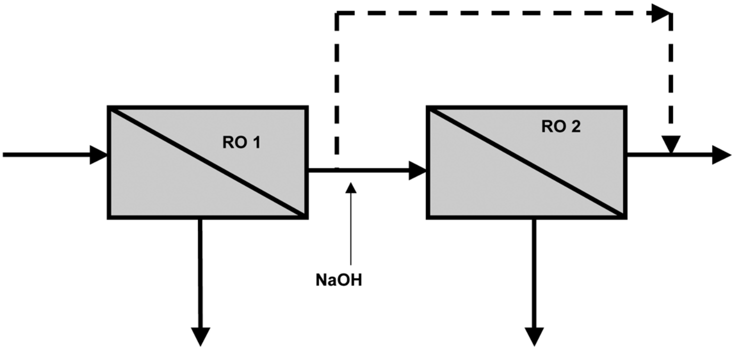


FIG. 11.28 Configuration alternative for boron reduction: two pass RO system with inter-pass pH adjustment.

The problem associated with the two pass configuration is related to an increased of overall operating cost due to reduced overall system recovery. In the RO2 unit, part of the feed water is further discharged as concentrate. Therefore, the combined quantity of permeate produced by the overall RO system is reduced by 10–15%. Therefore, all operating cost components associated with RO1 have to be increased according to the above ratio in estimation of the final cost of operation. Another problem is the potential for scale formation in the RO2 unit. The RO1 permeate contains same small concentration of alkalinity, calcium and magnesium. Scale inhibitor is usually added to the RO2 feed to prevent scaling by calcium carbonate or magnesium hydroxide under elevated pH conditions. Initially, at the beginning of operation, when membrane elements are new, the scaling tendency in RO2 is low since initial rejection of calcium and magnesium by RO1 is relatively high. However, as operation proceeds, passage of all ions including calcium and magnesium tends to increase, resulting in increased concentration in RO1 permeate. Elevated concentration of calcium and magnesium may limit the upper value of pH in the feed to the RO2 and therefore will result in lower boron rejection. Alternatively, recovery rate of RO2 needs to be reduced.

Single pass with boron specific ion exchange. System configuration of single pass RO seawater unit followed by boron specific ion exchange unit is shown in Fig. 11.29. This configuration does not require pH adjustment of the first pass (RO1) permeate stream. The ion exchange unit produces effluent with

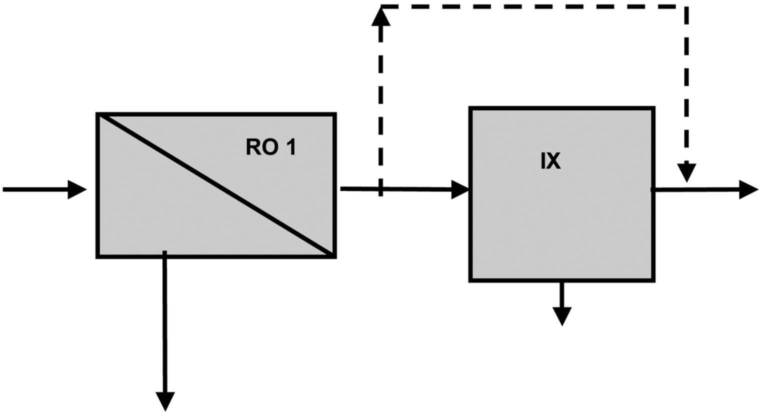


FIG. 11.29 Configuration alternative for boron reduction: single pass RO system with ion exchange polishing unit.

very low boron concentration (~0.1 ppm) and recovery rate of ion exchange process is typically very high, close to 98%. Resin regeneration is conducted with acid, followed by sodium hydroxide flush (to convert resin to a sodium form). Boron removal capacity of boron specific ion exchange resin is about 1–1.3 ppm of boron per liter of resin (28–36 ppm/ft^3). For resin regeneration about 0.08 kg of H_2SO_4 and 0.05 kg of NaOH per m^3 of process water (0.7 lb and 0.4 lb per kgallon respectively) is required. This ion exchange process is very reliable. The only disadvantage of ion exchange unit is high equipment cost, resulting from high cost of this specialty resin and also high cost of chemicals required for ion exchange regeneration. So far there is no large seawater RO system that utilizes ion exchange for additional boron removal. However, if the price of ion exchange resin will decrease, it is possible that this technology could become competitive with the current solution of second pass RO operating at high feed pH.

Multi-pass systems. The problems of low recovery rate and risk of scale formation in two pass RO systems and concerns of high capital cost in systems employing born specific ion exchange step might be effectively addressed by a multi-pass system design. Two variations of multi-pass configuration have been developed and applied in full-scale RO seawater systems (Figs. 11.30, 11.31). In both configurations, additional RO unit is employed to process and recover concentrate from the second pass RO. One approach, shown in Fig. 11.30, consists of treating concentrate from the second pass RO with a softener. After soft-

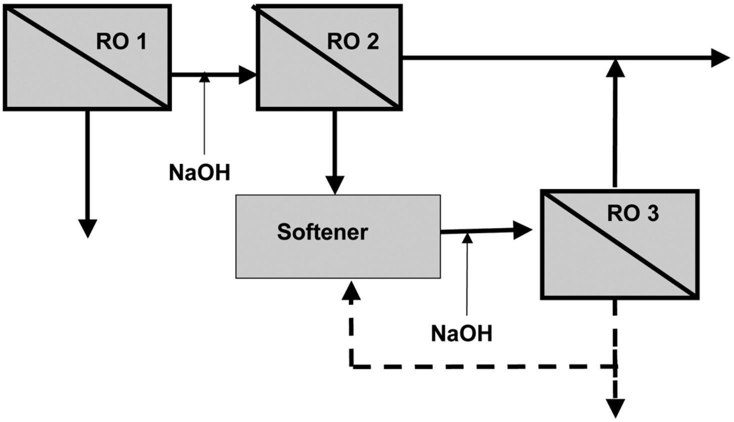


FIG. 11.30 Configuration alternative for boron reduction: two pass RO system with concentrate processing (A).

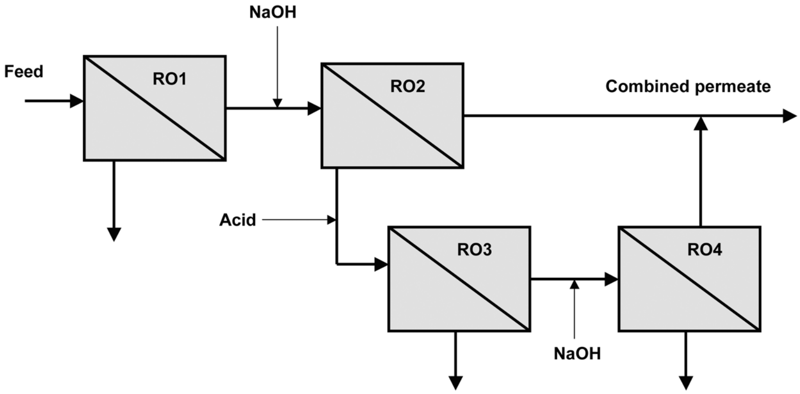


FIG. 11.31 Configuration alternative for boron reduction: two pass RO system with concentrate processing (B).

ening, the effluent from the softener, after adjusting pH to 10–10.5, is further processed with RO3. Due to very low concentrations of calcium and magnesium in the feed water, there is no risk of scale formation in RO3 at such high pH. The concentrate from RO3 is practically pure NaCl solution. Therefore, this stream can be used for regeneration of the brine softener, and reduce cost of operation. This system design is being used in 136,000 m^3/d (36 mgd) RO seawater system in Palmahim, Israel, which has a boron limit in permeate of 0.4 ppm.

Another cost effective design approach, mentioned already, is shown in Fig. 11.31. In this configuration, the concentrate from the second pass RO (RO2) is treated with another RO unit (RO3). To prevent scale formation, pH of the RO2 concentrate is reduced by acidification, prior to processing by RO3. The permeate from RO3, which has very low concentration of calcium or magnesium, is further processed with next RO unit (RO4) at elevated pH. This configuration has been implemented in a 330,000 m^3/day (87 mgd) RO seawater plant in Ashkelon, Israel, which has the same 0.4 ppm boron limit as the Palmahim plant. Even though the multi-pass configuration is more complicated than the two pass design, the multi-pass system produces permeate water at lower cost then the former design options, mainly due to overall higher recovery rate.

Two pass system with pH adjustment of seawater feed. The permeate of first pass RO typically contains a relatively low concentration of calcium and magnesium. Therefore scaling might be avoided within the range of pH required for improvement of boron rejection. According to data presented in Figs. 11.22 and 11.23, the pH required for 50% ionization of boric acid is approximately

9.5 for low salinity waters. However, for a high salinity solution corresponding to seawater feed and RO concentrate, at recovery rate of 45–50%, the pH required for the same ionization fraction is about one pH unit lower, approximately 8.5. Therefore, in seawater systems a meaningful improvement of boron rejection can be achieved at a lower feed pH than the pH required for similar boron rejection when treating a low salinity feed. Process configuration shown in Fig. 11.32 is based on this design approach. It consists of primary and secondary RO units. Caustic soda is added to feed water to primary (seawater) RO to increase feed pH to about 8.4–8.6. The secondary RO also operates at elevated feed pH. However, because the boron rejection of seawater RO has been improved the boron concentration in permeate is lower and secondary RO can operate at lower feed pH then would be required with the previous two pass design (shown in Fig. 11.28) and achieve the same boron rejection.

In addition to simplicity of the overall process configuration, this process option provides the following advantages:

1. Operation of first pass at elevated pH reduces concentration of CO_2 in the seawater feed water and subsequently reduces alkalinity in the feed to second pass RO. Therefore, scaling tendency of the feed water to second RO is lower and RO2 can be operated at higher pH or higher recovery rate with lower risk of scale formation
2. Due to lower alkalinity in feed water to the second pass RO, quantity of caustic required to increase feed pH is lower. This should be compared with additional cost for NaOH dising to the feed of first pass

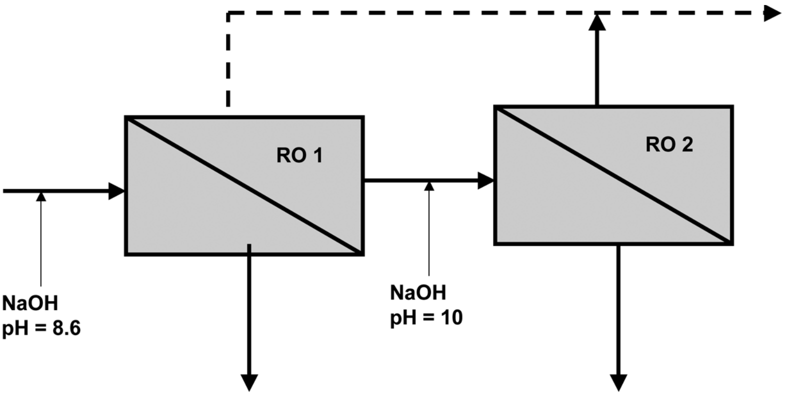


FIG. 11.32 Boron reduction options: two pass RO system with 1st pass And 2nd pass feed pH adjustment.

RO. Common seawater has relatively low alkalinity in the range of 140–160 ppm. The quantity of NaOH required to increase sweater pH to 8.4–8.6 range is about 20 ppm (as 100% concentration).

3. The method of increasing feed pH to the first pass is a flexible option of improving boron rejection. It can be applied when necessary, for example during the period of high feed temperature or elevated boron concentration, without need for additional equipment investment.

All of the above advantages should translate into improved of process economics.

11.3.2.Nitrate reduction with brackish RO membranes

Nitrate contamination of potable water is usually a result of agriculture run off water infiltrating potable water aquifer. The methods available for reduction of nitrate level are dilution with other water supply sources, use of nitrate specific ion exchange resins, electro-dialysis or reverse osmosis technology. Rejection of nitrate ion by RO membranes is lower than chloride or sodium, but still sufficient to produce potable water quality from any common, nitrate contaminated, water source. There is a significant number of RO plant operating with primary objectives of nitrate reduction in potable supply sources. Majority of these plants are in inland location and one of major issues, affecting process economics, is disposal of concentrate. Flow rate of concentrate is function of the recovery rate, which in most cases is determined by concentration of sparingly soluble salts. The common salts of concern are silica and calcium sulfate. Both of very limited solubility and require effective scale inhibitors for efficient, high recovery, operation of the RO unit.

One of the first plants in California that was built specifically for reduction of nitrate level in potable water aquifer is the Arlington Desalter. Recovery rate of the Arlington plant is limited by presence of silica in the raw water which is about 40 ppm.

Arlington Basin Desalter. The Southern California Arlington Basin contains approximately 370 Mm^3 (300,000 acre feet) of water. The ground water in this basin has degraded by agricultural leachate from historic citrus grove farming operations. The agricultural drainage has increased salt concentration to a level that this ground water is no longer usable for domestic purposes. Because of poor ground water quality, all pumping has been discontinued in the area. This resulted in impaired ground water seeping to the surface and draining into the Santa Ana River and other adjacent ground water basins, thereby degrading downstream water supplies.

The Arlington Basin Desalting project initiated, as a solution to this problem, serves the following purposes:

a) Reduce salts entering into the Santa Ana River.

b) Provide a clean water supply.

c) Restore the ground water in the Arlington Basin to a usable condition.

d) Restore the condition of the Arlington Basin for future water storage.

The Arlington reverse osmosis plant was designed by Camp Dresser & McKee of Ontario, California. Hydranautics furnished, installed and commissioned the RO system. The RO system was designed to provide 23,000 m^3/day (6 MGD) of blended product water containing less than 500 ppm TDS by mixing 15,000 m^3/day (4 MGD) of degasified permeate from the three RO trains (each rated at 5,000 m^3/day, 1.33 MGD) with 8,000 m^3/day (2 MGD) of ground water treated by granular activated carbon followed by aeration stripping. The RO system operates at a permeate recovery rate of 77%. The RO plant is designed to allow for expansion up to 30,000 m^3/day (8 MGD), by addition of a fourth RO train. The plant data is summarized in Table 11.10.

Process description. The schematic process flow diagram of the Arlington Desalter is given in Fig. 11.33 and plant basic information listed in Table 11.10. Feed water from the five local brackish wells is pumped to the plant site where it is split into two streams. Out of the total raw water flow of 26,500 m^3/d (7 MGD), provision exists for passing 8,000 m^3/d (2 MGD) through Granular Activated Carbon (GAC) filters, to remove dissolved organic compounds, mainly dibromochloropropane (DBCP). At present, due to lower than expected concentration of DBCP in the ground water, the GAC filters are bypassed and the blend stream is only treated by aeration stripping. The remaining flow, 19,000 m^3/d (5 MGD), is used as feed for the RO system. The RO Feed water is treated by dosing of scale inhibitor and sulfuric acid to a pH of 6.9 and is filtered through 5 micron cartridge filters.

After the filtration feed water is pressurized to approximately 14.5 bar (210 psi) with Afton vertical turbine pumps, the pressurized feed enters three parallel RO trains operating at 77% permeate recovery. Each train contains 44 pressure vessels, 8″ diameter, in a two pass 33:11 array. The pressure vessels each contain six Hydranautics spiral wound, composite, membrane elements, model 8040-LSY-

TABLE 11.10
Arlington Desalter Plant Data

Start up:	July 1990
Commissioning:	September 1990
Plant capacity	22700 m^3/d (6 MGD)
RO permeate	15100 m^3/d (4 MGD)
GAC effluent (blend water)	7600 m^3/d (2 MGD)
Number of RO trains	3
Array	33:11
Number of elements per train	264
Element type	8040-LSY-CPA2
Permeate recovery	77%
Feed Water type	Well water: 1200 ppm TDS, 90–100 ppm NO_3 44 ppm SiO_2
Product water quality	<500 ppm TDS
Requirement, after blending	40 ppm NO_3
Pretreatment	pH adjustment, scale inhibitor dosing, cartridge filtration
Design feed pressure	11–14 bar (160–205 psi)
High pressure pump energy consumption	0.34 kWh/m^3 (1.3 kWh/kgal)

(1) Pump equipped with Energy Recovery Turbine

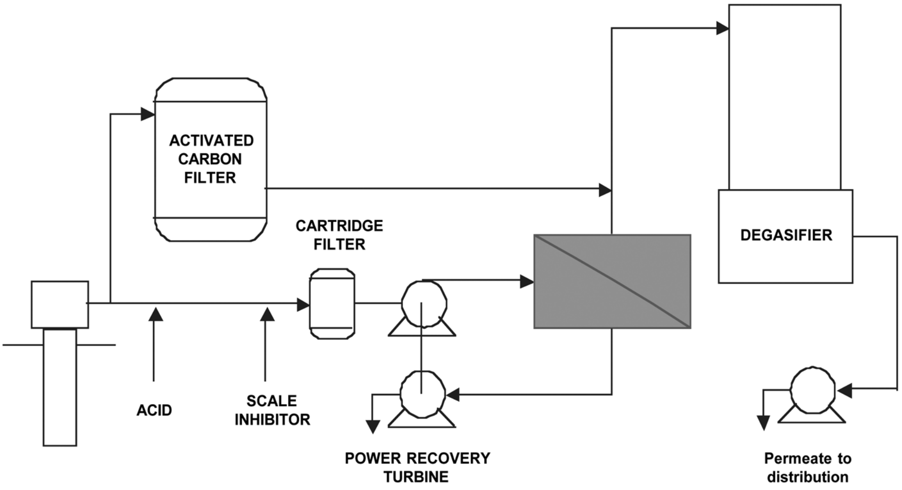


FIG. 11.33 Flow diagram of brackish RO plant for nitrate reduction and Arlington, CA.

CPA2. The average water flux rate of the membranes is 23.5 $l/m^2/h$ (13.8 gfd). The first pass averages 22.5 $l/m^2/h$ (15 gfd) and the second pass averages 18.7 $l/m^2/h$ (11 gfd). Permeate flow from the RO trains is combined with the blend stream at the ratio 2:1. The design blend ratio was based on the projected concentration of nitrate in the wells and in the permeate water, with a target concentration corresponding to California drinking water standard of not more than 45 ppm of nitrate in the total plant effluent. This blended effluent is of potable water quality and flows to the storm water channel and eventually recharges the ground water basin.

The concentrate stream from each RO train passes through an energy recovery turbine, which is a reverse running pump mechanically coupled with the high pressure pump. The combined concentrate from the plant is conveyed to the Orange County Sanitation District through the Santa Ana Regional Interceptor (SARI) line. After mixing with municipal sewage, and primary and secondary treatment of the Sanitation District effluent is split for further treatment by OCWD Water Factory 21, or direct disposal to the ocean. The flow diagram for the Arlington Basin Desalter Plant is shown in Fig. 11.33. The RO unit incorporates permeate throttling as a way to control permeate flow. The objective of using permeate throttling rather than conventional feed throttling to control productivity is to eliminate loss of recoverable energy and maintain constant concentrate pressure to the energy recovery turbine.

The other features are related to prevention of membrane scaling. One serious concern during system design was the projected high level of silica (40 ppm) in the ground water. It was customary in RO technology to accept concentration of approximately 120-160 ppm of SiO_2 as a limit for safe operation in the concentrate. The kinetics of silica scaling are not well understood and at the time of system design there was no commercially available silica scale inhibitor. The initial requirement of permeate recovery of 80% would result in concentration of SiO_2 in the concentrate of about 200 ppm. Calculation, element by element, of silica concentration indicated that the customary limit of 160 ppm of SiO_2 would be exceeded at the 5th element of the second pass. From this point, the brine stream residence time in the system is about 10 seconds. It was estimated that during such short residence time, at a temperature of 21°C, no silica scale should be formed, especially considering the quality of the raw water which has very low concentration of suspended solids, which could act as crystallization centers. Even though, at this condition, no silica precipitation was expected, as an additional safety feature, each RO train was equipped with a 4-inch diameter monitor tube. The monitor tube contains a single 4-inch element and operates on the concentrate stream from the RO train as feed water. The assumption was that silica scaling, if any, would start in this monitor tube, giving early warning to change the operating conditions.

Plant operation. The RO system at the Arlington Desalter was commissioned in September 1990. The composition of feed and permeate as analyzed from samples taken during acceptance test are presented in Table 11.11.

TABLE 11.11
Water analysis of water samples taken during acceptance test at the Arlington pant.

Constituent	Feed water, ppm	RO permeate, ppm	Salt passage, %	Blended water, ppm
Calcium	140	0.3	0.22	46
Magnesium	42	0.01	0.24	15
Sodium	168	7.1	4.21	66
Potassium	3.8	0.2	5.22	1.5
Bicarbonate	367	8.4	2.72	140
Chloride	162	2.3	1.44	60
Sulfate	243	0.6	0.21	89
Nitrate	93	8.3	8.96	37
Fluoride	0.4	0.1	24	0.2
Silica	40	0.1	.25	15
TDS	1260	28	2.21	460

12

System design verification through operation of a pilot unit

Pilot unit operation is an essential part of the membrane plant design process. Pilot unit operation can serve number of objectives:

1. Verification of product or process performance
2. Optimization of process operating parameters.
3. Verification of process economics

Verification of product or process performance is usually conducted if site conditions are unique in respect of feed water quality or if required performances are deviating from regular results obtained with given type of product or process. In some instances objective of pilot unit operation could be development of process parameters and verification if required performance are possible. Common examples of pilot unit operation are verification of level of performance of membrane elements and stability of operational parameters in field conditions. Another common pilot operation task is optimization of system configuration and operating parameters of feed water pretreatment process. For a well established application type, such as for example treatment of brackish well water, pilot operation is seldom conducted. The exception would be brackish or NF projects that require unique rejection characteristic from membrane elements, for which very limited data is available. Regardless of water source, operation of pilot unit is quite common if water to be treated has high fouling potential. Quite often, operation of pilot unit serves as an initial screening of membrane elements for a given project. In some cases operation of pilot unit could be part of scope of supply. Then, the pilot unit operation is required to demonstrate that process or product proposed by OEM is capable to perform according to project specifications. Pilot unit can be configured to include only feed water pretreatment section or it could represent a complete system.

Procurement of pilot unit could be quite expensive. Typical cost of pilot unit ranges between $50,000 and $150,000. Sometimes a leasing of pilot unit is possible. As pilot units are usually operate at sites with limited infrastructure, logistics of operation of pilot unit could be quite difficult and operating cost high. Never the less, operation of pilot unit is quite important as it can provide essential information for process design.

Usually, prior to pilot unit procurement or lease and operation, initial testing is conducted in laboratory using bench scale equipment to determine range of initial operating parameters. Initial tests could include operation of flat membrane test cells with representative feed solution to determine rejection rate of specific constituents. Another common initial test is a "Jar test" conducted to determine initial dosing rate of coagulants and polymers to be used in media filtration systems. Jar test procedure involves flocculation experiments of different dosing quantity and types of coagulants, conducted in parallel beakers under well defined mixing conditions (39). The usual approach is to build pilot unit in a way that its configuration and operating parameters mimics the future commercial plant. This applies both to the pretreatment and membrane sections of the pilot. The common pretreatment equipment tested will be either media or membrane filtration. Sometimes, number of water treatment technologies are combined together, either in series or in parallel configuration. Series arrangement could include for example dissolved air flotation followed by media filtration or media filtration followed by membrane filtration (MF) or UF. Parallel configuration could consist of two or more parallel lines of different pretreatment equipment followed by NF or RO units.

The preferred configuration of NF or RO pilot unit is that the recovery rate and average flux rate of the pilot unit is the same as the future commercial unit. For example the commercial unit is designed to operate at recovery rate of 80%, flux rate of 27.2 $l/m^2/h$ (16 gfd) and the membrane array of the train will be 20:10 pressure vessels, each housing seven elements, of 37 m^2 (400 ft^2) membrane area each. To mimic exactly this plant configuration, pilot unit would require 21 4040 elements in a 2:1 array. A 100 mm (4″) diameter pressure vessels housing seven elements are not very common and most likely not available as an off the shelf item. Therefore, the pilot unit can be configured as 2:1 array with six elements per vessel, total 18 elements. 4040 elements. The 4040 element has 7.8 m^2 (85 ft^2) membrane area. Therefore, pilot unit at 27.2 l/m^2/h will produce 3.81 m^3/h (24,200 gpd). At recovery rate of 80%, feed water flow required will be 3.9 m^3/h (21 gpm). If budget available for pilot study is limited then pilot operation can concentrate on verifying a critical segment(s) of the future commercial process. For example in seawater systems, recovery rate and scaling potential is usually not a concern. The major concern is potential blockage of the feed channels of the elements in the lead position by colloidal particles or bacterial slime. This phenomena can be adequately tested by following pressure drop and permeate flow in a pilot unit containing only 2–3 elements in a single pressure vessel. The size of the RO section of pilot unit will be smaller in this approach. The size of pretreatment unit will depend on feed flow selection. If objective of the pilot unit operation is to compare stability of water permeability of different membrane elements, then a number of elements can be tested in a single unit in parallel operation as shown in Fig. 12.1, which provides schematic flow diagram of pilot unit operating on well water. Different elements (A, B and C) are positioned in the first stage parallel vessels. The type of element(s) installed in the second stage can be changed periodically or maintain the same depending on the results.

The experimental matrix of test parameters is listed in Table 12.1. The tests matrix includes operation at various recovery rates, with and without sand filters as well as effect of two types of scale inhibitors. According to this matrix, test duration is scheduled to be 10 months on line time.

The duration of the pilot operation is dictated mainly by the available budget. If possible, it is beneficial to operate pilot through the whole cycle of seasonal changes of raw water quality. During operation of pilot unit it is important to maintain high on line factor. In addition, test results should be evaluated periodically to determine performance trends. A typical pilot unit operation program should include daily readings of flows, conductivities, pressure, feed

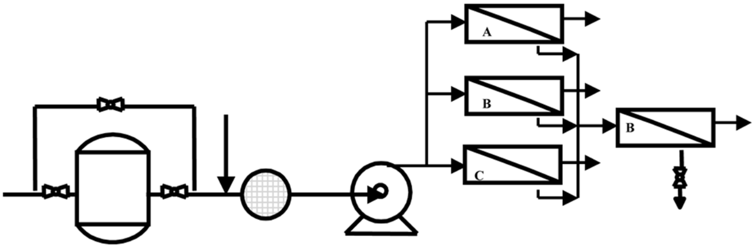


FIG. 12.1 Example of pilot unit configuration.

TABLE 12.1
Experimental matrix of pilot unit operation

Period, months	3	3	3	1
Recovery rate, %	75	80	85	90
Sand filter	–	+	–	–
Cartridge filter	+	+	+	+
Scale inhibitor 1	+	–	+	+/–
Scale inhibitor 2	-	+	–	–/+

water temperature, pH, turbidity, SDI and pump operating hours. It is very common recently, to have operation of pilot unit monitored remotely equipped with sensors providing readings for on line display. According to test program and objectives, water samples for chemical analysis should be taken periodically. Performance should be normalized on regular basis. After completion of each experimental segment a progress report should be prepared. During the last stages of the test program the pilot unit should operate at the conditions that could lead to the membrane fouling. This is to determine limits of operational parameters. In addition, membrane cleaning operation should be conducted during the test program to determine effectiveness of cleaning procedures.

13

System commissioning

System commissioning involves preparation for system startup, testing of functionality of major equipment and initial operation. Initial operation usually includes conductance of an acceptance test, either for internal or external purposes. As a preparation for system startup a plan for system commissioning should be developed. The plan should outline tests procedures that will be conducted prior to system startup, listed their expected results, including pass/fail values and provide details for system startup (schedule, procedures, expected outcome, duration and responsibility). Storage conditions of membrane elements prior to loading should be addressed as well. After completion of the construction phase, as a preparation for system commissioning, all major equipment, including connecting piping should be inspected for debris left from construction. If system include media filters, their internal piping should be tested for leaks prior to loading filtration media. Media filters should be loaded and filtration media should be conditioned according to recommendations of equipment supplier. All piping and pressure vessels should be flushed, disinfected with chlorine and flushed again to remove residual chlorine. Special attention should be paid to any FRP tanks or piping in the system. The FRP components should be sufficiently cured prior to feed water being stored or pumped through. Chemicals, leaching from FRP components, may cause severe decline of membrane permeability. Also all FRP components should be painted with completely opaque piping to prevent any light translucency. Even slight translucency could result in algae and bacterial grow in the system.

As a preparation for membrane elements loading, operating conditions of pretreatment system should be optimized. This applies mainly to coagulation and flocculation parameters. Equipment and the whole plant operation should be tested at the representative operating conditions of flow and pressure, prior to loading membrane elements. For hydraulic testing of the RO trains, orifice plugs should be placed in the permeate ports of pressure vessels. The orifice diameter should be such that feed pressure will be reduced to an atmospheric pressure at the rated permeate flow. Operation with orifices simulates normal operation of the whole plant. Membrane elements should be loaded into pressure vessels under supervision and in accordance to loading procedure provided by membrane manufacturer. It is recommend to record serial numbers of elements as they are loaded and relate these to their position in pressure vessels. This information is very useful during maintenance procedures, including future element replacement. It is very important that during loading only chemicals that are approved by membrane manufacturer will be used for lubrication of o-rings and brine seals. For example, it is known that use of perolum based lubricants, could result in chemical induced fractures of plastic components of membrane element. Only gliceryne should be used to lubricate pressure vessel for easier loading of membrane elements. Pure silica based grease should be used to lubricate inerconnector o-rings.

As a first step of acceptance test, operation of pretreatment system should be verified again. For such operation of pretreatment, the feed piping has to have provision for diverting pretreatment effluent to outfall without passing through RO membrane trains. Membrane operation should only start after feed water quality will be maintain consistently within the design limits for a period of few hours. When starting membrane operation for the first time it is very important to vent any air locked in the system, prior to applying high feed pressure. Pumping system should be design for a “soft” startup, i.e., slow increase of feed pressure. Feed pressure should increase slowly, at the rate not exceeding 0.5–0.7 bar (8–10 psi) per sec.

Acceptance test duration and conditions are usually detailed in the project specifications. Common duration of the acceptance test is one to four weeks of interrupted operation. Usually acceptance test procedure specifies what type of shut downs and duration will be allowed, without necessity to repeat the acceptance test. During the acceptance test, performance of major equipment, permeate capacity, permeate quality, feed pressure and sometimes power consumption, use of consumable and cost of operation, are being evaluated.

With completion of the acceptance test, a report is prepared and submitted for evaluation to the engineering firm supervising the project. It is quite common that after acceptance test some correction of the plant equipment has to be made. So called “punch list” contains list of equipment and procedures that require corrections. A definite time limit for completing of corrections is usually specified. Usually with a satisfactory completion of the punch list items, plant is accepted for operation.

At some locations, permeate quality has to be verified by government certified laboratory, prior to allowing transfer of product water to the potable water network. The initial tests of product water quality usually include also tests for presence of pathogens. Sometimes, due to contamination of piping network or contamination introduced during elements loading, permeate water my show positive results for a coliform bacteria test. This contamination, if occurred is at very low level. Newer the less it is required to be eradicated prior to water acceptance. With polyamide membranes strong oxidizing disinfectant can not be used. The usual approach in such a case is to soak the membrane unit with 1–3% solution of sodium bisulfite for 12–24 h. Another method is to expose membrane elements to low salinity water at low pH (pH 4–5) for a similar period of time.

14

System operation

Two most important issues during system operation are operation of RO membranes with feed water of good quality and prevention of structural damage of RO elements. Structual damage could result from mechanical shock ("water hammer") or operation with excessive feed–concentrate pressure drop. In majority of cases well water is of good quality and does not required additional filtration to remove colloidal particles. Surface water practically always requires removal of colloidal matter prior to membrane elements. Pretreatment processes and feed water quality indicators were already discussed before (chapter 8.2.3). However, it is important to realize, that even short exposure to feed water of poor quality may result in contamination of elements, effect of which could be very difficult to reverse. Therefore, parameters of the pretreatment process should be continuously optimized to achieve and maintain good feed water quality. For example, it is quite common that parameters of coagulation (dosing rate, water pH) have to be adjusted to compensate for seasonal changes of raw water quality.

Membrane elements are designed to provide stable performance under operation with high feed pressure. However, strong impacts during conditions of "water hammer" or continuous stress, due to operation with excessive pressure drop, can result in mechanical damage of membrane elements and shortening of useful membrane life.

It is quite common for membrane performance to stabilize during the initial 1–2 weeks of operation. Usually, both permeability and salt passage decline during this period. This demonstrates itself by a need to increase feed pressure in order to maintain permeate capacity and by decline of permeate salinity. This initial decline is more pronounced in seawater and wastewater systems then in brackish or nanofiltration plants. The usual range is about 10% decline from the values at the startup. If possible, the operation of the plant should be scheduled to provide long uninterrupted periods of operation of individual RO trains. Long operating period enables better determination of performance trends.

As mentioned already, on unit start up, feed pressure should be ramped up slowly at a rate not to exceed 0.7 bar (10 psi) per second. In the event of RO system shut down, it is common procedure to flush the system to replace concentrate feed solution. Minimum requirement is to replace 1–2 system volumes with RO permeate or filtrated feed water. System should be configured to activate flush of the RO trains even in the event of shut down due to power failure. This is especially important in brackish water systems. Usually, in brackish systems concentration of sparingly soluble salts is at saturation levels. Scale inhibitor, if added to the feed water, delays precipitation for some limited period of time. Therefore, saturated solution should be flushed out from pressure vessels and all system piping. Otherwise, if left in the RO system for long time, it may precipitate scale forming salts.

In seawater RO systems the danger of scaling from concentrate is small. Still, it is good idea to flush the system with RO permeate. One of the more important reasons for permeate flushing is that periodic presence of low salinity RO permeate on the feed side of the membrane, can mitigate biological fouling process through drastic reduction of osmotic pressure. In seawater systems flushing is sometimes supplement by incorporating a “suck back tank.” The suck back tank is part of train permeate piping and storage system and provides water for direct osmosis (reverse permeate flow) accruing immediately after unit shut down. During this step, there is a flow of permeate from the permeate side of the membrane to the feed side and dilution of concentration polarization layer at the membrane surface. The dilution process, after unit shut down is quite fast, and it comes to equilibrium in a short time of less than a minute.

During flushing of RO train it is important to have the permeate line open. Otherwise, a delamination of membrane in elements, located in the tail part of the system, may occur. Membrane delamination is caused by condition of permeate pressure being higher than the pressure applied on the feed side of the membrane. During flushing or normal operation, due to friction losses, the water pressure at the vessel exit (concentrate pressure) is lower than the water pressure at the entry to the vessel (feed pressure). This difference for a 7–8 elements long vessel can be 0.5–1.5 bar (15–22 psi), with clean elements. At fouling conditions, that cause partial blockage of element feed channels, this pressure difference can be much higher. If the unit permeate outlet is closed during flushing, permeate pressure will eventually equilibrate with feed pressure. Then for elements in tail positions the permeate pressure will be higher than pressure on the feed side of the membrane (concentrate pressure). If this pressure difference is higher than 0.5–1 bar (7–15 psi), it can cause membrane delamination.

Monitoring of system performance and operating conditions is very important task assuring reliable plant operation. In addition to data collected automatically by the control system, plant operators usually take daily readings at a frequency: 1–2 times per shift. These readings should include all operational data required for performance normalization. Performance normalization should be conducted preferably daily, but no less than once per week. Data normalization could be conducted manually (see chapter 7), but in most cases performance normalization is conducted using computer programs provided by membrane manufacturers. Sometimes, performance normalization routines are part of the data collection program, included in the PLC software. Then normalized data are available instantaneously on line, in the same way as the operational data are.

Not always is configuration of normalization programs flexible enough to reflect directly configuration of a RO train. In such instances data from each desalination stage or pass are entered as a separate train. The minimum set of data for performance normalization includes feed temperature, flow of permeate and concentrate, feed, concentrate and permeate pressure, feed and permeate salinity. The results of performance normalization are daily values of normalized permeate flow, salt rejection and pressure drop. These results should be plotted, preferably at daily frequency, to follow up on performance trend. The rate and magnitude of normalized performance changes determines when membrane cleaning should be applied. Membrane performances, in plants treating well water, are usually quite stable and very little changes are observed over time. In desalination plants treating surface water fluctuation of performance, due to variation of feed water temperature and salinity, are more frequent.

In addition to recording salinity of combined permeate and salinity of permeate from each stage, from time to time permeate salinity from individual pressure vessels should be measured. The water samples for salinity determination are usually collected at the sampling panel (see Fig. 5.3). Such panel has single tubing connected to permeate port of each pressure vessel in the train. The tubing should be connected to the pressure vessel permeate port which is connected to train permeate collecting manifold. Otherwise, permeate sample will represent only small fraction of permeate produced by the vessel. If an individual reading of permeate from a given pressure vessel is significantly higher then rest of the readings in the train or stage, it is indicative of an internal leakage.

This leakage can be result of faulty interconnector or inboard o-ring or feed to permeate leak in a bad element. Location of the leak in a pressure vessel can be confirmed by "permeate probing." To conduct permeate probing a small ball valve (about 12 mm, 0.5″ diameter) and special adapter has to be installed on the pressure vessel permeate port that is not connected to the train permeate collecting manifold.

Permeate probing procedure involves of inserting small diameter (~9.5mm, 0.375″) tubing through the valve, up to the end of the vessel. During this procedure the inserted tubing travels inside elements permeate tubes, which are connected to each other and to permeate end adaptors at each end of the vessel. At the point that the probe tubing reached the end of the pressure vessel, permeate sample is collected. This is the position of the end adaptor. Next the probe tubing is pulled 50 cm (20″) out from pressure vessel. This position corresponds to middle of the first element. Next the tubing is pulled out additional 50 cm (20″). This position corresponds to interconnector between two adjacent elements. This process is repeated till the other end of the vessel is reached. At each position permeate sample is collected.

Conductivity of water samples at each position provides conductivity profile that can be compared with calculated values. Conductivity profile depends on performance of membrane element (salt passage and water permeability), its position in the vessel and operating parameters associated with a given array of pressure vessels. Another important parameter is direction of permeate flow relatively to feed flow. Fig. 14.1 shows relative permeate flow along the pressure vessel, containing eight elements, operating in a single stage seawater RO system, treating seawater of salinity about 40,000 ppm TDS, recovery rate 50%. Case A in Fig. 14.1 corresponds to configuration of concurrent flow of permeate and feed. Case B corresponds to opposite directions of flows of permeate and feed.

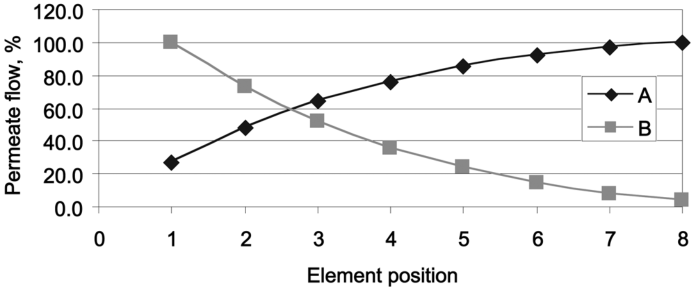


FIG. 14.1 Cumulative permeate flow along pressure vessel vs. direction of permeate flow.

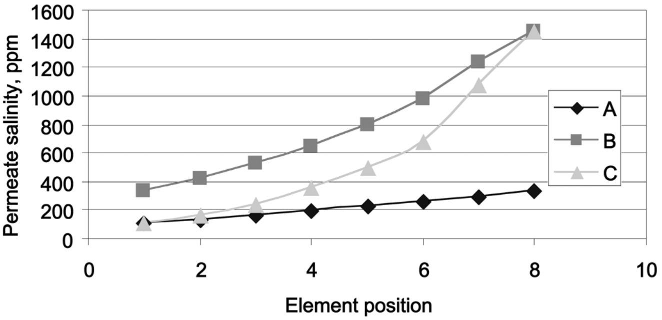


FIG. 14.2 Permeate salinity distribution along pressure vessel vs. direction of permeate flow.

Figure 14.2 shows permeate salinities associated with element positions. Line designated as case C shows permeate salinity contributed by individual elements. Lines for case A and B shows combined permeate salinity and correspond to opposite directions of permeate flow. Case A corresponds to concurrent flow directions of feed and permeate. Case B represents results for countercurrent directions of flows. In the actual field conditions the value measured is conductivity and not TDS.

The corresponding conductivity distribution along the pressure vessel, in concurrent (case A) and countercurrent (case B) flow, is shown in Fig. 14.3.

Although, the permeate conductivity at the outlet from pressure vessel is the same in both cases, the conductivity along the vessel is quite different. Significant deviations of measured conductivity from the calculated values are indicative of position of internal leaks.

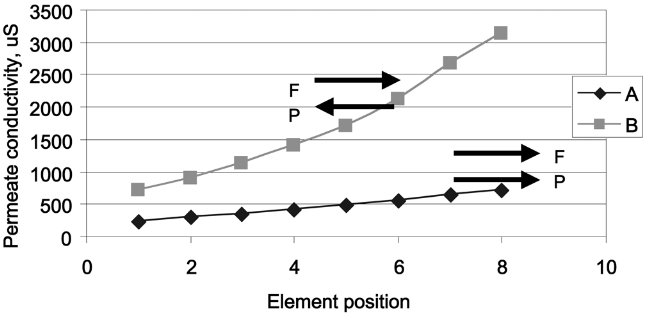


FIG. 14.3 Permeate conductivity distribution along pressure vessel vs. direction of permeate flow.

The results in Figs. 14.1–14.3 were calculated based on assumption that all membrane elements in pressure vessel have the same standard test conditions (nominal) performance of permeate flow and salt passage. In reality the nominal elements flow can wary ±15% and salt passage can wary ±50%. In addition to these variations elements performance could be affected by membrane compaction and fouling phenomena.

It is quite difficult to account for the above variation of performance as the permeability and salt passage of individual elements, after being installed in the RO system, can not be measured directly. At present, the only way to obtain the above information is to remove elements from the vessel and test them individually at some defined test conditions. These results then can be used to calculate projected elements performance at the operating conditions of RO unit.

Inability to obtain direct information on individual elements performance during system operation is a serious issue to be considered when evaluating the feasibility of improving system performance through partial elements replacement. The most efficient way of improving system performance would be to locate and replace membrane elements that are below certain performance level. At present, the only way to implement this is by replacing all elements in a given position in a train or through a staged replacement process. I the first approach an estimation is made regarding the location of elements that most likely are below some level of performance and target them preferentially for replacement. For example in case of scaling deposit (brackish water unit) it is likely that elements positioned at the concentrate end will be most affected. In case of colloidal fouling (surface seawater systems) very likely elements in the lead position will have the highes level of feed channel blockage and highest contribution to excessive pressure drop. In a staged membrane elements replacement approach all elements in one train are replaced with new elements then the old elements are tested individually. Old elements with acceptable performances are used for the next train replacement. For the purposed of individual elements testing, a separate, one element test unit can be used. In a simplest version, such test unit could be just a single element pressure vessel, equipped with flow meters, gauges and disconnecting valves, attached to a train, and operating in parallel to the main unit.

A complete membrane replacement is usually scheduled on the assumption of 5–7 years membrane life. However, in the intermittent period some quantity of membrane elements may require to be replaced due to performance problems (usually salt rejection). A single element test unit is quite useful to help make determination which membrane elements should be replaced. Single element test unit is also very useful tool in determination of the nature and causes of performance changes observed in the RO unit. Such a unit can also help to evaluate effectiveness of cleaning procedures by measuring performance of individual elements before and after cleaning.

15

Membrane elements fouling and performance restoration

15.1. Membrane elements fouling process

Membrane fouling affects membrane performance and can occur with any type of feed water and membrane types. Fouling is even encountered in secondary RO units processing RO permeate. Fouling affect is moderate in its initial stages, accelerating fast, and resulting in significant performance deterioration if not addressed on time. Summary of major fouling phenomena is presented in Table 15.1. Each one is presented as a separate category, however, in field conditions combined fouling processes are occurring either in parallel or one process is precursor to another one. Correction of fouling conditions is a three step process:

1. Early detection of fouling process
2. Identification of fouling conditions and their mitigation
3. Correction of membrane performance

The most effective way of early detection of fouling process is through periodic evaluation of trends of normalized performance: product flow or water permeability, salt passage or salt transport and pressure drop. Any changes of normalized performance values, beyond the initial decline of water and salt transport are indicative of fouling process. The usual process design assumptions are that due to fouling, permeability will decrease and salt passage will increase by 5–10% per year. Performance deterioration at higher rate can only be

TABLE 15.1
Summary of major membrane fouling categories

Fouling factor	Initial fouling stage effect	Advanced fouling stage effect	Potential membrane damage
Exposure to free chlorine	Some permeability and salt passage decline. Initially in the lead element(s)	Increase of permeability and significant increase of salt passage	Irreversible damage of membrane barrier
Colloidal matter	Some increase of pressure drop. Initially in the lead element(s)	Significant increase of pressure drop, some decline of permeability and increase of salt passage	Element telescoping and extrusion of brine spacer
Dissolved natural organic matter (NOM)	Some permeability and salt passage decline	Moderate decline of permeability, same salt passage decline	None
Biological matter	Some increase of pressure drop. Some permeability and salt passage decline. Initially in the lead element(s)	Severe increase of pressure drop. Some permeability and salt passage decline	Element telescoping and extrusion of brine spacer
Inorganic scale	Some increase of pressure drop. Some permeability decline. Initially in the tail element(s)	Severe increase of pressure drop. Some permeability decline and salt passage increase	Severe blockage of feed channels
Petroleum products	Significant permeability decline in the lead element(s). Small effect on salt passage	Severe decrease of permeability. Small effect on salt passage	None at low concentration. At high concentrations barrier integrity damage
Composite foulants (organics + colloids)	Some increase of pressure drop. Some permeability decline. Initially in the lead element(s)	Severe increase of pressure drop. Severe permeability decline and some salt passage increase	Severe blockage of feed channels. Element telescoping

tolerated if additional safety margin was applied during process design. Otherwise, system performance will be outside project specifications.

The identification of fouling conditions is a complex step wise process. It always starts with evaluation of updated composition and quality of the feed water. As mentioned previously, it is quite common that the process design of a new RO plants is based on approximate composition of raw water as feed water sources are usually being developed during system construction. The actual feed water should be analyzed in respect of concentration of scaling constituents (mainly in case of brackish sources) and RO feed water quality indicators such as turbidity, SDI and TOC. In instances that raw water has high fouling potential, useful information about nature of the foulants can be obtained by analyzing deposits on the SDI filter pad using scanning electron microscopy (SEM) and energy dispersive x-ray analysis (EDX). These tests may provide some indication about effectiveness of pretreatment and help to determine what type of foulants are arriving to membrane elements with the feed water. An example of comprehensive analysis of fouled membrane elements is included in reference 137.

Fouling process in RO system is usually identified through results of performance normalization. In brackish water multistage systems more insight into fouling phenomena is obtained through normalization of performance of individual membrane stages rather then normalization of total unit performance. Higher rate of performance decline in the first stage are indicative of foulants arriving with feed water. Typical fouling phenomena observed in the first stage are pressure drop increase due to blockage of the feed channels by colloids or biological growth. Another process, observed preferentially in the first stage at the beginning of the fouling process, is flux decline due to adsorption of organics. If the elements in the last stage are more affected then it is likely that fouling process is result of high concentration of fouling constituents due to an excessive recovery rate. The most common fouling phenomena, observed in the last stage, is formation of inorganic scale: either carbonate or sulfate. Less frequently scale of polymerized silica precipitates from the concentrate stream. Sometimes a mixed layer, containing both inorganic and organic foulants, is formed on the membrane surface. The mixed foulant layers are quite common as precipitation of one component, usually serves as precipitation centers and initiates precipitation of other constituents, which are at saturation concentrations.

Once the presence of fouling process in the RO system has been established through performance normalization, the next step is to remove elements for evaluation. For meaningful evaluation a lead and tail elements are required.

TABLE 15.2
Foulants and performance recovery evaluation matrix

System component	Evaluation procedure	Results
Feed water	Chemical tests and physical determinations	Feed composition, turbidity, SDI
RO system	Normalization of system performance data.	System normalized flow, salt passage and pressure drop
RO system	Evaluation of system operation • Starup and shut down procedures • Operation of the pretreatment system	Testing for possible equimnet malfunction and/or process design mistakes
RO element	Element appearance and weight. Element performance test. Element integrity test	Comparison with element specifications
Flat membrane	Performance tests Membrane surface analysis	Comparison with specifications and with performance of new membrane
RO element	Testing of effectiveness of cleaning methods	Identification of effective cleaning procedure

Sometimes, a full load of a single pressure vessel is removed for testing. The first step is to examine elements appearance and determine their weight. Presence of slime on element outside surface indicates biological fouling. Accumulation of particles on the inlet element area indicates inadequate operation of the filtration system. Reddish-brown deposit usually indicates carry over of iron flocculant from the pretreatment system.

The next step in the evaluation process is to test membrane element performance at nominal test conditions and compare results with ex-factory test results. An example of such evaluation is shown in Table 15.3. In this case a complete load of one pressure vessel, containing eight elements, from a single stage seawater RO unit, was removed for testing.

The results clearly indicate that elements maintained ex-factory values of permeate flow and salt rejection but some experienced blockage of the feed

TABLE 15.3
Results of evaluation of one complete load of eight elements from one pressure vessel

Position	Ex-factory Rej, %	Ex-factory m^3/d (gpd)	Retest Rej, %	Retest, m^3/d (gpd)	Retest DP, bar (psi)	Increase of nominal value,%
1 (lead)	99.79	20.26 (5354)	99.78	19.78 (5227)	0.65 (9.5)	195
2	99.70	19.40 (5126)	99.84	20.89 (5519)	0.34 (5.0)	54
3	99.74	21.68 (5729)	99.81	22.76 (6013)	0.31 (4.5)	41
4	99.78	21.86 (5775)	99.81	22.99 (6074)	0.28 (4.0)	27
5	99.80	20.54 (5427)	99.88	20.66 (5459)	0.26 (3.8)	24
6	99.74	21.01 (5511)	99.83	23.35 (6171)	0.24 (3.5)	18
7	99.80	23.19 (6127)	99.84	23.08 (6097)	0.21 (3.0)	—
8 (tail)	99.80	23.19 (6127)	99.85	22.99 (6074)	0.22 (3.2)	—

channels. As expected, the pressure drop is highest across the lead element, decreasing close to the normal levels for the elements located in 4th and 5th positions. The results included in Table 15.3 are representative for operating conditions of feed water containing high concentration of colloidal matter (high turbidity and SDI). The nature of the fouling process was determined through evaluation of normalized system performance data and EDX analysis of SDI filter pad, and membrane surface samples taken from the element in lead position.

Fig. 15.1 shows representative profiles of pressure drop along pressure vessel for different types of fouling. Case A corresponds to case of colloidal fouling (case discussed above). Case B corresponds to advanced stages of biofouling. Distribution of elements pressure drop values in case B is similar to distribution shown as case A. In both two cases, the pressure drop is highest in the lead element, gradually decreasing toward the concentrate end of the vessel. The final determination of nature of fouling process is based on evaluation of feed water quality and deposits found in pressure vessel and on the membrane

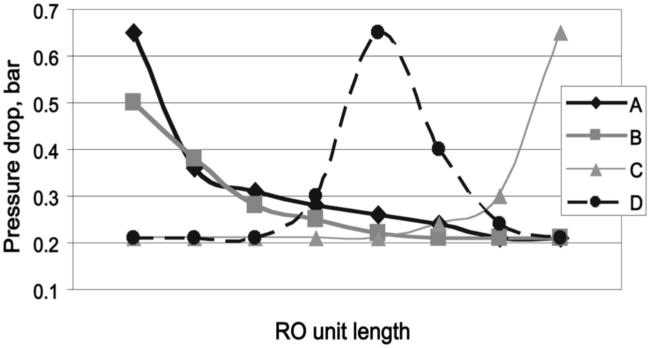


FIG. 15.1 Pressure drop distribution of fouled elements vs. element position.

surface. In case C, the pressure drop has its highest value at the concentrate end of pressure vessel. This situation is characteristic for membrane scaling process. However, in some rare cases the scaling deposits can be formed in elements positioned at some distance from the system end (case D). Such conditions can happened if the saturation concentration of scale forming salt is reached in the RO system in the middle section, somewhat before concentrate exit. If the following tail elements operate at very low fluxes, producing little water, and therefore providing small contribution to the overall recovery rate, scale formation rate will not be significant in the elements in the tail position. Usually pressure drop increase, related to fouling, shows similar pattern as the corresponding weight gain of individual elements.

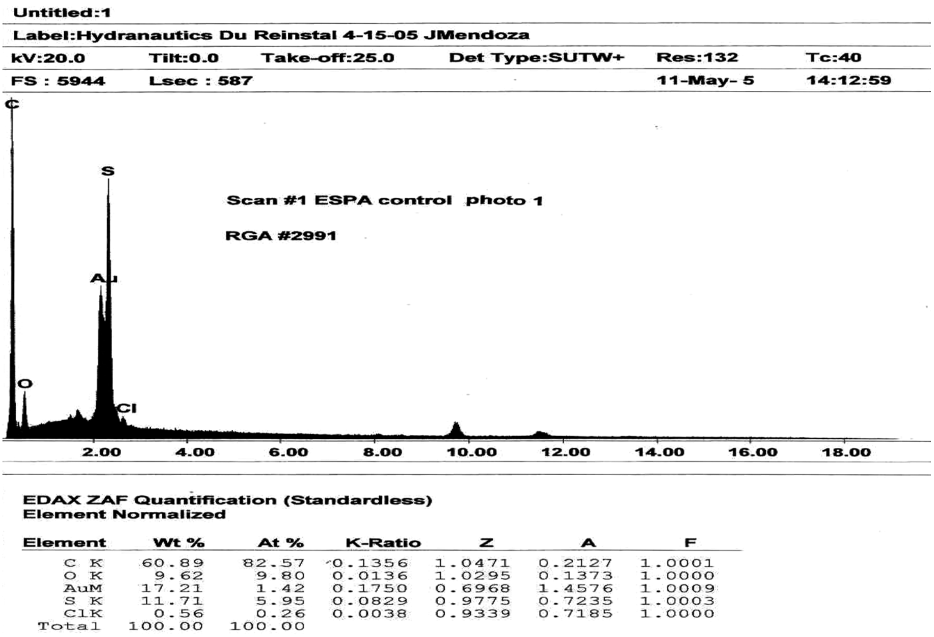
If individual elements test results indicate high salt passage, it is common procedure to check integrity of membrane element by applying dye passage test or conducting bubble test. The dye passage test involves operation of elements with feed solution containing dye (10% methyl violet solution) at nominal feed pressure. During this test dye will stain the membrane around location of leaks and will make color marks on the permeate spacer. These points of high dye passage can be located latter on when the element is unrolled and membrane leaves cut open. The dye is also a useful method for determination if membrane barrier has been damaged by exposure to harsh chemicals. Areas where membrane barrier have been chemically damaged have preferential adsorption of the dye indicated by strong coloration when element is open (dissected). The intact membrane surface shows very little coloration after the dye test. In case of very large leaks, dye presence in the permeate stream can be observed. Conducting of the second type of integrity test, the bubble test, involves closing one side of permeate tube with a stopper and application of clean air at low pressure (about 0.2 bar, 3 psi) to permeate side of the element. During this test the element is immersed in water and continuous stream of air bubbles is indicative of lack of membrane integrity.

According to results of individual elements examination and testing, the next step may include either screening effectiveness of cleaning procedures or conducting autopsy of one of the element to get more insight to nature of the foulant(s). Element autopsy is a destructive test. Therefore, it is more common to conduct it in case of fouling problem in large capacity systems. In this process element is open and leaves unrolled. The surface of membrane leaves are visually inspected for presence of membrane defects and presence and appearance of fouling deposits. If the initial element performance test indicated significant loss of rejection an effort should be made to identify location of leak(s). In case of heavy fouling of membrane surface, samples of foulant are collected for analysis. Collected foulant sample can be analyzed using conventional and instrumental methods of chemical analysis. Some insight regarding nature of the foulant can be obtained by testing fouling deposit for loss of ignition. High loss of ignition, in excess of 40%, is indicative of organic (containing carbon) fouling, usually of biological nature. Rapid, qualitative method of determination of carbonate scale presence is exposure of foulant material to mineral acid. If carbonates are present, effervescence due to release of carbon dioxide will be observed.

After visual examination of membrane surface small segments of membrane are cut out and submitted for surface analysis The useful surface analysis methods are SEM, EDX and FTIR. SEM provides high magnification pictures of the membrane surface. It enables visual assessment of the conditions of membrane surface. Examples of SEM pictures of clean and fouled membrane surfaces are shown in Fig. 17.8. When evaluating SEM pictures and EDX spectra it is important to remember that size of the membrane samples used for these analysis are very small, usually less than 1 cm^2 (0.15 in.^{2}). Therefore, it is important to select for this type of instrumental analysis a representative membrane samples. EDX analysis of membrane surface results in a spectra which are characteristic for specific atoms. However, majority of EDXequipment can not detect atoms below atomic number of 11 (Na). EDX is an effective method for identification of inorganic deposits on the membrane surface. However, the electron beam that initiates atom specific X-ray emission, penetrates about 2 micron into bulk of the membrane layer. Therefore some interpretation is required in reading the EDX spectra regarding origin of the emission peaks. Figs.

15.2, 15.3, and 15.4 show EDX spectra taken from a clean and two heavy fouled polyamide membrane surfaces respectively. On Fig. 15.2 the emission peaks designated as carbon (C), oxygen (O) and sulfur (S) originate from building components of polyamide membrane polymer and polysulfone support. Gold (Au) peak comes from the minute layer of gold deposited on a drayed membrane sample to provide electric conductivity. The EDX spectra also provide indication about relative concentration of the components in the surface layer. The weight fraction shown in the bottom of Fig. 15.2 indicates that for clean membrane the predominant components are carbon, followed by sulfur and oxygen.

Spectra shows on Figs. 15.3 and 15.4, for heavy fouled membrane surfaces, are quite different than those on Fig. 15.2. The relative fraction of carbon and sulfur decreases and other peaks shoved up. The new significant peaks are of silica (Si), iron (Fe), aluminum (Al) and phosphorus (P). The oxygen peak size increased compared to a spectrum of a clean membrane. The decrease of carbon and sulfur peaks is a result of lower penetration of electron beam into membrane due to shielding by the foulant layer. The oxygen peak increase is most likely associated with additional presence of oxygen in the foulant layer as a component of oxides (Si, Al) and hydroxides (Fe).



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kV:20.0 Tilt:0.0 Take-off:25.0 Det Type:SUTW+ Res:132 Tc:40

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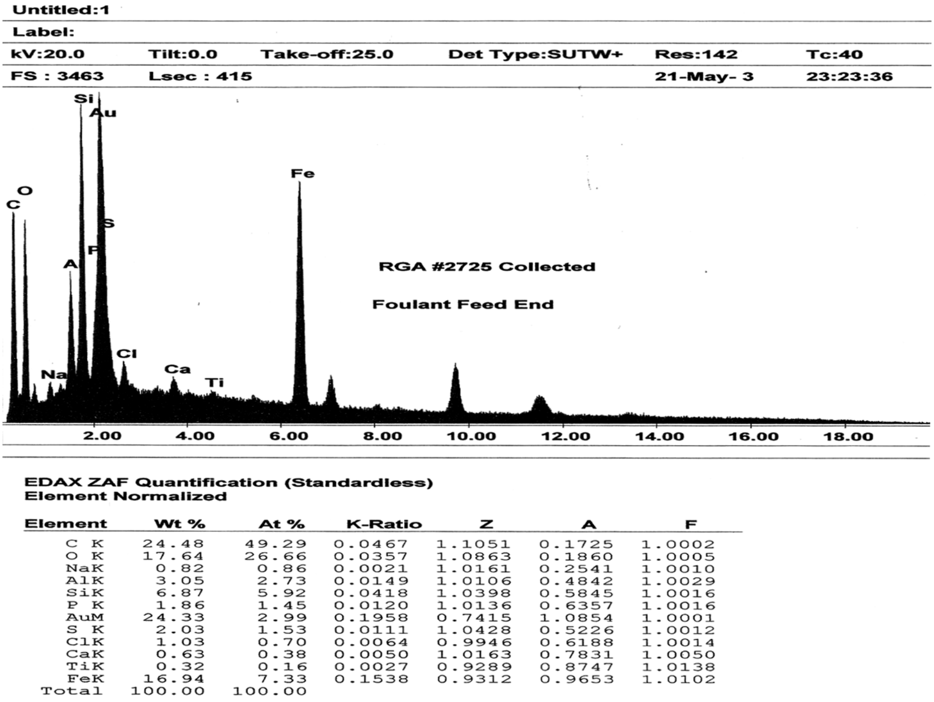
EDAX ZAF Quantification (Standardless)
Element Normalized

Element	Wt %	At %	K-Ratio	Z	A	F
C K	60.89	82.57	0.1356	1.0471	0.2127	1.0001
O K	9.62	9.80	0.0136	1.0295	0.1373	1.0000
AuM	17.21	1.42	0.1750	0.6968	1.4576	1.0009
S K	11.71	5.95	0.0829	0.9775	0.7235	1.0003
ClK	0.56	0.26	0.0038	0.9339	0.7185	1.0000
Total	100.00	100.00				

FIG. 15.2 EDAX spectrum of a clean composite polyamide membrane surface.



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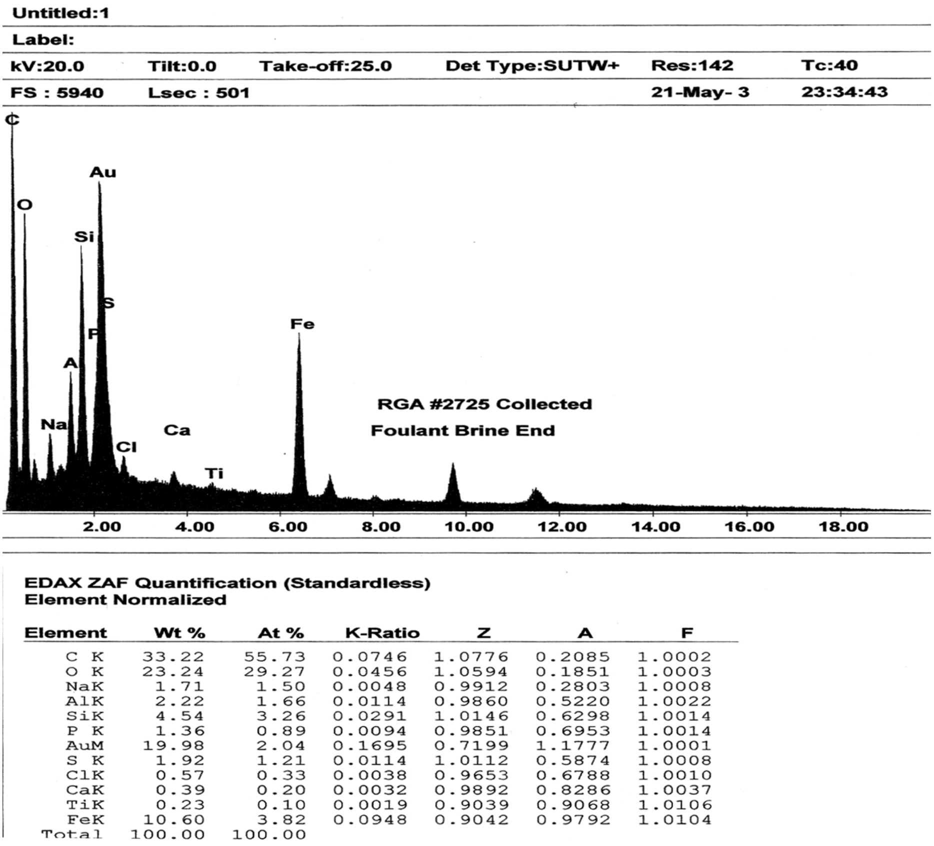
EDAX ZAF Quantification (Standardless)
Element Normalized

Element	Wt %	At %	K-Ratio	Z	A	F
C K	24.48	49.29	0.0467	1.1051	0.1725	1.0002
O K	17.64	26.66	0.0357	1.0863	0.1860	1.0005
NaK	0.82	0.86	0.0021	1.0161	0.2541	1.0010
AlK	3.05	2.73	0.0149	1.0106	0.4842	1.0029
SiK	6.87	5.92	0.0418	1.0398	0.5845	1.0016
P K	1.86	1.45	0.0120	1.0136	0.6357	1.0016
AuM	24.33	2.99	0.1958	0.7415	1.0854	1.0001
S K	2.03	1.53	0.0111	1.0428	0.5226	1.0012
ClK	1.03	0.70	0.0064	0.9946	0.6188	1.0014
CaK	0.63	0.38	0.0050	1.0163	0.7831	1.0050
TiK	0.32	0.16	0.0027	0.9289	0.8747	1.0138
FeK	16.94	7.33	0.1538	0.9312	0.9653	1.0102
Total	100.00	100.00				

FIG. 15.3 EDAX spectrum of composite foulant sample from membrane element in the lead position.

The EDX spectrum in Fig. 15.3 is from membrane sample taken from lead element and the one on Fig. 15.4 corresponds to membrane sample take from element in tail position. Comparison of two spectra clearly indicates higher fractions of fouling components on the membrane surface in the element in lead position: Si, Fe and Al. This would be consistent with colloidal materials, silt and iron flocculant carry over from the pretreatment, arriving to membrane elements with the feed water. It is not clear however, why the oxygen fraction in the surface layer is highest on the membrane sample from the tail position (about 23%) element. This is even though the apparent quantity of foulant in the tail element is lower.

The Fourier Transform Infra Red (FTIR) analytical method is sensitive to the vibration mode of molecule and is applied to identify functional groups of organic compounds attached the membrane surface. Figs. 15.5 and 15.6 show FTIR spectra taken from a clean and heavy fouled polyamide membrane surfaces respectively. FTIR is most effective in identification of organics adsorbed



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Label:

kV:20.0 Tilt:0.0 Take-off:25.0 Det Type:SUTW+ Res:142 Tc:40

FS : 5940 Lsec : 501 21-May- 3 23:34:43

EDAX ZAF Quantification (Standardless)
Element Normalized

Element	Wt %	At %	K-Ratio	Z	A	F
C K	33.22	55.73	0.0746	1.0776	0.2085	1.0002
O K	23.24	29.27	0.0456	1.0594	0.1851	1.0003
NaK	1.71	1.50	0.0048	0.9912	0.2803	1.0008
AlK	2.22	1.66	0.0114	0.9860	0.5220	1.0022
SiK	4.54	3.26	0.0291	1.0146	0.6298	1.0014
P K	1.36	0.89	0.0094	0.9851	0.6953	1.0014
AuM	19.98	2.04	0.1695	0.7199	1.1777	1.0001
S K	1.92	1.21	0.0114	1.0112	0.5874	1.0008
ClK	0.57	0.33	0.0038	0.9653	0.6788	1.0010
CaK	0.39	0.20	0.0032	0.9892	0.8286	1.0037
TiK	0.23	0.10	0.0019	0.9039	0.9068	1.0106
FeK	10.60	3.82	0.0948	0.9042	0.9792	1.0104
Total	100.00	100.00				

FIG. 15.4 EDAX spectrum of composite foulant sample from membrane element in the tail position.

on membrane surface, such as filtration aid polymers, scale inhibitors and other contaminants.

The final step in the membrane elements examination process is evaluation of effectiveness of cleaning procedures. Based on the results of the above described elements examination, cleaning procedures are selected and tested initially on single elements. Membrane elements for conducting of the initial cleaning tests should be selected from positions in systems that have been most affected by fouling. The elements performances should be tested prior to cleaning and after the cleaning procedure is completed. If more than one cleaning solution is being evaluated, it is recommend to test element performance after each cleaning solution has been applied, at least during the initial cleaning attempts. The cleaning procedures that were found most effective in single elements

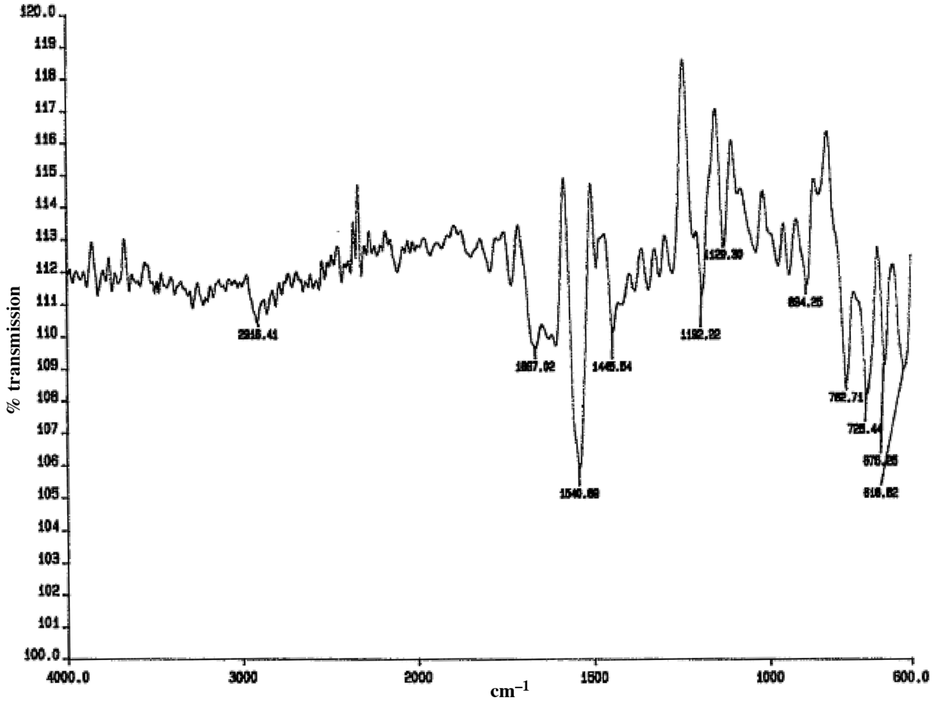

FIG. 15.5 FTIR spectrum of a clean aromatic polyamide membrane.

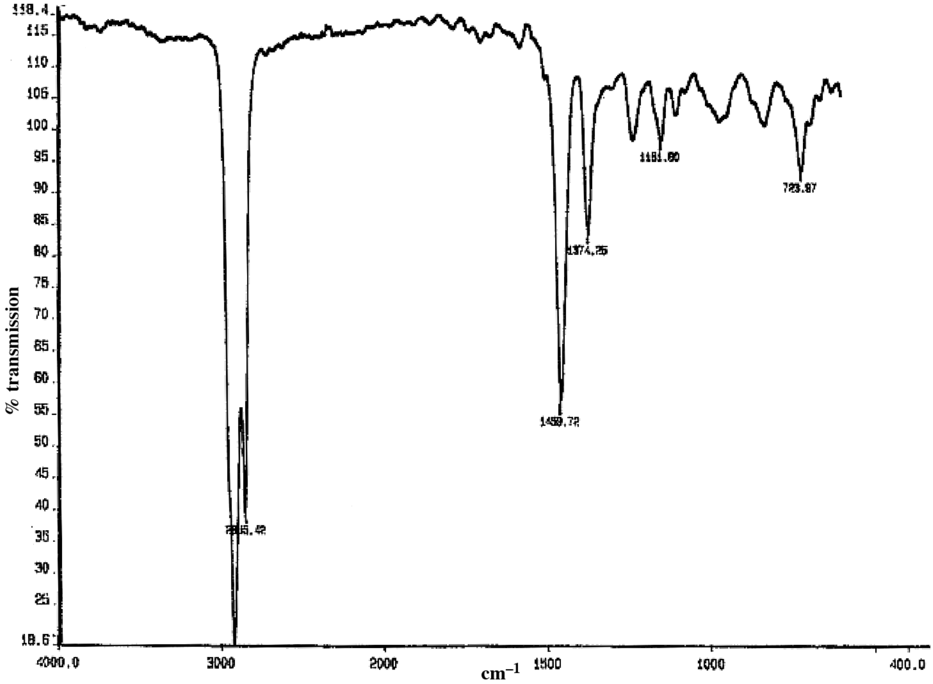

FIG. 15.6 FTIR spectrum of membrane surface contaminated with petroleum products.

cleaning tests are applied to restore performance of elements in the commercial trains. Extensive discussion of membrane evaluation process and interpretation of results is included in references 49, 50 and 137.

15.2. Performance restoration

Cleaning of elements in an RO train is conducted using the cleaning unit. The configuration of cleaning unit is shown in Fig. 15.7. It consists of cleaning tank, heater, recirculation pump, cartridge filter and connecting piping. Larger cleaning units also include separate tank for dissolving and mixing of cleaning solutions. Materials of construction of the cleaning unit should be selected to withstand low and high pH cleaning solutions (pH 2–11) at temperatures up to 50°C. The size of cleaning tank and capacity of cleaning pump is determined by the number of pressure vessels that will be cleaned at one time. During cleaning operation the flow rate of cleaning solution per vessel should be close to 7–9 m^3/h (~30–40 gpm). The cleaning tank volume should hold enough cleaning solution volume to provide at least 5 min of pump capacity. If, for example, 48 pressure vessels will be cleaned at one time, the operational volume of the cleaning tank should be about 19 m^3 (5,000 gallons).

In large RO systems the connecting piping of the cleaning unit is permanently attached to all trains. Valves or removable piping segments are used to connect/disconnect given train or train segment to the cleaning unit.

Cleaning operation sequence includes:

1. Flushing RO train with permeate water.
2. Connecting train or train segment to the cleaning unit.

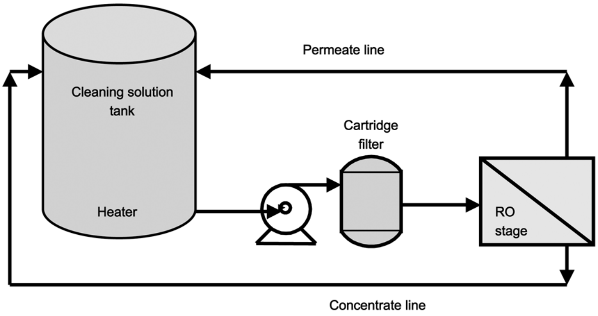

FIG. 15.7 Configuration of a cleaning system.

3. Preparing cleaning solution in the cleaning unit.
4. Recirculating cleaning solution for 1–4 hours through the RO train.
5. Flushing cleaning solution
6. Repeating steps 2–5 with next cleaning formulation or reconnecting cleaned train to high pressure pump and restoring normal operation.

Membrane cleaning, like any other dispersive process, is more effective at elevated temperature. Cleaning should be conducted at temperature of cleaning solution in the range of 35–40°C. Cleaning solutions can be purchased from specialized suppliers or generic cleaning formulations can be used. Composition of generic cleaning formulations can be obtained from all major membrane manufacturers.

One of the generic low pH cleaning formulation, frequently used, is 2% solution of citric acid. pH of such solution is about 2.5. Citric acid cleaning solution is very effective in removal of deposits of metal hydroxides and dissolving of carbonate scale. If it has been established that fouling deposit contains mainly calcium carbonate or metal hydroxides, temporary operation with feed water acidified to low pH (pH = 4.5–5) with mineral acid ($H2SO_4$ or HCl), may be sufficient to restore membrane performance. Cleaning, through operation at low feed pH, is only possible if discharge of low pH concentrate is allowed by local regulation at a given site.

The generic high pH cleaning formulations consist of solutions of NaOH in combination with EDTA or SDBS (surfactant). These cleaning solutions have pH of 10–11 and are effective in removal deposits of organic matter from membrane surface. It has been found (62) that EDTA or surfactants are essential components of high pH cleaning solutions and their presence contributes to improved removal of surface deposits that contain Ca ions imbedded in the organic fouling layer. In majority of cases fouling layer is of a mixed nature, it contains a mixture of inorganic and organic matter. The effective cleaning s equence is to apply low pH cleaning followed by application of high pH formulation.

RO systems operating on well water feed seldom need membrane cleaning. Cleaning frequency is usually less than one cleaning per 2–3 years of operation. RO systems treating surface water feed require more frequent membrane cleaning. In well designed and operated seawater RO systems cleaning is conducted at 1–2 year intervals. In seawater systems with inadequate feed water quality, required cleaning frequency could be much higher. For the purpose of operating cost estimation, budget for cleaning operation is usually based on two cleaning events per year. If more frequent cleanings are required, then it is an indication of inadequate pretreatment process.

Example 22

Calculation of annual cleaning cost.
System permeate capacity: 100,000 m^3/d (26.4 mgd),
RO unit configuration: 8 trains, 120 PV per train, 8 elements per vessel
Train segment size for a single cleaning: 60 pressure vessels.
Annual cleaning frequency: 2
Cleaning procedure: low pH cleaning followed by high pH

Free volume of pressure vessels: $60 \times 8 \times 0.025 m^3 = 12 m^3$ (3200 gal)
Volume of manifolds (10% of PV): $0.1 \times 12 m^3 = 1.2 m^3$ (320 gal)
Volume of connecting piping (50% of PV): $0.5 \times 12 m^3 = 6 m^3$ (1600 gal)
120 PV $\times$ 8 m^3/h $\times$ 3min/60 min = 24 m^3 (6,300 gal)
Total volume of cleaning solution: 19.2 m^3 + 24 m^3 = 43.2 m^3 (11,400 gal)

Chemicals quantity for annual cleaning operation:
Solution 1–citric acid
2% Citric acid: $0.02 \times 43.2 = 0.864$ t/cleaning
8 train $\times$ 2 segments $\times$ 2 cleanings/year $\times$ 0.864= 27.6 t
Solution 2 – NaOH + SDBS
0.2% NaOH: $0.002 \times 43.2 = 0.086$ t/cleaning
8 train $\times$ 2 segments $\times$ 2 cleanings/year $\times$ 0.086 = 2.8 t
0.02% SDBS: $0.002 \times 43.2 = 0.0086$ t/cleaning
8 train $\times$ 2 segments $\times$ 2 cleanings/year $\times$ 0.086 = 0.28 t

Annual cost of cleaning chemicals

Citric acid: 27.6t × \$2500/t =	\$69,000
SDBS: 0.28t × \$3000/t =	\$840
NaOH: 2.8t × \$250/t =	\$700
Total cost of cleaning chemicals per year	\$70,540

Cost per water produced: \$0.002/m^3 (\$0.008/kgal)

As indicated by the above example the cost of generic cleaning chemicals is not significant if cleaning frequency is limited to two cleanings per year. Additional cost, associated with cleaning operation, that should be considered is loss of production capacity. The system off line time required for cleaning is in the range of 1–1.5 days, which corresponds to about 0.4% of availability The major expense related to cleaning, at some locations, could be the disposal cost of spent cleaning solutions.

In majority of cases cleaning operation is capable to restore some of the lost permeability and reduce pressure drop. Very seldom salt rejection is improved. Usually, it remains the same or can even decline after cleaning. This is because foulant layer plugs imperfections and damaged areas in membrane barrier and effective cleaning opens them again to salt passage.

If cleaning attempts does not result in sufficient performance improvement, membrane element replacement is the only practical solution available for additional performance correction. As discussed in section 10.10, usually a considerable fraction of elements in the system have to be replaced to achieve noticeable performance improvement. Number of elements that require replacement can be reduced if elements with worst performance can be identified in the RO system. Discussion of such approach is included in section 14.

15.3. Acknowledgements

The author would like to acknowledge the assistance he received from a number of individuals who were helpful in preparation of this book by sharing information and providing illustrations of processes and equipment. In particular, the cooperation of contributing authors (listed on the cover page), who significantly expanded contents of this book, is highly appreciated. Special appreciation goes to our publisher, Miriam Balaban, who provided extensive help, advice, and constant encouragement during preparation of this book.

16

Nanofiltration technology and applications

*Craig Bartels**

16.1 Nanofiltration overview

A rapidly growing field in the membrane industry is nanofiltration. The membranes and membrane properties are quite diverse, but can generally be described as having rejection characteristics that are "loose" RO or "tight" ultrafiltration (65). The uniqueness of these membranes is highlighted by their ability to selectively reject dissolved salts, and have high rejection of low molecular weight, dissolved components. Thus, it is unlike RO which has high rejection of nearly all dissolved species, and it is unlike UF which passes nearly all low molecular weight dissolved species.

These properties can have very important consequences for engineers designing separation processes for specialized applications (69). For example, these membranes can be used to partially soften potable water, allowing some minerals to pass into the product water and thus increase the stability of the water and prevent if from being aggressive on distribution piping. These features are becoming more important in municipal water treatment because of the increasing regulations on water quality, requiring even low salinity water to be treated. (66).

Additionally, NF membranes are finding increasing use for purifying industrial effluents, and minimizing waste discharge. Since the 1970s, many of the industrialized countries have had a growing focus on environmental stewardship.

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One of the key consequences of this trend for industry is that they must treat waste streams to recover water, reclaim valuable chemicals, or reduce discharge costs. Generally, these separations are quite complex, and require a wide variety of treatment processes and technologies. Although most of the effort has been directed at treating the final wastewater effluent, greater benefit can be achieved by isolating certain wastewaters or process streams which are more easily treated. The rise of nanofiltration technology has provided engineers with a new tool which is well suited to such applications, since membrane systems are modular and can cost-effectively treat small streams.

Perhaps the greatest value of nanofiltration will be its use to recover valuable low molecular products that may be synthesized in the drug, semiconductor, textile, metal-plating or food industries. In the past, separation processes such as crystallization, distillation, ion exchange, and evaporation were used for such separations. These conventional technologies suffer from not being very selective, being energy intensive, or causing degradation of the chemical component of interest. Nanofiltration, if sufficiently selective for many applications, can avoid high energy costs and will not chemically affect the compound. In many cases, nanofiltration membranes can also be more selective then conventional technologies.

One of the key features that distinguish these membranes from RO membranes is the low operating pressures. Operating costs for most membrane systems is governed by the energy consumption, and thus the required operating pressure. Since NF membranes are considered loose RO membranes, they naturally have higher water permeabilities, but they also operate at low pressures because they pass more salt ions. The net result of this is that the osmotic pressure in the system is lower, and thus the applied pressure is lower.

The key to using nanofiltration for these applications is the selection of the proper NF membrane and the design of a suitable process. NF membranes are generally characterized by a high charge density and pore sizes in the range of nanometers. (67) The charge is most often negative and has the greatest effect on the selective nature of these membranes. The design of a NF system can often be more complex than a RO system. Because the required transmembrane pressure is so low, the pressure losses in the system due to hydraulic pressures and osmotic pressure can be significant and affect both the water quality as well as the fouling rate. This will be discussed in some detail. Additionally, new applications and membranes will be considered to further address the full potential of this new field, which has only just begun to be utilized.

16.2 Nanofiltration membrane characteristics

16.2.1. Membrane types

Most NF membranes in commercial use are thin film composites. These are characterized by a membrane layer that is formed in-situ by interfacial reaction and consists of a polyamide separating layer. This layer, which may be only 0.2 microns thick, is supported by a porous polysulfone support and a non woven polyester fabric backing. (Fig. 16.1) This layered structure results in a high permeable membrane with outstanding mechanical strength. Examples of commercial composite membranes and nominal performances are shown in Table 16.1.

The polyamide nanofiltration membranes can be either fully aromatic (Fig. 16.1) or a combination of aromatic and aliphatic functional groups. The ones which are fully aromatic have a chlorine tolerance similar to polyamide RO membranes, which is on the order of 2000 ppm-hrs. The polyamides which are a combination of aromatic and aliphatic groups typically have much higher chlorine tolerance, 10–100 times greater.

The polyamide membranes made with fully aromatic groups typically have a higher chloride rejection and are used for softening, while the aromatic/aliphatic types have much higher chloride passage and are used for selective separation of charged ionic species. Because of the difference in the interfacial reaction, the aromatic/aliphatic polyamide membranes have a much smoother surface which may reduce fouling for some specific applications (68). Although, this chemistry has a number of functional advantages over fully aromatic polyamides, this type of chemistry cannot be made into a high rejection RO membrane.

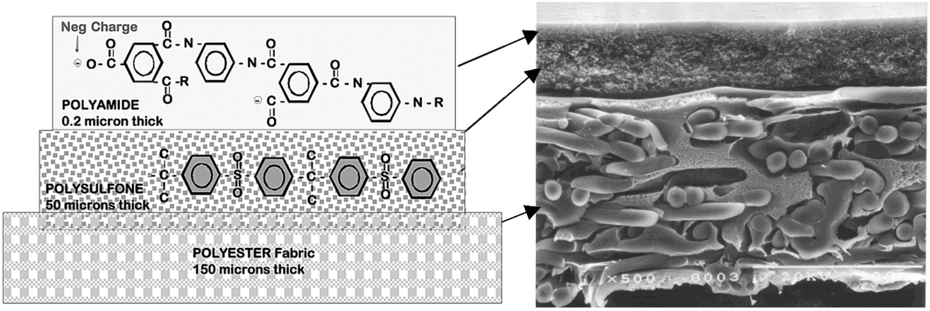


FIG. 16.1 Composition of a composite aromatic polyamide NF membrane.

TABLE 16.1
Common commercial nanofiltration membranes

Product Type	Memb Type	Nom Salt Rejection (%)	Nominal Flow (m^3/d)	Nominal Flow (GPD)	Element Area (m^2)	Element Area (ft^2)	Test Solution/ Concentration (ppm)	Test Pressure (bar)	Test Pressure (psi)
Dow NF200	PA	60	30.4	8000	37.2	400	500 ppm $CaCl_2$	4.8	70
Dow NF270	PA	97	47.5	12500	37.2	400	2000 $MgSO_4$	4.8	70
Dow NF90	PA	98	36.1	9500	37.2	400	2000 $MgSO_4$	4.8	70
Dow SR90	PA	NA	36.1	9500	37.2	400	2000 $MgSO_4$	4.8	70
Hydranautics ESNA1-LF	PA	90	31.2	8200	37.2	400	500 $CaCl_2$	5.2	75
Hydranautics ESNA1-LF2	PA	86	33.4	8800	37.2	400	500 $CaCl_2$	5.2	75
Hydranautics ESNA2	PA	99.5	36.1	9500	37.2	400	2000 $MgSO_4$	9.0	130
Hydranautics HydraCoRe50	PS	50	31.5	8300	33.9	365	500 NaCl	5.2	75
Koch TFC-S	PA	85	30.4	8000	37.2	400	500 NaCl	5.5	80
Koch TFC-SR2	PA	97	45.6	12000	37.2	400	2000 $MgSO_4$	3.8	55
Nitto Denko NTR 729-HF	PA	90	35.7	9400	37.2	400	500 NaCl	5.2	75
Osmonics Desal 5	PA	98	38.4	10100	37.2	400	2000 $MgSO_4$	6.9	100
Saehan NE8040-90	PA	99.5	34.2	9000	37.2	400	2000 $MgSO_4$	5.2	75
Toray SU650	PA	55	22.1	5812	34.9	376	500 NaCl	3.4	49
TriSep 8040-TS80-TSA	PA	99	34.2	9000	37.2	400	500 $MgSO_4$	7.4	107

Composite polyamide NF membranes have been used for many years in industry, one such example (69) refers to NF polyamide membranes from the 1980s. This is certainly a benefit to engineers who are designing systems, since there is historical data describing the performance of the NF membranes on a range of applications. Specific flux and rejection values from these studies can be used to estimate performance in comparable new applications.

The ability of these membranes to act as a selective rejection membrane comes from the presence of the negative charge on the aromatic ring when the acid chloride group reacts with water to form a carboxylate ion, COO^-. (Fig. 16.1) The number of these charged groups and the amount of crosslinking are altered to adjust the selectivity of this membrane for different dissolved species.

Another type of NF membrane is the charged polysulfone membrane. One example reported in literature by Nitto Denko (70) is the 7400 series membrane, which is a polysulfone support coated with a thin layer of sulfonated polyethersulfone. The surface charge is shown in Fig. 16.2 as a function of pH. Again, the combination of strong surface charge and relatively large pore size leads to a high flow membrane which can reject higher valence charged species.

Because this membrane is made of polysulfone polymers and not polyamides, it has much higher chlorine tolerance (70). A chlorine exposure of 1,000 ppm for 40 days had no effect on rejection, while the salt passage of a cellulose acetate membrane increased by 10 times in just 10 days at 100 ppm of chlorine. This is a big advantage as it allows the use of chlorine for cleaning or low doses of chlorine for continuous disinfection.

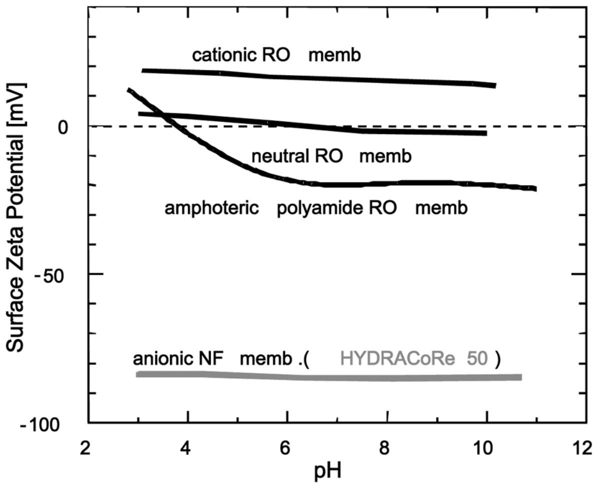


FIG. 16.2 Surface charge of a sulfonated polyethersulfone membrane compared to other RO polyamide membranes.

One rather unusual feature of this membrane is shown in Fig. 16.3. When challenged with a variety of ion pair salt solutions, it is apparent that the rejection is highest for divalent anions over monovalent anions, as one would expect. However, it also shows the rejection drops when switching from monovalent cations to divalent cations. This is rather surprising, and is likely due to the divalent cations being able to more effectively neutralize the negative charges at the membrane surface and form a thinner electric double layer. Thus, the electrostatic repulsion between membrane surface and ion is reduced, and rejection is reduced. The practical aspect of this feature is that the membranes have rather low salt rejection for common natural waters. The presence of calcium and magnesium cause the membrane to open up and readily pass salinity. For applications that require little or no salt rejection, but high rejection of low molecular weight polymers and negatively charged polymers, this feature can be a big asset. Examples will be given later.

Other types of nanofiltration membranes have been produced; these include cellulose acetate asymmetric membranes (71), positively charged interfacial membranes (72) and ceramic membranes (69). Again, each has specific properties which make it suitable for certain applications. The material properties and separation characteristics of these membranes should be reviewed before deciding which membrane to select for a given application.

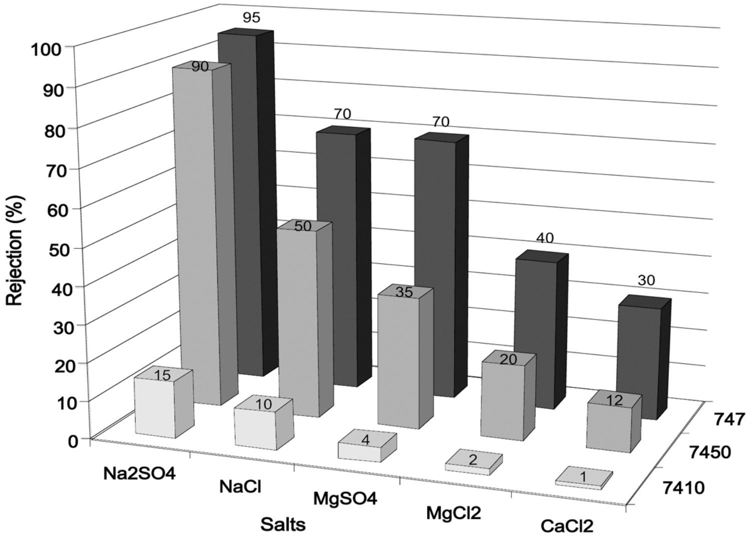


FIG. 16.3 Rejection characteristics of Nitto Denko 7450 membrane tested with various salt solutions.

Most of the NF membranes are prepared as sheets and thus rolled into a spiral wound element configuration. There are a few exceptions, one being the capillary polyamide membranes that are available from select vendors and tubular membranes which are common for the ceramic membranes.

16.2.2. Separation mechanism

Numerous papers have been written in an attempt to mathematically predict the performance of nanofiltration membranes. Most of these models have some success for a narrow range of operating parameters and feed constituents. In general these models become inaccurate when the feed constituents are varied significantly, due to the inability of these models to accurately model the separation mechanism. Most RO models are based on a solution-diffusion mechanism (73) and can adequately predict RO membrane performance over a broad range of operating conditions with very little laboratory characterization data. However, the application of the solution-diffusion model is not adequate to describe the charge effects and convective flows in NF membranes; thus it cannot extrapolate to extreme conditions. Various other models have been developed to predict NF separations properties (74).

Studies have shown that the size and charge of an ion can correlate to the rejection performance of NF membranes. Fig. 16.4 shows the ion rejection characteristics of an aromatic/aliphatic polyamide NF membrane treating a common 500 ppm salinity surface water operating at a flux of 25 lmh (15 gfd) and 75% recovery on a common surface water. Fig. 16.4a shows the rejection of the ions as a function of ionic radius. It is clear that the ion rejection does not correlate to the ionic size. However, when the same rejection data is plotted against the hydrated ionic radius (Figure 16.4b), there is quite a strong correlation when the anions and cations are considered separately.

Figure 16.4 clearly shows that a NF membrane more selectively rejects divalent anions, compared to divalent cations. The sulfate ion, which has a minus-2 charge, is most strongly rejected of any ion. This is because of the strong repulsion of the negative ion and the negative charges on the membrane surface. The divalent cations like calcium have much larger hydrated radii than sulfate, but lower rejection because there is no electrostatic repulsion. Interestingly, the chloride ion has a larger hydrated radius than the potassium ion and is negatively charged, but still is rejected less by the membrane. This is likely due to Donnan Equilibrium effects (75). For waters with mixed ions, the sulfate ion is strongly rejected, causing smaller anions to readily pass through the membrane

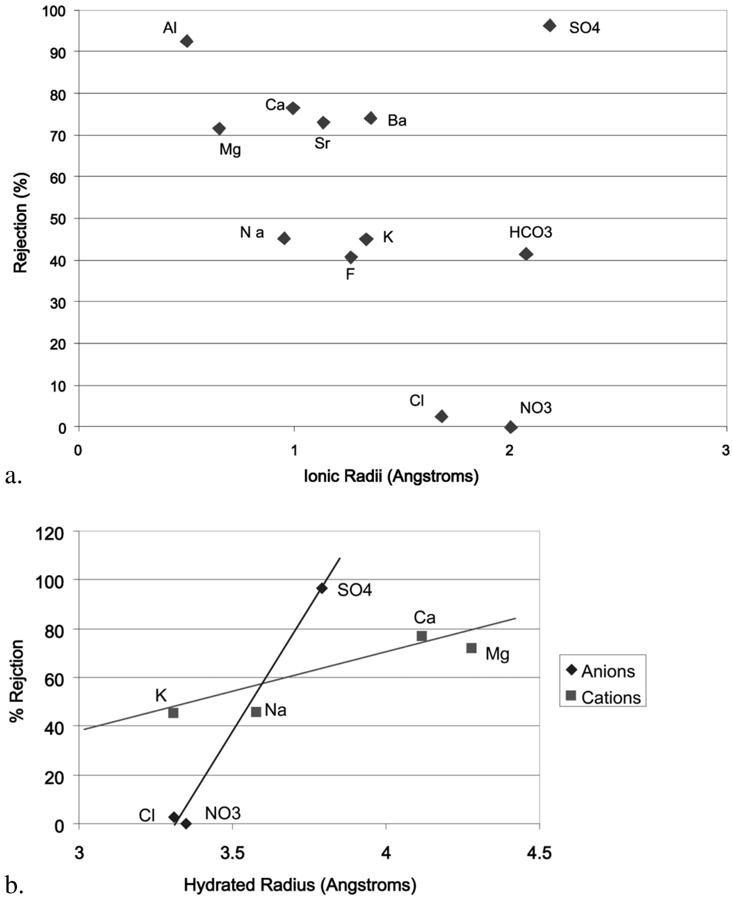


FIG. 16.4 Ion rejection for the ESNA2 polyamide membrane treating surface water plotted as a function of a) ionic radii and b) hydrated radii.

to neutralize the charge on the cations which pass through the membrane. For properly designed NF membranes, this behavior leads to the very desirable characteristic of high chloride passage, but high sulfate rejection.

One of the more successful models developed to describe the NF process was reported in 2002 (77). This approach models transport in nano-sized pores using Hindered Electro-Transport Theory and some very basic membrane characterization data. The development of an accurate model is important for expanded application of NF technology, since it will allow multiple users to estimate the performance at non-standard, non-tested operating conditions. Without such an advanced model, multiple membranes will have to be tested for each small variation of operating conditions.

16.2.3. Membrane properties

As described above, the rejection characteristics of a NF membrane vary greatly depending on the chemistry of the separating layer. Examples of the performance of some typical commercial membranes have been reported in a recent California Energy Commission (CEC) study (76). In the study, NF membranes were used to treat Colorado River water. The rejection of specific anions and cations is shown in Fig. 16.5. The results show the broad range of rejection that is achieved for different types of NF membrane, although the sulfate rejection of all the NF membranes is quite high. From this result, it also is apparent that the rejection of a given ion will depend on the particular mix of ions present in the feed water. For example, if there is a significant amount of sulfate in the water along with chloride, the chloride will be much more likely to pass through the membrane to neutralize the charge of cations that pass through the membrane, whereas, if there is bicarbonate ion and chloride ions in the feed water, the chloride rejection would be higher. Therefore, unless an accurate model is available, it is important to pilot test NF membranes for a given water unless there is already data available for similar water sources.

The selection of the best membrane for an application will depend on the feed water and product water quality requirements. For instances where high rejection of hardness is required, a tighter NF membrane is required. In contrast there are also cases where a customer desires little or no removal of salinity, but high rejection of organic material. In the latter case, a looser NF membrane should be selected.

There are a growing number of applications where removal of organic matter is critical. The organic material can be naturally occurring matter (NOM), natural or synthetic drugs, chemical intermediates, dyes, or polymeric products. Most NF membranes can be considered to have a molecular weight cut-off around 250 daltons, but this will vary depending on the type of NF membrane. A typical plot is shown in Fig. 16.6.

Molecular weight cut-off (MWCO) is most often considered the molecular weight at which 90% rejection is obtained. Based on this definition, the data shows that a tight NF membrane has a MWCO of about 180 daltons, while a loose NF membrane is considered if the MWCO is around 1000. It should be noted, however, that MWCO varies depending on the type of organic molecule used for the challenge. Less polar organic molecules would show higher values of MWCO. Thus, MWCO can only be used as a general guide with high accuracy for molecules similar to those used in laboratory challenge tests. Many studies have been made to relate the organic rejection of NF membranes to their

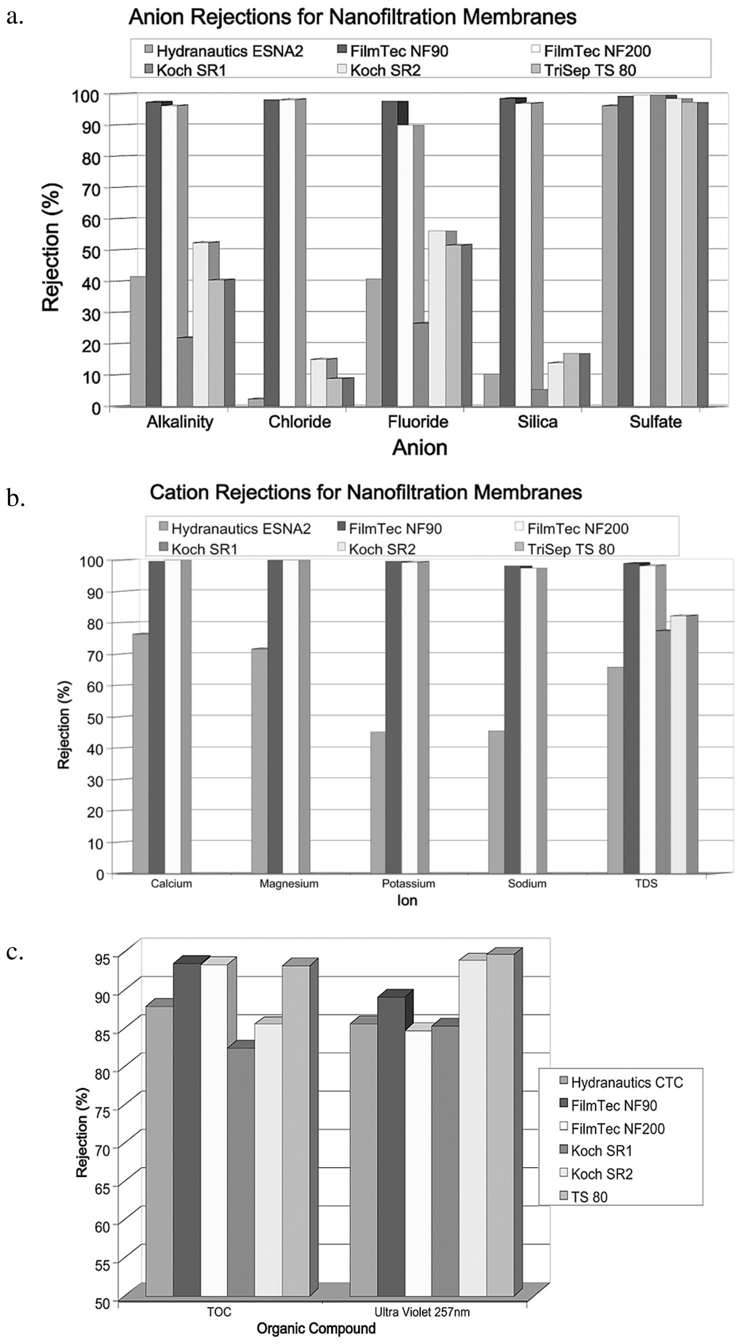


FIG. 16.5 Typical a) anion b) cation and c) organics rejection for various NF membranes treating a surface water.

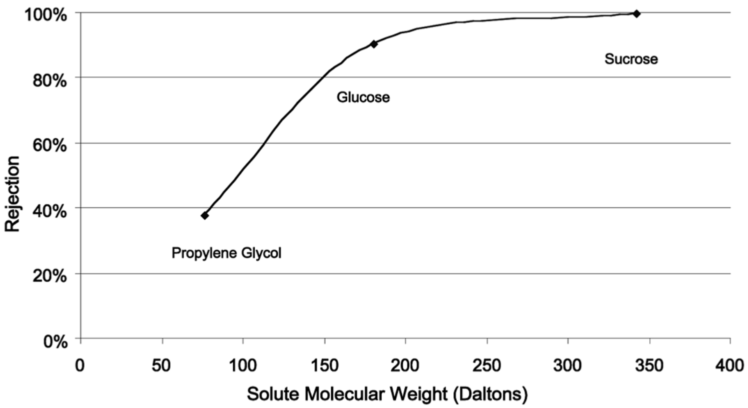


FIG. 16.6 Rejection of dissolved organic material as a function of membrane type and molecular weight for ESNA2 membrane. Organic compound.

physico-chemical properties (78), but the data in literature is still somewhat limited for the large number of different organic compounds.

In many cases, however, the separation objectives for a given application require organic rejections of 99+%. For example, most NOM found in surface water has an average molecular weight around 1000 daltons, and thus most of the membranes would give 99+% rejection and meet most water quality targets. In contrast, applications in the sweetener industry require the rejection of mono and disaccharides. These are in the range of 300 to 600 daltons and require tighter NF membranes to meet the rejection requirements.

NF processes will generally operate at pressures ranging from 5–14 bar (70–200 psi), with the majority operating around 7 bar (100 psi). For a typical plant, this will equate to a power cost of \$0.021/m^3 (\$0.08/kgal). As with RO membranes, NF membranes with lower rejection typically operate at lower pressure. This can be seen in Table 16.1, although the products are tested at a variety of pressures and salinities.

Since many NF membranes are used to remove organic materials, there is a likelihood that there can also be significant fouling. In addition to selecting optimum operating conditions which will be discussed later, low fouling NF membranes are also available on the market. Some examples of such membranes are the Duraslick™, which is reported to have a smooth surface (68), and the ESNA1-LF which is reported to have a smoother surface and a more neutral charge (79). More will be mentioned about these later.

16.3 Nanofiltration process considerations

Once an engineer has selected a NF membrane that has suitable rejection and permeability for the given application, optimum operating conditions should be applied. Many commercial membrane suppliers have software and technical recommendations that can help guide the selection of optimum design conditions, but these are often quite similar to the design constraints for RO membranes. Although the NF membranes are more permeable than RO membranes and often thought to have higher flux, they still operate in cross-flow mode, and thus require the proper balance of permeate-flow-to-reject flow to maintain the sweeping action that keeps concentration polarization to a minimum.

16.3.1. Pretreatment requirements

Almost all NF applications will utilize a spiral wound element configuration. The requirement for these elements should be SDI of less than 4 or turbidity of less than 0.2 NTU. Values higher (worse) than this can be treated, but there will be much more rapid fouling of the membrane. This fouling will be evidenced by increased differential pressure across (ΔP = inlet pressure – outlet pressure) the pressure vessel or by reduced membrane permeability. The foulant must be removed before it irreversibly affects the membrane. Thus, the system has more down-time and greater cleaning costs, and ultimately, shorter membrane life.

A wide range of pretreatment processes can be used to provide adequate water quality for NF, these include clarification, media filtration, dissolved air flotation, and micro/ultrafiltration. The latter is preferred as it gives the best quality. In all cases chemical coagulation should be used to remove excess colloidal material, and dispersants used to control precipitation of sparingly soluble ions. In the case of well water, pretreatment can be as simple as having a 5 micron cartridge filter. Some examples of different pretreatment will be given in the next section on actual plant reference sites.

An entirely different approach to NF applications is capillary NF membranes. (80) Since these do not have a traditional feed spacer and can be forward flushed like a capillary UF membrane, it is possible that poorer quality feed water can be treated without extensive pretreatment. According to study results, the NF capillary membrane was able to maintain stable performance with periodic forward flushing. This allows the NF membrane to treat more difficult water sources without significantly increasing pretreatment costs.

16.3.2. Process design features

The key process design features which must be considered are flux and recovery. The optimization of these parameters will insure that the fouling rate is not excessive and sparingly soluble components are not precipitated on the membrane surface. However, there are many operating parameters which can be varied to control these two critical parameters. Additionally, the choices available to a process designer are further limited by the water quality objectives that must be met.

Driving force. NF systems will run at low pressure, which gives rise to certain issues which must be considered by the process engineer. One of the key issues is that the process driving force, TMP, changes within the system (Eq. 8). A typical two stage NF system is shown in Fig. 16.7, where the feed pressure is only 6 bar (87 psi). At such a low pressure and high recovery, the driving force, expressed as the trans membrane pressure (TMP) is only 3.7 bar (54 psi) in the first stage, and 40% lower in the second stage. This is caused by the increasing osmotic pressure as the salts are concentrated and by the hydraulic pressure losses that occur as the water passes through each element. When the TMP drops by 40%, the flux will also drop by the same percentage. This leads to higher salt passage in the second stage. However, there are some ways that design engineers can minimize this effect.

Permeate flux. As mentioned, the quality of the feed water is the most dominant factor affecting the selection of flux. Thus, a lower average system flux (16–22 lmh, 9.4–12.9 gfd) should be used for poor quality feed water, while a moderate flux (23–30 lmh, 13.5–17.6 gfd) for good quality feed water. Definitions of poor and good quality feed water are given by the membrane suppliers; it will depend on SDI, turbidity and TOC content. The flux will affect water quality of the NF product. A relatively higher pressure will give a higher

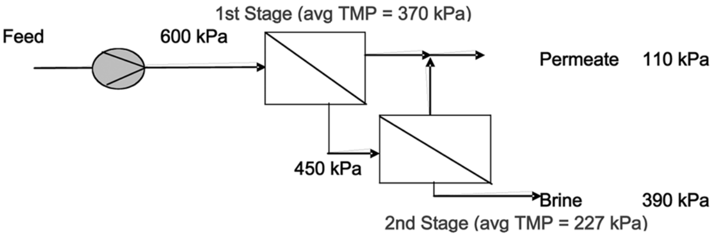


FIG. 16.7 Typical pressures in a 2 stage NF system.

flux and the better quality of permeate. This is due to that fact that the salt transport rate is not affected by the trans-membrane pressure, while the water permeation rate is. In general, a 20% increase in water flux will result in a 20% decrease in the apparent contaminant passage rate.

Poor quality feed waters can be characterized as having total organic carbon (TOC) levels in excess of 5 ppm, BOD levels of 10 ppm or more, turbidity greater than 0.5 NTU, and/or SDI greater than 4. However, it may be more cost effective to improve the pretreatment as opposed to running at low flux and having frequent membrane cleaning.

Flux distribution within the NF system is also important and often difficult to control due to the low applied pressures. Fig. 16.8 shows the flux distribution within the same 2 stage NF system shown in Fig. 16.7.

As long as the crossflow is sufficient and sparingly soluble salts do not precipitate, the lower flux rates should not cause an operational problem. The main effect will be the increased salinity in the second stage permeate. The other aspect, which must be considered by a designer, is the lead element flux. Most membrane companies have technical recommendations on the maximum flux of a lead element. Naturally, as the flux of the tail elements drops, the lead elements have to have higher flux to keep the target average system flux. Care should be taken not to exceed the manufacturers recommended lead element flux. A general guideline is that the lead element flux should not exceed the average system flux by more than 50%.

When treating high fouling water, it may be desirable to lower the lead element flux, and thus balance flux throughout the system. This can be done a number of ways, but each has a penalty associated with it. One means to balance flux is to use higher permeate back pressure in the first stage compared to

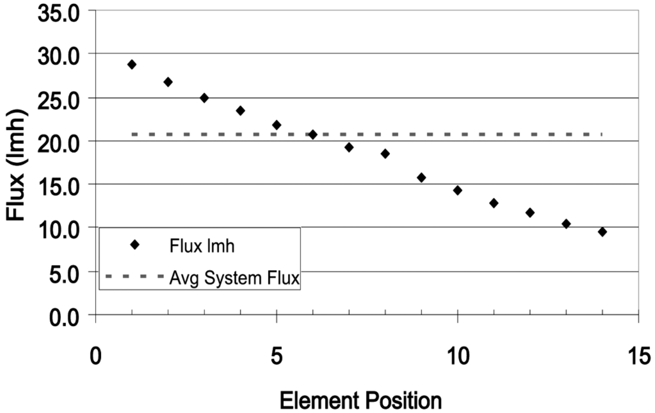


FIG. 16.8 Flux distribution in a 2 stage NF system.

the second stage. The problem with this concept is that it increases the overall pressure in the system, and thus the operation cost. A second alternative is to use a booster pump between stages. This will not significantly increase the pressure in the system, but it will increase the capital cost and make the system more complex. The booster pump can be a much smaller pump; for the system shown in Fig. 16.7, a 1.5 bar (22 psi) booster pump would be required. A third alternative, a hybrid design, will be discussed later.

Hybrid designs . Many engineers have turned to membrane selection as an optimum way to balance flux without increasing capital cost or obtain desired permeate salinity. Thus, a lower rejection, higher permeable membrane is put in the second stage. One example of this is at Fort Myers, FL plant which uses ESPA1 (RO) membranes in the first stage and ESNA1 (NF) membranes in the second stage. Other examples are the Boynton Beach NF plant that uses NF90 in the first stage and NF270 in the second stage, or the Ft. Lauderdale plant which will use high flux ESPA4 (RO) in the first stage, ESPA4 in the first 4 positions of the second stage vessel and ESNA1-LF2 (NF) in the remaining 3 positions of the second stage vessels. In this way the flux is increased in the second stage, but again the salt passage will be higher. (81) To some degree, the higher flux of the second stage element will compensate the lower rejection of this looser membrane. The only complexity of this concept is the need to manage two different membrane types at the site. For this type of solution, a membrane with about 10% higher productivity should be used in the second stage.

Recovery and scaling. It is always desirable to design a system with the highest recovery possible. This minimizes the size of the pretreatment system and reduces the amount of wasted water. The practical limit in recovery for NF systems is the potential for scale formation, membrane fouling, or diminishing pressure. Whenever natural waters are treated, some salt species will be concentrated to the point where precipitation occurs. The use of suitable antiscalants and dispersants can allow higher recoveries, but these chemicals will eventually become ineffective if the supersaturation is too great. Thus, great care should be taken to assess the potential for forming common scale materials, such as silicates, calcium carbonate, calcium sulfate, and barium sulfate. Most companies have software that will calculate scaling indices and warn a user when the scaling potential is dangerous.

The desired product water quality also affects the selection of the recovery rate. This is particularly true for nanofiltration systems because the rejection for a number of ion by NF membranes is not extremely high as it is in RO. An illustration of this is given in Fig. 16.9. The Fig. shows three curves which

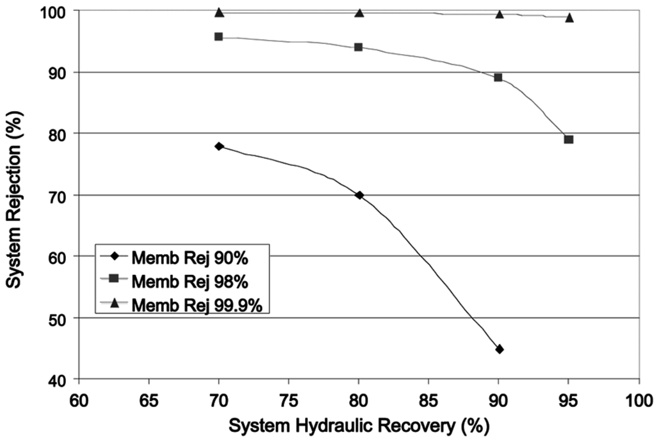


FIG. 16.9 Effect of membrane rejection and recovery on product water quality.

represent the rejection of three different compounds. The figure shows the system rejection (= [Feed Conc-Permeate Conc]/Feed Conc) as a function of system recovery. The rejection values, 99.9, 98 and 90%, are those that would be measured with a single element on a model test solution. At a high recovery of 90%, the compound which is rejected 90% by the membrane shows a system rejection of only 45%, whereas the compound rejected at 99.9% still has a high system rejection of 99.5%. The single element intermediate rejection of 98%, has also dropped significantly to 89% at the 90% recovery. Thus, for example, if some naturally occurring matter is rejected at 99.9%, it may be possible to go to 90% recovery and achieve the desired product water quality of this compound. However, if the designer also needs to achieve over 90% reduction of calcium, the membrane rejection of calcium must be greater than 98%. Alternatively, the recovery could be reduced to 85% if the membrane can achieve 98% rejection of calcium.

This analysis shows that a designer must identify the critical compounds early in the membrane selection process. A full analysis of the feed water is necessary and clear targets for the product water need to be established also. Again, to ensure sufficient rejection of the items of interest, pilot testing may be required.

16.4 Nanofiltration applications

16.4.1. Potable water

Softening–Boca Raton. One of the largest NF plants in the world is the 151,000 m^3/d (40 mgd) plant in Boca Raton, FL, USA. This plant has a dual

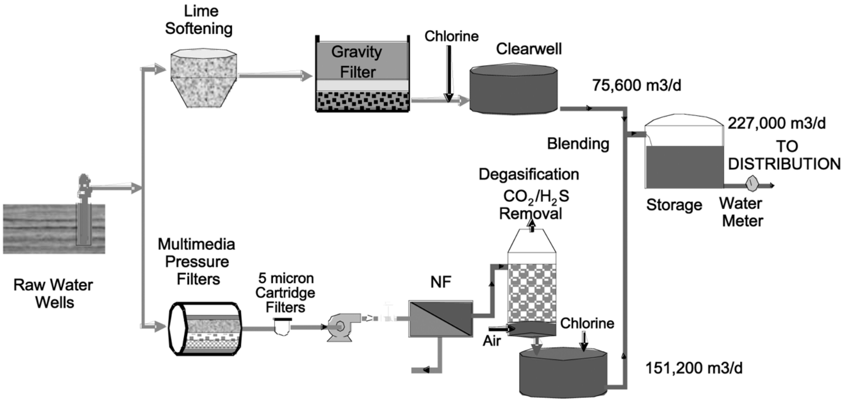
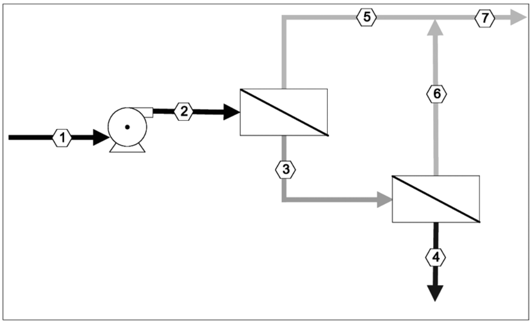


FIG. 16.10 Boca Raton water treatment plant process design.

process, chemical softening and membrane softening, to provide a blended water that meets municipal water quality targets. A schematic of the process is given in Fig. 16.10.

A majority of the water (2:1 ratio) is processed by the membrane plant, which, through blending, helps improve the water quality from the lime softening plant. To meet the final water quality goals, the target performance of the membrane plant is shown in Table 16.2. This plant is a good demonstration of the complexities of designing a NF system. To meet the water quality goals, the hardness needs to be reduced from 387 ppm to 50-80 ppm (as $CaCO_3$). This specific hardness range was specified by the city to avoid corrosion problems while providing a softer water. To meet federal regulations on disinfectant/disinfection by-products (D/DBP), the TOC needed to be reduced from 12 mg/L to less than 1 mg/L. Thus, the organics which lead to trihalomethane or haloacetic acid formation (THMFP or HAAFP) needed to be reduced below 42 and 30 ppb, respectively, so that the blended water would meet target values of 64 and 48 ppb, respectively. These requirements led to the selection of a specific nanofiltration membrane.

The plant was designed with 10 trains having 72 vessels in the first stage and 36 vessels in the second stage. Each vessel has 7 Hydranautics ESNA1-LF2 nanofiltration elements. These two stages were designed to recover 85% of the feed water, while two additional trains with a 36 × 18 array were added to recover concentrate from the first trains or be used as additional stage 1 and 2 trains. These trains contained a higher rejection ESNA1-LF3 element. Ultimately, it was decided that Trains 11 and 12 should be used on the raw feed



	1	2	3	4	5	6	7
Flow m3/hr	682.6	682.7	264.1	102.4	418.5	161.7	580.2
Pressure bar	0.0	6.0	4.6	3.4	0.8	0.0	0.0
TDS (ppm)	465.8	465.8	1014.4	2107.5	119.7	322.2	176.1

FIG. 16.11 Boca Raton NF process design for stages 1 and 2.

water. The first stage has 0.84 bar (12 psi) of backpressure to better balance the flux between the two stages(the RO process is shown schematically in Fig. 16.11). The design flux rate is 20.7 lmh (12.2 gfd) and the temperature is consistent around 25°C (77°F). The conservative flux value was selected for this ground water due to the high organic content of the water.

Table 16.2 also shows the actual permeate recovered from the two stage plant. It shows that all water quality targets were achieved, with some safety factor. This has led to a noticeable improvement in the delivered water quality compared to the water previously provided by lime softening alone.

The high level of NOM in this water also led to unique challenges and benefits. When the feed water is treated as filtered, it has a pH near 7.2 and the concentrate has a Langelier Saturation Index (LSI) of 1.8. Normally, water with such an LSI would require the addition of acid or antiscalant to prevent scaling when the recovery is in the range of 75–85%. However, pilot testing in this project demonstrated that it was possible to operate at such LSI without acid or antiscalant and not have scale formation. This was attributed to the high levels of NOM in this water (TOC of 12 mg/L), which acts as a natural antiscalant. However, the high levels of organics also pose a problem–a high membrane fouling rate. Some of the membranes tested in the pilot program did show high rates of fouling that led to higher than desired feed pressure and higher than desired rejection. The pilot testing did demonstrate that a low fouling nanofiltration membrane was suitable to prevent organic fouling while still meeting the

TABLE 16.2
Boca Raton Performance Requirements and Achievements

Parameter	Feed	1st and 2nd stage permeate	Target value
Total hardness (mg/l $CaCO_3$	387	75	50-80
HCO_3 (mg/l as $CaCO_3$)	209	74	<175
TDS mg/l (sum of ions)	387	137	<300
TOC (mg/l)	12	<0.8 (THMFP 16 ppb) (THAAFP 15 ppb)	<1.0 (THMFP 42 ppb) (THAAFP 30 ppb)
Color (CU)	35	1.2	<2
Recovery		85%	85%
Trans hembrane pressure @ 25°C		4.8 bar	<5.6 bar
Average flux (lmh)		20.7	20.7

water quality targets. As reported in a recent publication (82), the use of this membrane and unique process has led to substantial savings. The amount saved included, not adding acid ($572,000), caustic ($97,000) or antiscalant ($258,000), plus the additional energy savings from the enhanced permeability of the membranes, was in excess of $48,000 per year. This easily paid for the additional cost of pilot testing and using low fouling membrane.

The plant started up late 2004 and has had steady operation, aside from some operational upsets. Fig. 16.12a shows the normalized permeate flow during the first eight months of operation for train 3. During this period the membranes have not been chemically cleaned. The system performance has improved with a once/day permeate flush to remove colloidal material. There has been a ~10% decline in normalized flux, but it has not reached the point of needing chemical cleaning. Cleaning is usually recommended when flux drops 15% from initial values. Likewise the normalized salt passage is shown in Fig. 16.12b. This plot shows that the salt passage (or rejection) has been stable for the entire period.

Stable rejection is important for these applications because the water quality goals have both a maximum and minimum rejection limit (50–80 ppm hardness as $CaCO_3$). Any small trend of improving or worsening rejection will cause the product go out of specification in a short time. It is very important for plant operators to plot and track the normalized data for an operating system.

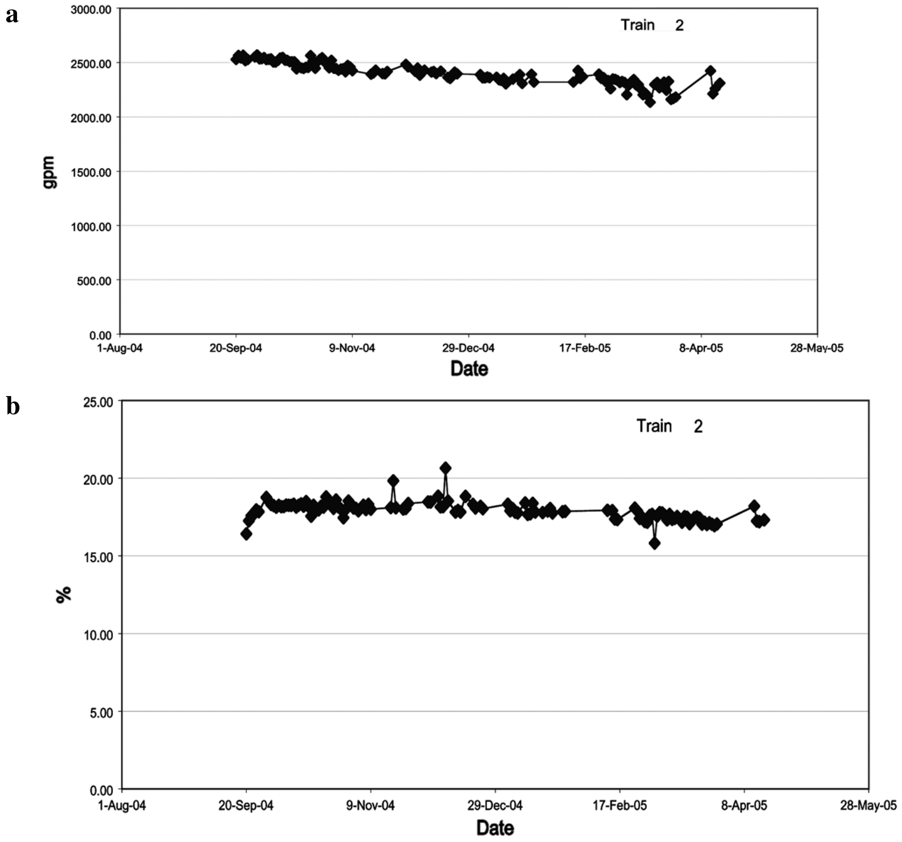


FIG. 16.12 Boca Raton train 2 performance a) Normalized permeate flow and b) Normalized salt passage.

This allows them to determine the fouling rate and when cleaning should be done. Most membrane companies provide software that will normalize the NF plant operating data. It also allows them to see when unwanted events occur that degrade the performance of the plant. The data should be used to follow normalized salt passage, permeate flow, and normalized pressure drop across the system. It is additionally beneficial to record the data for each stage individually. Then it is possible to see where the fouling is occurring. Fouling in the second stage will cause a decrease of the normalized permeate flow. This would often be an indication of scale formation. Such information helps the plant manager to target the type of cleaning that should be done. In summary, the Boca Raton NF softening plant shows that stable performance can be achieved for a high TOC feed water and the membrane can be tailored to meet a narrow permeate hardness range. These kind of process and product improvements results in substantial economic savings for the end user.

Color removal–Irvine Ranch. Another example of a nanofiltration application is the removal of color from water. The NF membrane required for these applications varies, depending on the salinity of the water. In some cases the water needs to be softened and color removed, such as the application described at Boca Raton. In other cases, the color needs to be removed, but the salinity should remain relatively unchanged. These applications require loose NF membranes. One such example is the plant at Irvine Ranch, CA. This NF plant treats deep well water which has a salinity close to 350 ppm, a color of >200 CU and produces 26,500 m^3/d of high quality potable water.

The feed and permeate quality of the water is given in Table 16.3. The target for the product water was Color <5 CU and salinity between 100–300 mg/L. Again, this project required a lengthy pilot study over many months. (83) Eighteen different NF elements were tested to select a membrane with the right properties. As a result, the engineers selected a loose NF membrane which could pass most of the salt, but reject the organic matter. As with most NOM, the molecular weight is too low to be highly rejected by UF membranes. The advantage of a loose NF over a conventional tight softening NF membrane is shown in Table 16.3 as well. Both membranes can achieve the target color values, but the loose NF has a permeate conductivity of 350 uS/cm compared to 48 uS/cm for the conventional. Thus, the permeate from the conventional NF would be more corrosive and the concentrate will have a much higher TDS and scale potential. The latter issue was quite important to the plant designers, since the plant was targeting 92% recovery.

The commercial plant was designed with 3 trains in a 36:18:8 array with 7 elements per vessel. The feed water is at a temperature of 33 C, with an SDI of 3.7, turbidity of 0.7 NTU and the operating flux is 26 lmh at a pressure of 6 bar.

TABLE 16.3
Summary of Irvine Ranch NF Plant Performance

Parameter	Feed	HYDRACoRe NF	Conventional NF
Color cu	200	<5	<5
Conductivity uS/cm	500	350	48
Calcium mg/L	13	8.5	0.2
Specific Flux Lmh/bar		10.6	11.8

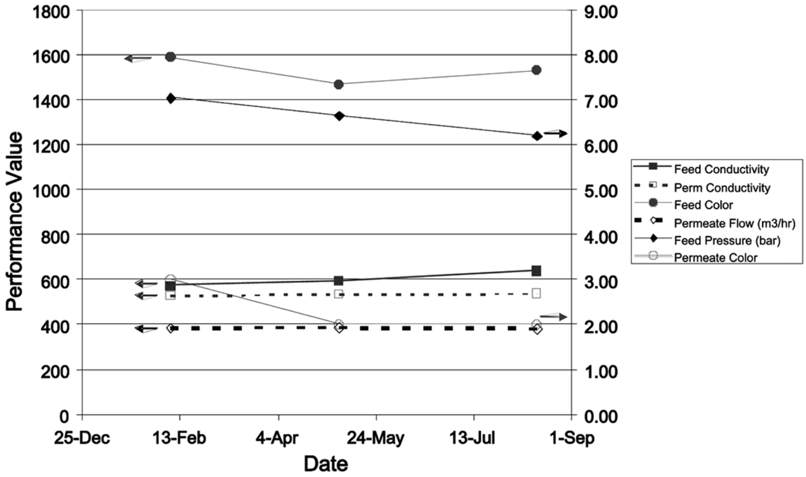


FIG. 16.13 Irvine Ranch Water District NF plant performance. Historical performance of the Irvine Ranch Water District Plant with HydraCoRe membrane.

The concentrate from this plant has conductivity of 1274 uS/cm and a color reading around 2835 CU, as seen in Fig. 16.13. This plant has also operated stably, requiring no cleaning in the first year of operation.

Pesticide removal—Mery Sur Oise. Another application for NF membranes has been to filter harmful pesticides from potential potable water sources in addition to the removal of organic matter. One example of this occurs in France at the Mery Sur Oise treatment plant. (84) This 140,000 m^3/d NF plant treats surface water from the Oise River near the city of Paris, France and blends with 50,000–210,000 m^3/d of conventionally treated water. Historically, the TOC levels in the river water vary, having values as high as 10 mg/l. The conventionally treated water has had an average TOC of 1.8 ppm, which requires a high level of chlorine in the distribution system. This imparts a unpleasant taste to the water. Additionally, the river has been contaminated with a variety of pesticides which need to be reduced from a few micrograms to a tenth of a microgram. The reason for NF at this plant was the goal of high organic and pesticide rejection while slightly lowering the hardness of the water. High rejection RO or NF would demineralize the water too much and make it corrosive against the lead containing distribution pipes. This requirement led to the development of a selective nanofiltration membrane which could achieve these targets.

The NF plant was installed and commissioned at the end of 1999. The design of the NF plant is shown in Table 16.4. The NF system design includes 8

TABLE 16.4
Process design for the Mery-sur-Oise NF plant

Plant	Mery-sur-Oise
Source	Oise River
Total capacity (m^3/d)	140,000
Pretreatment processes	Coagulation/Floculation/Settling/Ozone/ Media filter/Cartridge filter
Pretreatment chemicals	Antiscalant, Acid
NF trains	8
NF stages	3
Array	108 × 54 × 28
Elements/pressure vessel	6
NF membrane flux (lmh)	17.2
Feed pressure (bar)	10
Recovery (%)	85
Membrane element	NF200 B-400
Feed hardness (mg/l as $CaCO_3$)	300
Feed: permeate Atrazine (µg/l)	850 : <50
Feed: permeate TOC (mg/l)	~2.8 : 0.67

trains, each with a 3 stage design to achieve 85% recovery. The system has 9120 pieces of Dow NF200 B-400 membranes operating at 17.2 lmh.

The pretreatment is much more extensive than at the Boca Raton plant since it treats a high turbidity (15–100 NTU), high TOC surface water. The raw water is treated by coagulation, flocculation, settling, ozonation, dual media filtration, and cartridge filtration. The pretreatment system produces water with SDI <3 and particles <200/ml. The plant uses 0.2 g/m^3 of polyelectrolyte, 1.2 g/m^3 of ozone, and 5 g/m^3 of coagulant before the media filter.

The plant has operated for more than five years with the original membranes. These membranes have produced permeate with 90% rejection of pesticide and 50% reduction of conductivity. The amount of chlorine dosed in the water has decreased from 0.4 to 0.2 mg/l. The reduction of chlorine has resulted in a better tasting water, while the NF membrane has reduced the health risk of pesticides and provided a softer water.

Pollutants Removal–Metals and Other Inorganic Ions. There are other applications for using NF membranes to produce potable water. These generally consist of the removal of heavy metals such as radium, iron, manganese or arsenic, where the customer wants to pass some of the other salts in the water.

In a study of synthetic and natural surface water, it has been shown (85) that a NF membrane (NF-300 from Osmonics) could achieve 95% rejection of As V that was present as 100 Ìg/l in the feed. The permeate value of 8 Ìg/l met the regulation target of 10 Ìg/l, while TDS rejection was only 68%. Compared to a RO membrane, NF would have an advantage in this case of running at lower pressure, being less likely to cause scaling, and having less affect on product water stability.

Another example is the study that was done with NF membranes for removing perchlorate from a surface water. (86) This study of UF, NF and RO membranes found that the ESNA NF membrane had perchlorate rejection of 87–92%, the TFC-S NF membrane had perchlorate rejection of 97% and the LFC1 RO membrane had rejection as high as 99%, while the UF membrane had only 10% rejection at best. Naturally, the trend for chloride rejection was similar with LFC1>TFC-S>ESNA. This again shows the trade-off that is often present when selecting membranes. Selection of the best membrane will depend on the many factors specific to the project.

16.4.2. Industrial process fluid purification

For industrial applications, NF is being used in a growing number of applications because of its selective rejection properties. There are a broad range of applications from wastewater treatment to product recovery/purification. These applications must be pilot tested to properly select a NF membrane and to optimize the process. In addition these applications also have challenging operating environments such as high temperature, high or low pH, and solutions with aggressive organic constituents. A few representative applications are described below.

Dye waste. The textile industry requires large amounts of water and generates large volumes of wastewater. Plants typically use activated sludge treatment, but the effluent from this conventional treatment is often colored, thus requiring more treatment and generating more waste. Recently, engineers have been turning to membrane technology to more effectively clean the waste and possibly recover some of the wastewater for reuse. (87) The dye wastes that are generated in these processes contain as much as 20 g/l of NaCl and dye concentrations of 100–500 ppm.

Experiments have shown that the membrane can be operated at a flux around 25 lmh with a pressure of 200 kPa. The Koch TFC-SR2 NF membrane had dye rejection of 97% and showed that the flux could be restored after several uses. The low pressure was achieved because the NF membrane passed most of the NaCl. This again is the key feature of the membrane and governs which membrane should be selected for this application. The final purified water did not have color and was suitable for reuse.

Offshore reinjection fluid. One of the larger applications for NF membranes in industrial applications is the treatment of seawater for re-injection practiced by the petroleum industry. In this application off-shore oil platforms need to inject water into the oil-bearing formations to enhance the oil recovery. The seawater which they treat and inject contains high levels of sulfates, while the water in the production well will contain a saline water that often has high levels of barium. There is a proven concern that the sulfates from the seawater will mix with the barium-containing produced water in the well and result in the formation of barium sulfate scale. This insoluble scale can then plug a producing well and reduce the oil output. Many facilities have opted to use the patented process (88) which takes advantage of NF membranes to remove sulfate from the seawater before injecting it into the well. This type of application requires a NF membrane which can selectively reject divalent anions while passing monovalent ions.

In this application the raw seawater is treated by some prefiltration technology to reduce the turbidity to <0.2 NTU. The filtered seawater contains sulfate values around 2500–4000 ppm and TDS values of 36,000 to 43,000 ppm. A typical feed water is shown in Table 16.5. (88) Of particular interest are the sodium and chloride concentrations. To operate the NF process at low pressures, the NF membrane must have a high passage rate for the chloride. This is indeed achieved with the Dow NF45 membrane, as illustrated in the patent. The typical results for a 75% recovery NF system are shown in Table 16.5. The particular advantage of membrane process is that it is compact and has relatively low weight and low chemical consumption compared to other sulfate removal technologies. This is very important for a remote offshore platform.

The feed chloride concentration was 19,750 ppm while the permeate value was 19,000 ppm, thus having only 4% rejection of chloride. This keeps the osmotic pressure to a minimum, and allows the NF system to run at low pressure, 15.9 bar in this case. The optimum flux rate depends on the pretreated water quality, but typical values are around 24 lmh. The system is designed with 2 stages operating at 75% recovery. Operation at higher recovery will significantly increase

TABLE 16.5
Water quality for a seawater NF plant

Ion	Feed water	Product water	% Rejection
Sodium	11,200	10,688	5
Potassium	370	320	14
Calcium	400	309	23
Magnesium	1,400	330	76
Chloride	19,750	19,000	4
Sulfate	2,650	48	98
Bicarbonate	140	20	86
Total	35,910	30,715	

At 25°C and 75% recovery

pressure and will increase the sulfate levels in the permeate. In the future, NF membranes with higher SO_4 rejection and lower Cl rejection would make it possible to run at higher recovery and thereby decrease the size of the pretreatment. The sulfate ion is reduced to less than 50 ppm from the initial 2650 ppm, or 98% rejection. The concentrate sulfate concentration would be approximately 10,600 ppm. It is generally believed that the sulfate values need to be reduced below 100 ppm to prevent $BaSO_4$ formation. The actual sulfate concentration allowable will depend on the site, but it is expected that even lower values of sulfate can be achieved.

In a related application researchers have been looking at the use of similar NF membranes to precondition seawater for further RO treatment or as pretreatment for multi stage flash (MSF) distillation (89, 90). A number of reports have been produced on this subject, and there is at least one commercial plant operating in this mode (91). The desalination plant at Layyah has a historical capacity of approximately 159,115m^3/d. Five units employ the multi-stage flash evaporation process. Recently, one of the original MSF units was rehabilitated and expanded, which included the addition of a NF pretreatment process. By lowering the sulfate and calcium concentration of the feed water, the MSF could then operate at higher temperatures without concern about controlling calcium sulfate scale. The use of NF treated water resulted in a water production increase from 22,700 m^3/d (5 mIgd) to 32,730m^3/d (7.2 mIgd). (19)

There are still many questions about the economics of this approach, especially for NF as a pretreatment to SWRO. With NF pretreatment, the salinity to the SWRO will be reduced and allow it to run at lower pressures and run at higher recovery. However, the combined recovery of the NF + RO will be less than just a SWRO process. For example if the data in Table 16.5 is used, the feed to the SWRO is 30,715 ppm. This can be run at 59.4 bar (862 psi) and 55% recovery with current state of the art SWRO membranes. Thus the NF-RO process overall system recovery is 0.75 × 0.55 = 41.2%. The same SWRO membranes can be run on the feed to the NF at the same pressure 59.3 bar (860 psi) by lowering the recovery to 51%. Therefore, the overall recovery is better at 51% versus 41% and the size of the pretreatment will be less in the latter case and the SWRO system would be more economical. Additionally, the NF-SWRO and the SWRO system make the same permeate quality (253 ppm) due to the differences in the recovery. A more thorough investigation of NF-SWRO versus SWRO (93) has concluded that only in a very specific case would the NF-SWRO be more cost effective. This case—no energy recovery, permeate recovery below 35%, and high TDS water—is rarely found in commercial plants these days, and would thus indicate such a process will not be cost effective for commercial plants. In all other cases that were evaluated, the single SWRO process was more cost effective, by more than 20 percent.

Recently, researchers at Long Beach Municipal Water District have evaluated a NF-NF system for seawater desalination (93). The proposed approach is to use a complex two pass NF system to make potable water. The study found that the process could produce permeate with less than 500 ppm TDS, operate at less than 550 psi in the first stage and 250 psi in the second stage. However, there are questions about the economic viability of this approach as well (92). In this work, the membranes which were used could be more accurately described as low pressure RO membranes, since they have salt rejections around 99% or better. Such rejections are needed to reduce a TDS of 37,000 ppm seawater to less than 500 ppm at reasonable recoveries. Further scale-up of this concept is in progress; a 1900 m^3/d (500,000 gpd) system is being constructed to compare the NF-NF and SWRO concept side-by-side.

Paper mill industry. This is another industry that requires large volumes of water, generates copious amounts of wastewater and would benefit from a low-energy membrane reclamation process. The wastewater from these plants is highly contaminated with salts and a variety of organic materials. UF membranes have been used to clean up this wastewater, but NF is needed to give quality that is adequate for reuse (94). A few NF plants have been used to reclaim such wastewater. One of these has a capacity of 10,000 m^3/d. These plants however require extensive pretreatment to make the water suitable.

Recent tests with the Dow NF270 membrane showed that high removal rates can be achieved for TOC and color while low pressure is required. However, the rejection characteristics were found to vary with pH, which would impact the selection of the treatment process. Also, testing showed that the treatment temperature must be limited to less than 50°C. This also posed some restrictions, but in general the membrane showed promise because of its ability to be cleaned once fouled. Further research and testing should open additional opportunities in this industry.

16.4.3. Other applications

There are a variety of other industrial applications for which NF membranes can be used, including dairy wastewater (95), electroplating industry (96), lubricant oil purification (97), and chloralkalai treatment (98) to name a few. Another set of examples of applications is where the NF membrane is used to reject and recover valuable product from a process stream. These applications would include sugar fractionation, drug and other specialty chemical recovery, metals recovery, monomer purification, and many more.

16.5 Conclusions

This chapter does not contain an exhaustive list of all NF applications, but rather has tried to give guidelines for how the properties of NF membranes can be utilized in a commercial process. It has also given general guidelines for how to use a NF membrane in an effective treatment process. When these guidelines are followed, there is a strong chance for success. The extent to which NF membranes are used depends to a great degree on the education of process engineers who are designing processes and the availability of useful tools to evaluate potential NF solutions.

17

Wastewater treatment and reclamation by RO and NF process

*Craig Bartels**

17.1 Introduction

Increasing growth in heavily populated regions of the world has strained available water supplies in many regions (99). There are a limited number of options that these communities have to supplement their current water resources, but these generally involve the use of lower quality water and require more extensive treatment. Alternatively, these communities may import water from more remote sites, but high transportation costs also make this an undesirable choice. Aside from importing new water or improved conservation, water resources are typically supplemented by methods such as the treatment of brackish well water, seawater desalination, or wastewater treatment and reuse (120). The latter option can be a very economical approach due to abundant and convenient supply of wastewater, and the relatively low salinity of most wastewaters. The salinity of wastewater usually is in the range of 800–1500 mg/L of total dissolved solids (TDS), but repeated reuse of the water will cause salinity to continue to rise; therefore, desalting is required. This is much lower salinity than brackish wells which may range from 1500 to 3000 mg/L of TDS, and seawater which will be around 32,000 to 40,000 mg/L of TDS. The energy required to treat brackish well water will be about two times that of wastewater, while about six times of that for seawater. However, seawater is plentiful, while

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brackish wells can often be overdrawn, and thus have reduced production. Recently, many communities have opted to reclaim wastewater for uses such as irrigation, industrial water or barrier protection of groundwater sources in coastal regions (100, 101). For example, two large treatment plants greater than 30,000 m^3/d, have been built in Singapore to reclaim municipal wastewater, and there is a 312,000 m^3/d RO-based wastewater treatment plant in Kuwait. Although it can be a difficult water to treat, municipal wastewater is abundant and generally located near populated areas.

Likewise, plant operators are finding that it is difficult to obtain high quality water needed for their industrial processes. Advanced treatment of municipal wastewater can be a viable alternative to procure a source of high quality water, while others opt to reclaim their own wastewater, and thus decrease their disposal costs as well. The latter option can be quite difficult, depending on the nature of their wastewater, and it may not be consistently available.

17.2 Background

Since the largest volume wastewater applications involve municipal sources, the majority of this chapter will focus on issues related to the treatment of municipal wastewaters. Such waters can be challenging to treat with reverse osmosis (RO) due to the high organic material loading, high biological activity, excessive colloidal material, and high potential for scaling. Traditionally, raw wastewater is treated by coagulation and primary treatment to remove large solids, followed by secondary treatment which is the activated sludge (Fig. 17.1). In this step organic matter is biologically degraded to CO_2, H_2O, and possibly ammonia by bacteria. The water then goes through a coagulation and settling phase to clarify much of the remaining solids from the water. The effluent from this treatment, called the secondary effluent, is still high in suspended solids and cannot be treated directly by RO without excessive fouling rates. Typically, the secondary effluent is filtered by various means, including sand filters, lime clarification, microfiltration or ultrafiltration. The type of pretreatment has a significant impact on the design and operation of the RO system. Much of the original work on wastewater treatment with RO utilized the conventional filtration as pretreatment.

Previous papers have documented the high fouling rates that can be observed for typical systems using RO membranes when treating conventionally

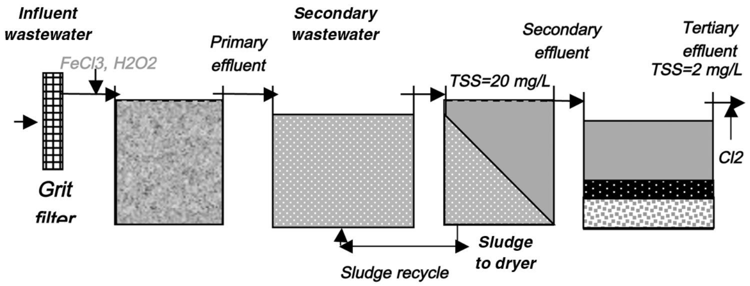


FIG. 17.1 Typical municipal wastewater treatment process.

filtered municipal wastewater. (102, 103) In extreme conditions, the pressure can increase from 13.8 bar (200 psi) to 18.1 bar (260 psi) in a matter of days to maintain constant permeate flow. Such fouling rates will make the RO system uneconomical for this application.

For example, the original wastewater treatment plants at Orange County Water District (OCWD) and West Basin Water Recycling Plant utilized lime clarification and media filters, while the Jewell plant (SUT Seraya-High Grade Industrial Water) in Singapore used dual media filtration as pretreatment to RO. Each of these had operated with some success, although significant membrane fouling required frequent membrane cleaning.

Meanwhile, extensive pilot and demonstration studies at OCWD and other sites, have demonstrated the value of using microfiltration or ultrafiltration pretreatment to improve the operation of the RO process (104). The improved performance has been mainly attributed to the improved removal of fine particles and minimization of "composite" fouling (foulant layer consisting of colloidal and organic matter). They estimated that a membrane pretreatment/polyamide RO system would have an O&M cost that was 40% less than a conventional clarification/cellulose acetate RO system. Also, the new design would take only 20% of the space to make comparable flow.

The success of treating wastewater effluents by reverse osmosis has also made significant gains recently due to the improved performance of the reverse osmosis membranes and improved process design. By adhering to these new guidelines and selecting the optimum membrane, wastewater treatment by RO can prove to be an economical, long-term solution for drought proofing water-short municipalities.

17.3 Membrane selection

There are many membranes to choose from, ranging from conventional polyamide membranes to low fouling membranes. The choice of the optimum membrane is a difficult matter that is best done by pilot testing.

The earliest commercial plants used cellulose acetate (CA) membranes. These membranes needed cleaning every 4–6 weeks and were characterized by high pressure requirements and relatively poor permeate water quality (Fig. 17.2). (105) With the advent of the composite polyamide membrane, the pressures were lowered and the water quality improved (Fig. 17.2), however, there were still frequent issues with fouling. The one advantage of CA membranes was their resistance to chlorine, but they were very sensitive to pH extremes, which somewhat limited cleaning options.

More recently, low fouling membranes became available in the late 1990s. The nature of these is not fully revealed by the manufacturers, but some information has been provided to understand the properties of these membranes. One type of low fouling membranes is characterized by having a surface that is less charged (more neutral) and more hydrophilic. Conceptually, this was thought to prevent the attachment of organic materials, as well as bacteria, and thereby minimize the fouling rate. A list of commercial low fouling membranes is given in Table 17.1.

A variety of approaches can be used to prevent various fouling mechanisms. The LFC membrane, utilized for the Bedok wastewater treatment demonstration facility, has a similar topography as a conventional polyamide

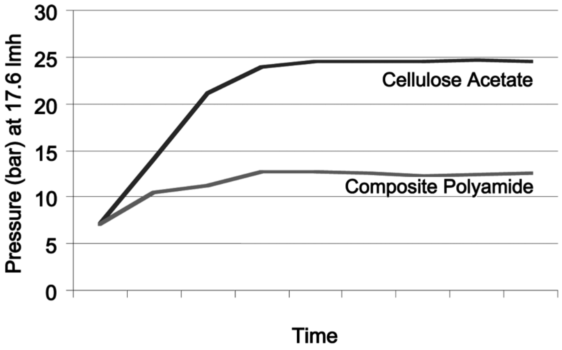


FIG. 17.2 Comparison of cellulose acetate and polyamide membranes treating wastewater at a commercial wastewater treatment plant.

TABLE 17.1
Performance characteristics of commercial low fouling polyamide RO membranes. Test pressure 15.5 bar (225 psi)

Product type	Salt rejection (%)	Nominal flow (m^3/d)	Nominal flow (GPD)	Element area (m^2)	Element area (ft^2)	Test solution/ concentration (ppm)
Dow BW30-FR	99.5	39.9	10500	37.2	400	2000 ppm NaCl
Hydranautics LFC1	99.5	41.8	11000	37.2	400	1500 ppm NaCl
Hydranautics LFC3	99.7	36.1	9500	37.2	400	1500 ppm NaCl
Saehan RE8040-FL	99.5	38.0	10000	33.9	365	2000 ppm NaCl
Toray TML20-400	99.7	38.8	10200	37.2	400	2000 ppm NaCl
TriSep 8040-X201-TSA	99.5	34.2	9000	37.2	400	2000 ppm NaCl

membrane, but the chemical nature of the surface has been modified to reduce the typical negative charge at the surface and increase hydrophilicity. The surface is essentially neutral at the normal pH of the wastewater, and it is hydrophilic. (106, 107) This combination is specifically designed to minimize the interaction of organic molecules with the membrane surface. A test was carried out to characterize the surface of an LFC and a conventional polyamide membrane. The initial flux of both membranes was measured, and recorded as J_o. The membrane was then exposed to a water soluble organic surfactant that was positively, negatively, amphoterically, or neutrally charged. The membrane was exposed for 12 hours in 1000 mg/L solution of the surfactant. The membrane was then removed from the solution and rinsed with RO water for 3 hours. The flux of the membrane, J_f, after being exposed to the surfactant solution was measured again and the flux retention, J_f/J_o, was calculated. The higher the value of J_f/J_o, the better the flux retention, and the less likely the membrane will be significantly fouled by the water soluble organic. The results of all tests are shown in Fig. 17.3 and demonstrate that the LFC membrane is significantly more stable to cationic, amphoteric and neutrally charged organic species. The response of both membranes was exceptionally good to anionic compounds as expected. These organics tend to be repelled by the negative charge of the polyamide, and are not strongly attracted to the neutral charge of the LFC surface.

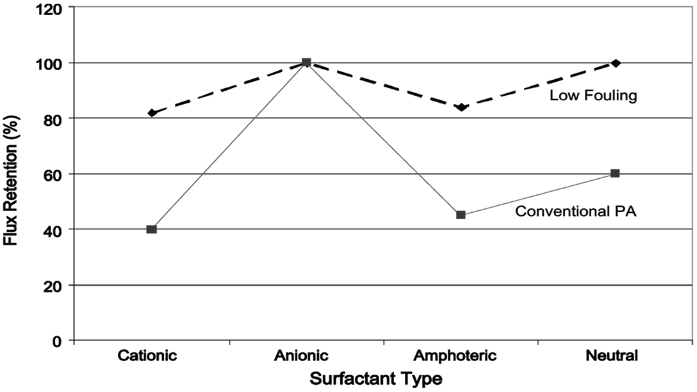


FIG. 17.3 Effect of surfactant fouling on conventional and low fouling membrane.

Likewise, Saehan (108) has recently patented a low fouling membrane chemistry (108) which involves coating a hydrophilic, polyfunctional epoxy layer onto the surface of a polyamide membrane. This layer is crosslinked to make it permanent. They show that this layer is permanently attached to the membrane surface and reduces protein fouling (from dried milk) by about 50%.

Other low fouling membranes claim different features which render the membrane fouling resistant. The Dow BW30-FR membrane is believed to have a modified polyamide surface, which reduces the adsorption of biofilm polymers (109), while the TriSep X20 claims that the unique polyamide-urea membrane layer does not have the carboxylate groups of a traditional polyamide membrane, and thus is more neutral (110).

In addition to selecting a membrane that has less affinity for organic compounds, membranes with various flux and rejection can be chosen. In general, the required quality of the water will determine which membrane is optimum for the application. A user will select the membrane which has the rejection that will meet the quality of the product water specification, while giving them the lowest operating pressure. As always, there is a trade-off in membrane science where higher rejection usually results in lower water permeability. Reference to Table 16.1 gives the engineer some guidance to select the appropriate membrane; however, it is always best to use projection software provided by the vendor to estimate the performance for a full system.

One emerging requirement for wastewater treatment plants is the need to meet water quality objectives for multiple ions and organic compounds, not just the overall TDS. For this purpose, an accurate projection program and/or pilot testing is required. Some of the components which are being regulated and need

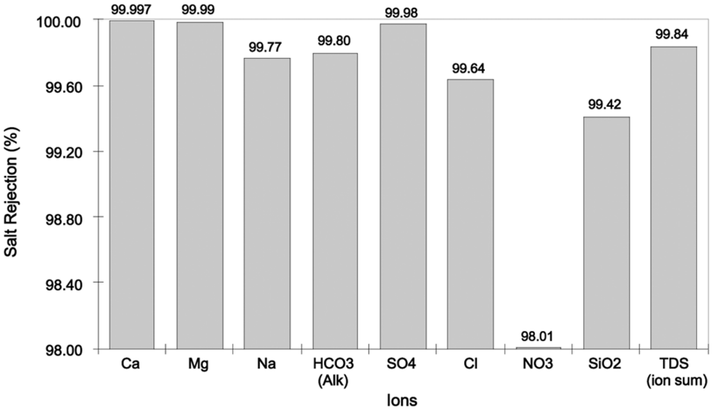


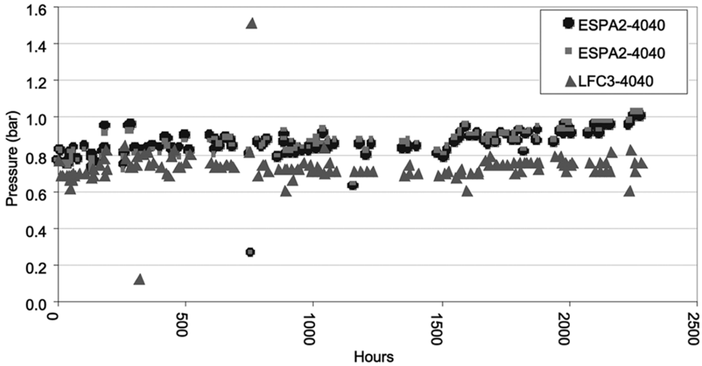
FIG. 17.4 Ion rejections for a commercial low fouling membrane. Feed concentration 2076 mg/L, Feed temperature 21 C, and Flux 33 lmh (19.5 gfd).

specific removal rates include, hardness, iron, ammonium, boron, silica, nitrate, chloride, endocrine disruptors, and general TOC. Fig. 17.4 gives the measured rejection of a variety of ions measured for the LFC3 membrane.

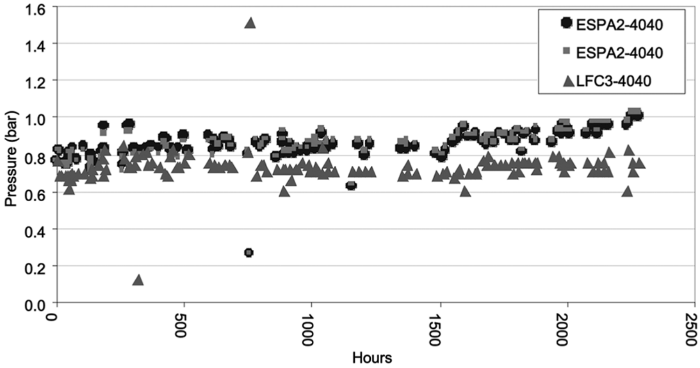
It should be noted that the neutral, hydrophilic charge of the LFC surface is designed principally to minimize organic adsorption. If the water is high in colloidal matter, this membrane will foul at about the same rate as a conventional polyamide membrane, because such fouling is due to the hydrodynamics of the system. The flux rate versus crossflow rate will determine how rapidly solids are brought to the surface and thus, collect on the membrane surface. This type of fouling can only be minimized by reducing flux rate or by increasing crossflow velocity; however, studies have indicated that the nature of the LFC surface makes it slightly less susceptible to biofilm formation. (111) This is likely due to the reduced adsorption of the organics to the membrane which provide a food source for the bacteria.

Recently, further research has shown that in some circumstances, conventional polyamides may be as effective as the low fouling membranes for treating municipal wastewater. In one paper (112), they report a side-by-side pilot study where MF or UF pretreated wastewater was provided to a 2:1 array of 4 inch RO elements. One vessel in the first stage had a conventional polyamide membrane (ESPA2) while the second stage contained the high rejection low fouling LFC3 membrane. All element types were operated at the same flux rates. The results show (Fig. 17.5) that the low fouling LFC3 had slightly more stable normalized flow, and the conventional polyamide membrane had the lowest pressure (highest specific flux).

a. Normalized specific flux



b. Normalized pressure drop



c. Normalized salt passage

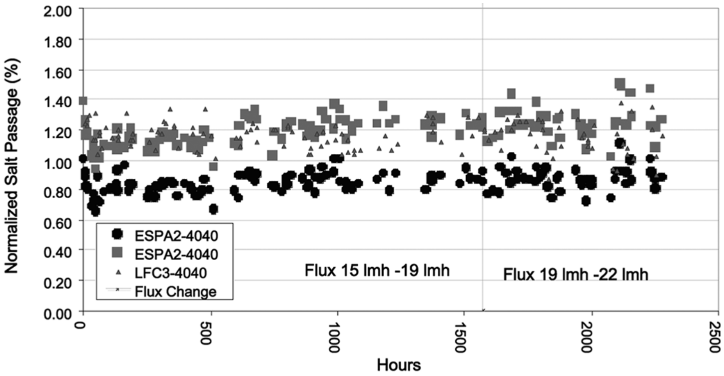


FIG. 17.5 Performance of conventional and low fouling membranes treating membrane pretreated wastewater.

In a contrasting report (113), researchers have reported studies with conventional and low fouling membranes which suggested that some conventional polyamide membranes were equal to or superior to low fouling membranes. Their studies were done at commercial wastewater treatment plants in which a number of vessels were replaced with new membrane of different types from different membrane suppliers. A variety of fouling rates were seen depending on the product type and the wastewater treatment plant, but statistically, there appeared to be no advantage from using the low fouling membrane. In some cases they note that the low fouling membrane had worse performance compared to traditional polyamide membranes. The data did indicate that there was often an initial period of fouling in which the initial membrane permeability dropped. This drop, which is attributed to organic adsorption, was significantly different for the various membranes. Some membranes were able to achieve stable operating flux rates for an extended period of time. An example of one of the studies is shown in Fig. 17.6.

It is still difficult to fully interpret all of these results. Certainly, it can be said that under the proper operating conditions, it is possible for some conventional polyamide membranes to operate as well as or better than low fouling membranes. This may be in part due to the nature of the fouling that is seen in state-of-the-art RO wastewater plants. If the main mode of fouling is due to the presence of negatively charged humic acid organic material, the negatively charged polyamide surface may be good at repelling these organic materials, as would be expected based on Fig. 16.1. However, if the fouling is due to neu-

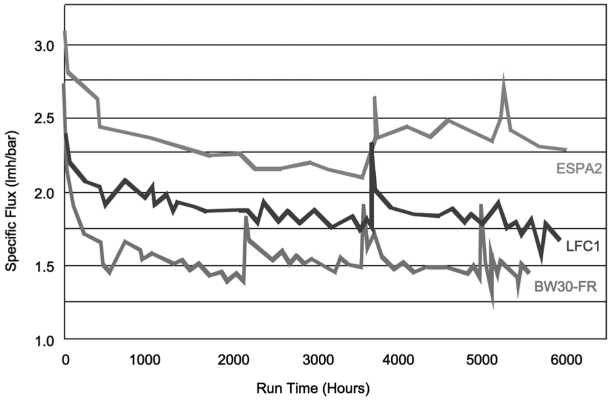


FIG. 17.6 Performance of conventional and low fouling membranes at a commercial wastewater plant (113).

trally charged polysaccharides from bacteria or other types of non-charged or positively charged organics, one would still expect the low fouling membranes to be superior. Based on these findings, pilot testing is highly recommended to select the best membrane and process conditions.

17.4 Wastewater reclamation RO process design

As already described, the use of ultrafiltration or microfiltration will greatly reduce colloidal fouling of the RO membrane, and new membranes are available which are more resistant to adsorptive organic fouling. These two improvements have made RO membrane technology effective for wastewater applications, but there are other design features which are critical for optimized RO performance treating wastewaters. A state-of-the-art wastewater treatment process is shown in Fig. 17.7.

17.4.1. Permeate flux rate

Flux rates for surface water treatment are generally in the range of 17 to 25 lmh (10–15 gfd), depending on the quality of the source water. For wastewater, flux values lower than this range have been used to minimize the fouling rate when conventional media pretreatment is used. However, this results in higher capital costs and lower permeate quality. Multiple studies (102, 113, 115, 116) have shown that higher flux rates can be achieved with stable performance when using MF or UF pretreatment. With such pretreatment, flux rates of 15–20 lmh (9–12 gfd) can be used. One study (112) showed that the fouling rate of the RO was about the same whether UF or MF was used as the pretreatment. Even with these higher flux rates, fouling rates are lower due to the reduced colloidal fouling which is otherwise prevalent with wastewater treatment. As previously

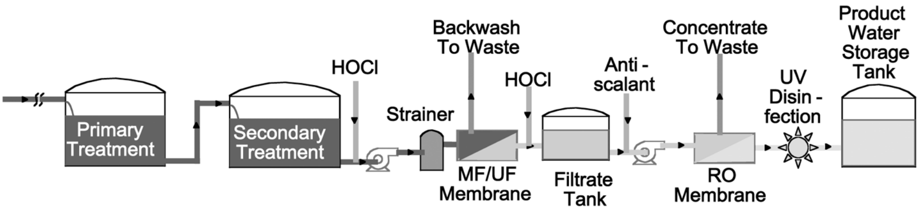


FIG. 17.7 Typical wastewater treatment process.

explained above, there appears to be an initial decline in permeability, followed by a period of stable permeability. This is likely due to the adsorption of dissolved organic matter present in wastewater. Once the adsorption sites on the membrane surface are saturated, the membrane operates at a stable permeability.

17.4.2. Biofouling control

Biofouling is a common occurrence for RO membranes when treating a bioactive wastewater or surface water. This can result in frequent cleaning and thus greater operational expense. A variety of strategies have been employed to minimize this problem. For conventional surface water treatment, chlorination in pretreatment followed by dechlorination has often been used, but typically bacterial regrowth can still cause rapid fouling. This type of treatment can be detrimental due to chlorine breaking down the organic substance and creating a more bioassimible food source for the surviving bacteria. Also, chlorine can damage the membrane at a dose of a few ppm over long operational periods.

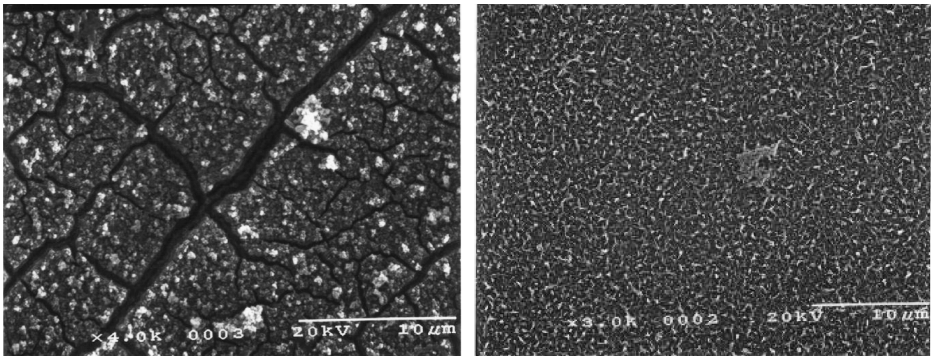
For wastewater treatment plants, however, most plants utilize chloramines as a biostat to prevent the growth of the bacteria in the RO membranes. Chloramine is a much weaker oxidant than free chlorine, and thus the membranes can tolerate 30,000–100,000 ppm hours of chloramines exposure. Additionally, chloramines are easily formed in wastewater due to the abundance of ammonia in the water. Chlorine when added to the wastewater, will rapidly form chloramines. (116) Some sites use dual dosing sites, opting to use the chloramines to control biofouling in the UF/MF pretreatment, followed by a second dosing between the UF/MF and the RO. This is done to carefully control the chloramines concentration, which may drop after the initial dose due to the oxidant demand of the organic matter in the wastewater. Careful control is needed to ensure chloramines concentration to the RO, since even chloramines will more quickly damage the RO membranes when dosages exceed 5 ppm for an extended period. Regardless of the dosing location, the chloramine concentration to the RO should be controlled between 2 and 4 ppm, and not exceed 5 ppm for maximum product life. Historical experience shows that biofouling has never been a significant issue for plants that utilize such biofouling control strategy. However, some wastewater plants opt to remove the nitrogen from the water, which means there is no ammonia present to form the chloramines. In such cases, ammonia should be dosed as well as the chlorine. One way to do this is the addition of ammonium chloride. Typically, ammonia to chlorine ratio should be 3–5 to ensure the absence of free chlorine. In such plants, it is essential that monitors (ammonia flow transmitters, oxygen reduction potential (ORP) meters, etc.) be installed to prevent unwanted presence of free chlorine to the RO membranes.

17.4.3. Recovery rate

Water recovery rates of 75–85% are typically achievable, the actual value depending on sparingly soluble salts in the feedwater. As always, higher recoveries are desirable, since they minimize the size of the pretreatment system and maximize the use of the source water. Although the salinity of wastewater is not typically high, it can be high in certain ionic species which can scale the membrane, or dissolved organic compounds which can more likely adsorb on the membrane surface as they are concentrated. The latter can be combated by using membranes that do not readily adsorb the organic species or by adding coagulants in front of the UF/MF to adsorb the organics. The inorganic scale can only be controlled by chemical addition or reduced recovery. Chemical addition can be costly, while reduced recovery increases the cost of the pretreatment system. Of the many plants that have treated wastewater, the most common scaling problem has been the formation of calcium phosphate. Phosphates are common in wastewater due to detergents used by domestic households. Additionally, the amount of phosphate can vary significantly. This poses a significant problem for membrane treatment.

For example, an analysis was made of a fouled LFC1 element from a wastewater treatment RO plant. The element operated for about one month in the last position of the third stage vessel. The abstracted element was tested under normal conditions, 15.6 bar, 7.7 m^3/h brine flow rate, and 1500 ppm of NaCl. Compared to the initial test, the flow of the element had dropped 58%, while the salt passage had increased 250%. The element was cut open and membrane was extracted for surface examination. A SEM picture of the fouled and clean membrane surface is shown in Fig. 17.8. It is apparent that the membrane was heavily fouled, since the membrane surface cannot be observed. What can be seen is an amorphous layer of material which has formed cracks or fissures, which likely occurred during the vacuum drying of the sample. A sample of the foulant was removed for Energy Dispersive Analysis by X-ray (EDA X), to determine the nature of the foulant. The results are shown in Table 17.2.

From the Table 17.2, it is possible to identify a few unique foulants. The presence of phosphorous is certainly associated with phosphate, PO_4. Thus, this



a. Fouled membrane surface b. Clean membrane surface

FIG. 17.8 SEM micrograph of a (a) fouled LFC1 membrane from the Bedok demonstration plant and a (b) clean membrane surface.

would also account for all of the oxygen present. The counter ion is likely to be calcium, as the solubility of calcium phosphate can be readily exceeded. However, there is not sufficient calcium to match stochiometrically with the phosphorous. Alternatively, some of the phosphorous on the membrane surface is likely associated with other sources, such as organic phosphorous compounds, such as phosphonates. The calcium on the surface was not present as calcium carbonate, since the foulant did not bubble when exposed to acid, nor was the calcium present as $CaSO_4$, since there was no sulfate present. Thus, it is likely that the calcium present was associated primarily with calcium phosphate. If this is the case, then we expect that 4% of P and 16% of O is present as phosphate; the remainder being organic phosphorous. The latter may come from the antiscalant used at the site or other materials in the wastewater.

TABLE 17.2
EDAX Foulant analysis of membrane from a wastewater treatment plant

Element	Atomic %
Carbon	24.9
Oxygen	42.6
Aluminum	5.2
Phosphorous	12.4
Calcium	6.2
Iron	4.3
Other	4.4

The other major element found was carbon. This is common for foulants and is likely derived from the organic material present in wastewater. This was not further analyzed. A small amount of aluminum was present, suggesting the presence of aluminum hydroxides, oxides or possibly aluminum silicates. This would most likely be associated with colloidal material in the feedwater, but not removed by the MF pretreatment.

In general, it is apparent that phosphate scale and organic compounds were the major contributors to fouling in this element. This is common at other wastewater treatment sites as well. A typical wastewater may have 5 to 50 ppm of phosphate in the feed along with 20–100 ppm of calcium. When the water is concentrated by 4–7 times, the phosphate scaling index can easily exceed saturation levels. The solubility of phosphate scale is given by the following equation:

$$K_{sp}Ca_3(PO_4)_2 = [Ca]^3[PO4]^2/[Ca_3(PO_4)_2]$$

where K_{sp} is the solubility product for calcium phosphate. This value can then be used to calculate the Scale Index Factor which also includes pH and temperature dependence of calcium phosphate scaling. In general, when this factor exceeds 100, there is an expectation that scaling will occur, unless controlled by a suitable antiscalant, while values in excess of 1000 are considered to be high risk for scaling and not controlled by antiscalant. Increasing concentrations of calcium or phosphorous, increasing pH, or decreasing temperature will all increase the formation of calcium phosphate scale. Calculations were made to determine the phosphate scaling Index Factor for typical wastewater using software from French Creek™. An example of a typical wastewater is given in Table 17.3. For this water the Scaling Index Factor at pH 7.1 and 30°C is 406. At 85% recovery, the phosphate and calcium are concentrated 6.6 times, thus the scaling index is higher and greatly exceeds the guideline of <1000. Table 17.4 shows the calculated saturation index for the above feedwater at various pH, temperature and recoveries. At 85% recovery, a pH as low as 4.5 is required to achieve a reasonable Scaling Index Factor. It demonstrates the significant effect of recovery; even the pH adjustment does not effectively control phosphate scaling for such high values.

Therefore, wastewater treatment may require one or more of the following strategies to control scaling: use of antiscalant designed specifically for calcium phosphate scale, lowering the feed pH, dephosphonation, or lower recovery. This is an important analysis that must be made for each wastewater that is to be treated, along with the normal assessment of traditional scales such as silica,

TABLE 17.3
Bedok wastewater feed analysis (all values in mg/L)

Cations		Anions	
Calcium	34.1	Total alkalinity	69.3
Magnesium	22.7	Sulphate	80.5
Sodium	153.2	Chloride	228.6
Potassium	17.5	Fluoride	0.91
Barium	0.02	Nitrate	35.8
		Silica SiO_2	8.3
Other		Phosphate	8.9
TOC	11.4	Boron	0.09
pH	7.5	TDS (ion sum)	730.3

TABLE 17.4
Bedok wastewater calcium phosphate scaling index for various operation parameters.

Temperature (°C)	Ca (mg/L)	PO_4 (mg/L)	pH	Recovery (%)	$Ca_3(PO_4)_2$ Scale index*
30	34	8.9	7.5	75	2547
30	34	8.9	6.5	75	12
30	34	8.9	7.5	85	29037
30	34	8.9	7	85	4448
30	34	8.9	6.5	85	283
30	68	8.9	6.5	85	1425
30	34	17.9	6.5	85	796
15	34	8.9	7	85	2577

*Scaling indexes based on French Creek software calculations

calcium fluoride, calcium carbonate and calcium sulfate. An economic assessment of these options will be made later in this chapter.

17.5 Commercial plant data and design

17.5.1. Singapore wastewater treatment plants

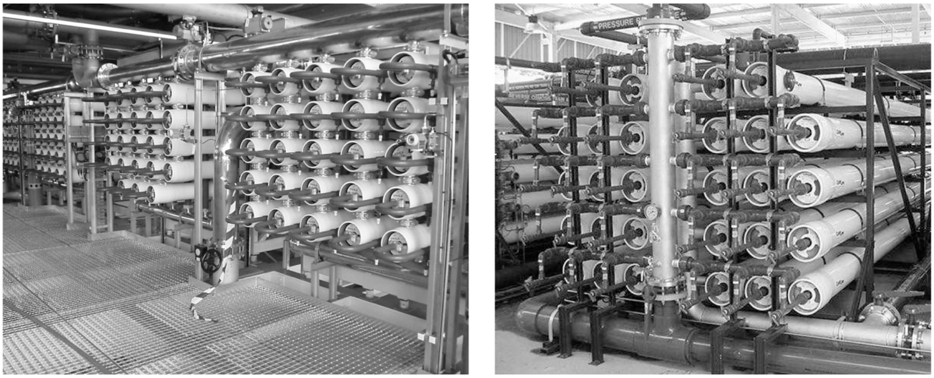
The Bedok demonstration plant has been treating wastewater with a membrane process since the year 2000. This plant is a 10,000 m^3/day plant that uses microfiltration to treat secondary effluent followed by RO treatment with LFC1 membrane. The elements are arranged in three stages with 30:14:8 vessels per stage, respectively, each with 6 elements per vessel. The plant initially operated at 85% recovery, which was later reduced to 81% to minimize fouling in the third stage. The average flux rate was 18.6 lmh (11 gfd) initially, and later reduced to 17.8 lmh (10.5 gfd) after increasing the first stage from 28 to 30 pressure vessels. Temperature was consistently around 30°C. The plant uses a two-point dosing of chloramine, one prior to the MF system and a second just prior to the RO system. The chloramine dose to the RO is adjusted to 2–3 mg/L.

After some initial changes in recovery, flux rate and antiscalant, the plant has operated consistently. During the first few years of operation, the first and second stage elements were only cleaned once every six months. The third stage has been cleaned every 3–4 months on average. The fouling has been most problematic in the third stage, which is typical for scale formation when recovery is high. The general absence of bacteria as verified by the SEM micrographs of the third stage element and the low pressure drop for the first stage, indicate that biofouling is not a significant problem for this site. This verifies the effectiveness of the chloramine dosing to control biofouling. However, some systems have been known to have biofouling issues. This can generally be attributed to by-pass of the pretreatment membranes, inadequate dosing of chloramines, dead-legs in the piping which allow significant biogrowth, or poor disinfection of the system prior to operation.

Additionally, the membrane has shown stable rejection during periods of controlled operation, suggesting that the chloramine dosing has not degraded the membrane. Over the first two years of operation, the membranes were exposed to an approximate dose of 8400 ppm hours of chloramines. In addition the RO system has had greater than 6 log removal of bacteria and greater than 4 log removal of viruses.

In November of 2002 two new full-scale plants were completed and began operation in Singapore. (Fig. 17.9) These two plants were designed with the optimum features gathered from the Bedok demonstration plant.

One plant was located at the Bedok site, while the other was located at the Kranji site. The feedwater design parameters for these two plants was similar to the Bedok demonstration plant; the process design parameters are given in Table 17.5. The Kranji plant utilized MF membrane pretreatment, while the Bedok plant utilized UF membrane pretreatment. Both plants selected low fouling composite polyamide membranes for the RO portion of the system. Compared to the Bedok Demonstration Plant, a few changes were implemented, including a reduction to two stages of 7 element vessels, lower recovery and



a. b.

FIG. 17.9 Commercial RO wastewater treatment plants at (a) Bedok and (b) Kranji.

slightly lower flux rates. This was primarily done to reduce the scaling issues seen in the demonstration plant.

Production of water from these two plants was primarily provided to the highly technical, wafer fabrication industry. The water quality has met expectations and the start-up data in Table 17.5 shows that the initial performance of the plants agreed well with projections provided for the project. Also, a comparison of start-up and later data is shown in Fig. 17.10. The main difference between projections and actual start-up data is the salinity of the feedwater. In

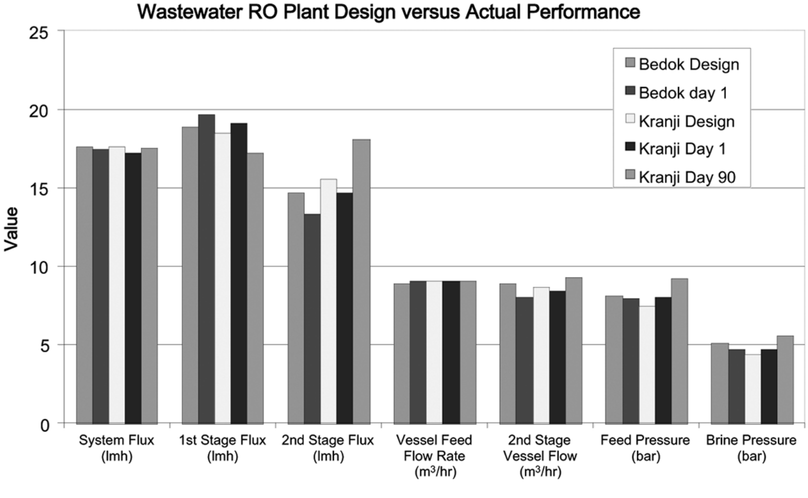


FIG. 17.10 Comparison of design, start-up and continuous operation data from two Singapore plants.

TABLE 17.5

Design and operation parameters for the two Singapore Wastewater Treatment Plants

Parameter (per train)	Units	Bedok design	Bedok day 1	Kranji design	Kranji day 1
System flux	lmh	17.6	17.5	17.6	17.2
1st stage flux	lmh	18.9	19.7	18.5	19.1
2nd stage flux	lmh	14.7	13.3	15.6	14.7
Lead element flux	lmh 1st/2nd stage	22.5, 19.2		21.7, 19.4	
Element array	1st × 2nd stage	50 × 23	50 × 23	49 × 24	49 × 24
Elements/vessel		7	7	7	7
Element area	sq m	37.2	37.2	37.2	37.2
Vessel feed Flow rate	m^3/h 1st stage	8.9	9.1	9.1	9.1
2nd stage Vessel flow	m^3/h 2nd stage	8.9	8.1	8.7	8.4
Vessel reject Flow rate	m^3/h 1st stage	4.1	3.9	4.3	4.1
2nd stage Vessel flow	m^3/h 2nd stage	4.8	4.6	4.6	4.7
Permeate Flow rate	m^3/d	8000	7968	8000	7848
System recovery	%	75	74.7	75	73.3
Feed pressure	bar	8.1	8	7.5	8.01
Brine pressure	bar	5.1	4.75	4.4	4.69
1st stage Permeate pressure	bar	1.5	2	1.4	1.1
Total trains		4	4	5	5
Total permeate flow	m^3/d	32000	31872	40000	39240
Temperature	°C	30	30	30	30
Feed TDS	mg/L	987	572	633	314
Permeate TDS	mg/L	46.6	27.15	41	12.25
Rejection		98.1	98.1	97.4	98.4
pH		7.1	6.7	7.5	5.6
Total chlorine	mg/L		1.8		2.1
TOC	mg/L				15.7

both cases the salinity of the feedwater was lower than expected, and the pressure was at or slightly above projected. The fact that the pressure is slightly higher than projected is quite common for wastewater treatment with RO membranes. This has been reported in other papers (113) and is attributed to the presence of the dissolved organic material.

During the first 3 years of operation for the Bedok and Kranji plants, there were some operational issues, but most of these have been resolved and the plants are operating at full capacity and making water that meets customer requirements. Further expansion in the Singapore region has continued with the 24,000 m^3/d (6.3 mgd) Seletar Wastewater Treatment Plant (2003) and the 145,000 m^3/d (38.3 mgd) Ulu Pandan Wastewater Treatment Plant (expected 2007). With the innovations which have come from the demonstration plant, these plants will be a reliable source of high quality water for years to come.

17.5.2. Southern California wastewater treatment plants

Wastewater reclamation through the use of membrane technology has been commonly practiced for many years in Southern California. Principally, this has primarily been practiced at two key locations, Orange County Water District and West Basin Municipal District.

Orange County Water District (OCWD). The Water Factory 21 treatment facility at OCWD originally began operation in the mid-1970s, utilizing lime clarification pretreatment and cellulose acetate RO membranes. Eventually, the CA membranes were replaced with TFC-HR® polyamide composite membranes to achieve better permeate quality and lower pressures. With the advent of economical MF membrane pretreatment and through extensive pilot testing at their facilities, they converted the lime clarification to MF membrane. The current design of the Groundwater Replenishment System Phase I (GRS Phase I) plant is shown in Table 17.6. The product water has been used to reinject into the aquifer to act as a barrier and prevent intrusion of seawater into the potable water aquifer.

As is typical, the salinity of the water is not particularly high, being around 1000 ppm. A water analysis from the plant is shown in Table 17.7. Additionally, phosphate concentrations range from 0.3 to 3 ppm. It shows relatively high values of ammonia, commonly due to the activated sludge process, and high levels of TOC. As can be seen from the pre and post chemical addition, sulfuric acid is added to lower the pH to prevent scale formation at the high recovery rates used in this application. Also, the plant targets 3–5 mg/L chlorine residual

TABLE 17.6
Design of the OCWD wastewater treatment plant

Location	Fountain Valley, CA
Source	Secondary municipal wastewater
Startup	April 28, 2004
Total capacity (m^3/d)	19,000
Pretreatment filter	Microfiltration
Pretreatment chemicals	Antiscalant, acid
Feed salinity (ppm)	1000
RO trains	3
RO stages	3
Elements/press vessel	6
RO membrane flux (GFD)	20.8
Feed pressure (bar)	9.8
Recovery (%)	89
Perm salinity (ppm)	20
Membrane element	ESPA2

through MF, which then carriers over through RO. For the feed concentration given in Table 17.7, the concentrate salinity is expected to be higher than 10,000 mg/l. The concerns for scaling stem mainly from the possible formation of calcium phosphate, calcium sulfate (saturation index of the concentrate = 117%), silica (saturation index of concentrate = 116%) and calcium carbonate (LSI of the concentrate = 1.7). All of these saturation values are within range of control for commercially available antiscalants. The data in Table 17.7 also point out another typical result for wastewater plants—the RO permeate is better quality than projected by commercial software. This is believed to be due to the high level of dissolved organics which coat the membrane surface and render the membrane to be higher rejection. As explained previously, though, in some cases this can be uncontrolled and lead to continued normalized flux decline. The water quality from the Phase I OCWD plant has been quite consistent over time, as shown in Fig. 17.11. The normalized salt passage has averaged about 1%. This plant is designed at higher recovery and higher flux than the plants in Singapore. Based on the operating data, it appears that these conditions are reasonable for the water that is being treated. The operating data shows that there has been some decrease in normalized permeability at this plant. The increase

TABLE 17.7
Water analysis and projected performance from the RO system at OCWD

Constituent	RO Feed: Pre-chemical addition	RO Feed: Post-chemical addition	RO Permeate: Actual average permeate	RO Permeate: "Projected permeate* T = 25°C"
Ca (mg/l)	74.6	72.8	<0.1	0.6
Mg (mg/l)	30.2	29.1	<0.1	0.2
Na (mg/L)	247	240	6.7	9.1
K (mg/l)	20.2	19.4	0.5	0.9
NH_4 (as mg/l NH_4)	35.5	35.3	1.5	1.7
HCO_3 (as $CaCO_3$)	289	93.7	13.6	9.4
SO_4 (mg/l)	267	459	1.1	5.4
Cl (mg/l)	215	212	4.38	9.9
F (mg/l)	0.84	0.79	0.16	0.07
NO_3 (as mg/l NO_3)	<0.4	2.0	0.44	0.6
SiO_2 (mg/l)	18.3	18.3	<1	0.6
EC (uS/cm)	1810	1890	42.2	76.8
TDS (mg/l)	938	1040	22.3	38.4
TOC (mg/l)	13.5	13.3	0.9	—
Reported pH	8.1	6.7	5.9	5.7

*Projected values were determined with Hydranautics IMS DESIGN v8.8 Software

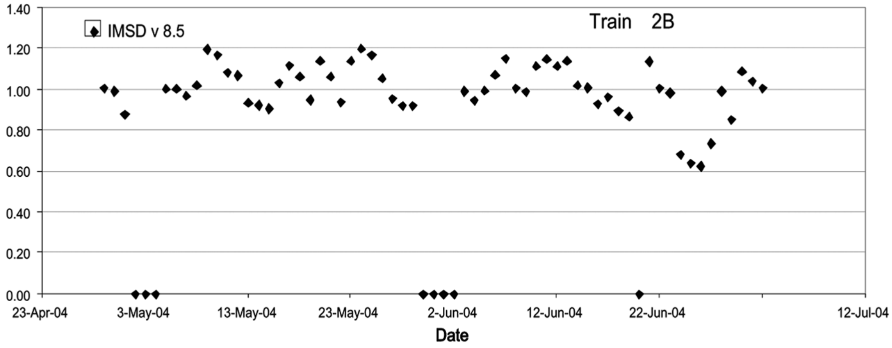


FIG. 17.11 Projected and actual normalized salt passage for polyamide membranes at OCWD (GRS phase I) train 2B.

in pressure has been steady over time and is mostly recovered by chemical cleaning. The pressure drop across each stage has only increased in Stage 1. The pressure drop information by stage is very helpful in diagnosing the cause of the fouling and potential cleaning.

As a result of the extensive testing and demonstration of RO-based reclamation processes they have done at OCWD, the full scale 70 mgd Groundwater Replenishment System will be implemented in 2007. This project will have a similar design, with 14 operational RO trains, 18,900 m^3/d (5 MGD) capacity for each, plus one spare train. The design flux is 20.2 lmh (11.9 gfd) and the recovery is 85%. The plant will use a 78-48-24 array with 7 ESPA2 elements per pressure vessel. The product water from the RO will be further treated by UV with hydrogen peroxide as an advanced oxidation process and reinjected into the aquifer where it can be drawn out (years later) by member water agencies who further treat the water before use. This project will likely be a key example of a safe, efficient way to reclaim water that will be mimicked by other communities.

West Basin Municipal Water District (WBMWD). The West Basin Water Recycling Plant has been operational since 1995, with a number of satellite treatment plants installed in subsequent years. (115) The water production at the site is in excess of 87,000 m^3/d (23 mgd) with 13% used as Title 13, tertiary effluent reuse, 28% for cooling towers, 28% for ground water injection, 23% for low pressure boiler feedwater and 8% for high pressure boiler feedwater. The RO systems at this site are both first pass and second pass systems; the latter used to make very high quality water for local refineries.

This site also started with lime clarification followed by media filtration as pretreatment and cellulose acetate RO membranes, but in the late 1990s it converted to MF pretreatment and polyamide RO membranes. The change of RO membranes from cellulose acetate to composite polyamide has lowered the pressure by more than 150 psi, while the permeate salinity has decreased by more than 50%. Likewise, the permeate TOC levels have dropped by more than 50%. These results again confirm the advantage of the new polyamide membranes for wastewater reclamation.

17.6 Operating cost advantages

Based on the preceding discussion, there are a variety of operating designs that can be considered when treating wastewater by RO. To assess the optimum design, the cost also needs to be considered. An economic model was developed to compare the various operating designs. The assumed values used for this analysis are given in Table 17.7 and were consistent with common literature values (118, 119). Operating and Maintenance (O&M) costs were determined for this case and other potential designs. The resulting cost of this base case is shown in Fig. 17.12.

Other cases were considered, including lower flux with conventional pretreatment, MF/UF pretreatment with 85% recovery with high level of antiscalant dosing, and MF/UF pretreatment with normal antiscalant but low pH. A summary of the alternate process conditions is shown in Table 17.8. Based on these values, the total O&M cost of the RO plant was calculated and is shown in Fig. 17.12. In general, the operating conditions reflect higher operating costs that are associated with wastewater treatment, such as higher cleaning frequencies, chemical costs, lower membrane life and higher downtime.

Case A. This is the reference case and assumes the lowest fouling rate. The values in Table 17.8 selected for this case were based on the actual performance values at Bedok demonstration plant. The cost analysis in Fig. 17.12 shows that this is the lowest cost scenario.

Case B. This is the case for using conventional pretreatment in place of UF/MF pretreatment before the RO membrane system. This flux was decreased for this process condition because of the higher turbidity expected from a media filter or lime clarifier compared to UF/MF pretreatment. Also, fouling rates are still expected to be high due to the high suspended solid that will pass to the RO system. Previous reports have verified that operation of RO membranes on

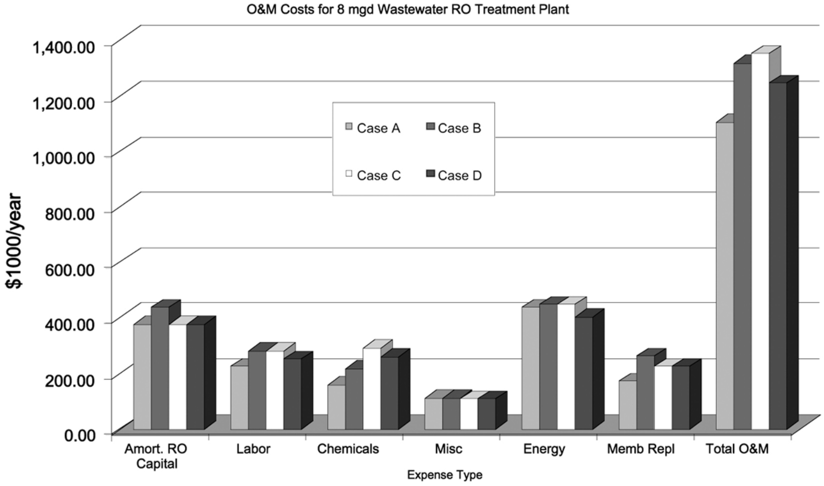


FIG. 17.12 Estimated O&M costs for 8 mgd wastewater RO treatment plant.

TABLE 17.8
RO system cost factors for economic comparison calculations

Pretreatment	Units	Case A UF/MF	Case B Med filter	Case C UF/MF	Case D UF/MF
Recovery	%	75	75	85	85
Flux	lmh	18.0	15.4	18.0	18.0
Pressure	Mpa	0.97	0.99	1.12	1.01
Operators	#	4.0	5.0	5.0	4.5
Acid	ppm	0	0	35.0	68.0
Scale inhibitor	ppm	1.4	1.4	2.8	1.4
Cleanings per year	#	2.0	4.0	4.0	3.0
Cleaning cost	$/L	0.08	0.15	0.15	0.11
Memb life	years	6.5	5	5	5
Element cost	$US	600	600	600	600
Electricity cost	$/kwhr	0.06	0.06	0.06	0.06
Amortization time	years	20	20	20	20
Amortization rate	%	6	6	6	6
Array	1st,2nd,3rd	44,22,0	52,25	39,19,8	39,19,8
Permeate TDS	mg/L	53	62	82	85

conventionally treated wastewater does result in higher fouling rates and more frequent cleaning. In one report (117), the different RO membrane was found to require cleaning every month. This higher fouling rate impacts many parameters, and must be partially compensated by running at lower flux rates. The net result is higher capital cost, higher chemicals, and higher replacement costs. The RO costs for this case were about 20% higher than the base case. If the pretreatment process and the disposal of sludge from the pretreatment process is less than 20% higher than the UF/MF pretreatment, this case would be potentially attractive.

Case C. This case assumes that the recovery of the RO plant is increased from 75 to 85%. As shown in Table 17.8, the high calcium phosphate scaling tendency means that additional acid and antiscalant are needed. The pH of this water was adjusted to 6 and antiscalant expense was doubled. This is seen as the most expensive case. This has the highest chemical cost of all the cases considered. This is a result of the large expense associated with lowering the pH and increasing antiscalant.

Case D. This is also a high recovery process, however, the scaling is primarily controlled by lowering the pH to 5. At this value the calcium phosphate scaling tendency is very low and easily controlled by minimal antiscalant cost. The O&M of this case totals to be about 15% higher than the base case. Even with the 10% savings expected for the reduced pretreatment costs, this Case still will not be as economically competitive as Case A. This process has a slightly lower energy consumption than the other cases due to the lower amount of feed water being pumped.

17.7 Other wastewater treatment applications

As stated in the beginning of the chapter, membranes are also being extensively used to treat industrial wastewaters as well as municipal wastewaters. Although these systems are generally smaller, they have the added benefit of lowering the waste disposal costs of the factory as well as reducing their need for purchasing process water. Examples of such treatment applications includes, textile plant waste, tannery wastewater, metal working fluids, refinery wastewaters, oil production facilities, and many more. As an example, we will describe the treatment of dye waste from the textile industry (122).

The textile industry is a large world wide industry that utilizes a water intensive process and generates a large amount of difficult wastewater. The treatment of the 5000 tons/year of cotton material can require 650,000 m^3 (170 mg) of water and generate 559,000 m^3 (148 mg) of wastewater (121). The wastewater comes from a number of different places in the dyeing process, as illustrated in Fig. 17.13 and can have a combined COD of 600–800 mg/l O_2 and suspended solids of 100-200 mg/l. The individual wastewater streams range from low turbidity and high color to low color and high turbidity. Many of the streams from post-dyeing continuous wash operations and periodic dye bath discharge contain sodium chloride concentrations of 10 g/l or greater, while some are a few g/l. For years these wastewaters have been treated by conventional coagulation, biological oxidation and media filtration processes. However, these processes generate additional solid waste in the form of inorganic sludge from clarifiers and organic sludge from the biological processes, and do not make water of sufficient quality for reuse. Therefore, engineers have been looking at membrane processes as a way to enhance the treatment of the wastewaters. Laboratory studies have shown that Koch TFC-SR NF membranes have potential for treating synthetic dye solutions (121), with the salt permeating

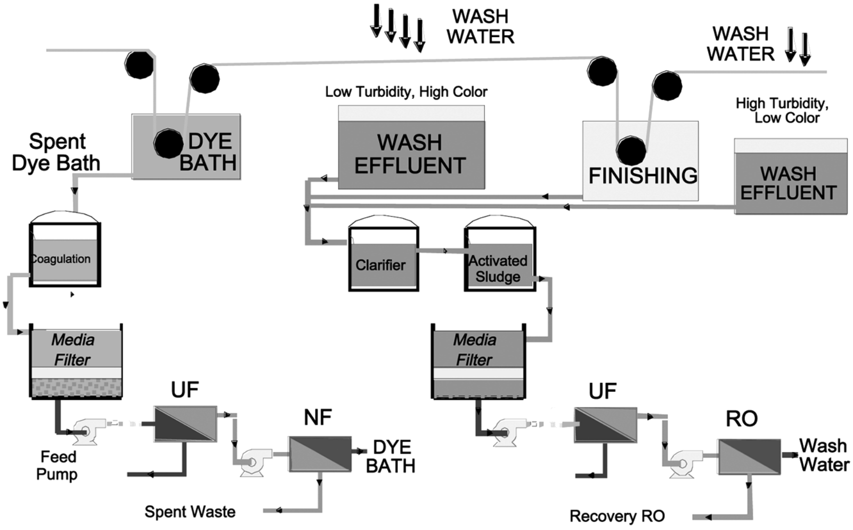


FIG. 17.13 Textile wastewater treatment process.

through the membrane while the organic can be removed and concentrated.

Recently, small commercial plants have successfully implemented integrated membrane processes to recover water and reduce wastewater generation. As shown in the Fig. 17.12, these membrane processes have been used to treat the low volume (75 m^3/d, 20,000 gpd) spent dye bath waste and the high volume (700 m^3/d, 0.2 mgd) wash water from dyeing and finishing operations. In both cases HYDRAcap UF modules (120,000 MWCO) were used to treat the wastewater from a conventional pretreatment process. The UF effluent was then treated by HYDRACoRe NF membranes or LFC1 RO membranes. For these wastewaters, the NF and RO were run at relatively low fluxes of 16.9–20.3 lmh (10–12 gfd). Both processes have been able to operate at stably at these conditions.

Examples of water from the various treatment steps are shown in Fig. 17.14. The UF system operates at a flux of 41-44 lmh (24–26 gfd) and only requires acid cleaning every 90 days. The feed to the NF membrane can have 10,000 to 70,000 mg/l of TDS. The HYDRACoRe NF membrane was chosen because it was able to remove the dye chemical while passing most of the NaCl. The NF membrane ran at 75% recovery, with the permeate going back to the dye bath, and the concentrate going to spent waste.

Meanwhile, the high volume waste from the wash water has much lower salinity, 3000–6000 mg/l TDS, because of the dilution of the wash water, but it

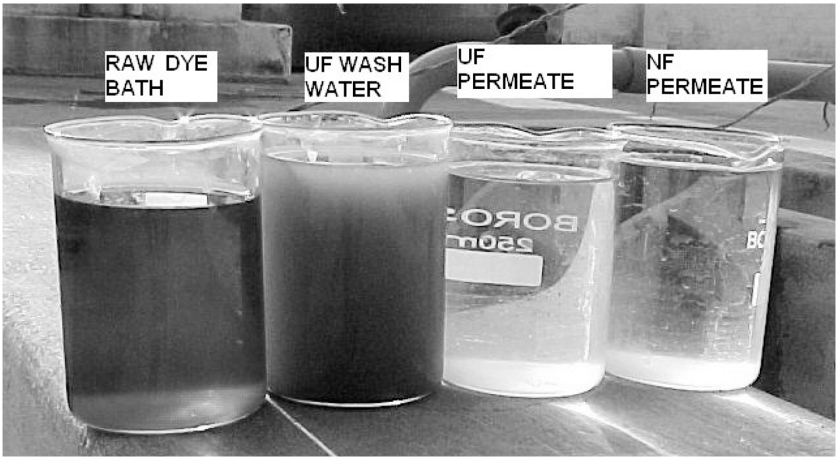


FIG. 17.14 Water samples from various treatment steps of spent dye wastewater.

contains high levels of turbidity from the textile material. This stream was treated with HYDRAcap UF membrane at 54–61 lmh after the conventional clarification, biological oxidation, and multimedia filtration. The UF filtrate which had virtually no suspended solids, was treated with the LFC1 RO membrane. This membrane required cleaning every 1 to 2 months. To further reduce the waste stream, the LFC1 concentrate was further concentrated with SWC3+ seawater membranes, due to the high salinity of the water. The SWC3+ was able to operate at 50% recovery, with overall recovery of 80–85%. RO permeate was reused as wash water, while the SWC3+ concentrate was sent to an evaporator. Thus, there was zero liquid discharge from this part of the process.

These systems have operated for more than a year and demonstrated stable performance. They have demonstrated cost savings as expected and show promise for use in large-scale applications.

17.8 Conclusions

The development of polyamide composite membranes and improved RO process design has made RO technology very effective for the treatment of municipal wastewater. The use of UF/MF pretreatment has proven to effectively remove colloidal foulant, and chloramines are effective to control biofouling in most cases. It is important to consider the water chemistry and scale formation potential when selecting recovery rates, or the plant will struggle to meet water production and chemical cleaning costs will be high. While some conventional polyamide membranes have been shown to have stable permeability, many plants are electing to use hydrophilic, neutrally charged RO membranes to ensure low fouling rates. Further experience is needed to determine which membrane will be best for a given wastewater reclamation project. One key factor which may be overlooked when designing a RO wastewater process is the potential for calcium phosphate scaling. Most wastewaters contain moderate to high levels of phosphate and calcium. When operating at high recovery, it is easy to greatly exceed the calcium phosphate solubility. There are many computational models present which can predict the occurrence of scaling. These indicate that in many cases the phosphate must be reduced by dephosphization, acidification or stabilized with improved calcium phosphate antiscalants. These factors and results point to the continued favorable use of this technology for reclaiming wastewater.

In the future, however, it is anticipated that MF/UF-RO treatment of secondary effluent will be replaced by MBR-RO treatment of primary effluent. This has already been proven effective (114), and will certainly save money, space and reduce waste compared to the current MF/UF-RO process.

17.9 Acknowledgements

The author would like to recognize the assistance received from a number of individuals in gathering this information. In particular, Mr. Rich Franks who has analyzed and prepared charts for much of the plant data reported here as well as Mr. Satish Chilekar, who provided much of the detailed information on the use of membranes in the dye waste treatment application.

18

Budgeting and implementation of desalination projects

*Nikolay Voutchkov**

The purpose of this chapter is to provide guidelines for estimation of costs associated with development, construction, and operation and maintenance (O&M) of seawater desalination reverse osmosis (SWRO) plants, and to discuss key methods of project implementation.

18.1. Overview of desalination cost estimating procedures and practices

18.1.1. Project cost definitions

The key economic parameters of a SWRO desalination project are:

- Capital costs
- Operation and maintenance (O&M) costs
- Cost of water

18.1.2. Definition of capital costs

Capital costs include all expenditures associated with the implementation of a given desalination project from the time of its inception, through design,

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permitting, financing, construction, commissioning and acceptance testing for normal operation. The largest component of the capital costs are the expenditures for project construction (construction costs). Construction costs encompass direct expenditures needed to build all project structures, and procure and install equipment, piping and service utilities. Because of their direct association with the construction of physical facilities, the construction costs are also referred to as “direct” or “hard” capital costs. Construction costs are typically 50–85% of the total capital costs.

The remaining portion (15–50%) of the capital costs are often referenced as “indirect” of “soft” costs. These costs are associated with all engineering, administrative and financing efforts necessary to bring the project to fruition, and to secure permits, funds and contractors needed to build and operate the plant.

The total project capital costs are typically presented in monetary units (i.e., US$) estimated either for the year of the beginning of project construction or expressed as costs referenced to the middle of the construction period. Depending on the type, length and term of the project financing, project capital costs are often converted in monetary units per year and referred to as amortized or annualized costs (US$/yr). In addition, both capital and construction costs are sometimes presented as costs per unit of desalination plant production capacity (i.e., US$/m^3/d, US$/gpd).

18.1.3. Definition of operation and maintenance costs

Operation and maintenance costs are all costs associated with: SWRO plant operations (power, chemicals, labor, and replacement of consumables); with maintenance of plant equipment, buildings and utilities; and with compliance with all plant operation and environmental permits and regulatory requirements. The operation and maintenance costs are typically expressed as the all-inclusive operational expenditures for a period of one year (i.e., US$/yr) or as operational costs for the production of unit volume of desalinated water (i.e., US$/m^3, US$/kgal).

Operation and maintenance costs may be divided in two main categories: fixed and variable. Fixed O&M costs are these that are incurred independent of the actual amount of fresh water produced by the desalination plant. Fixed costs include: labor costs (staff wages and fringe benefits); costs for equipment maintenance; environmental and performance monitoring; operational insurance; administrative costs; and other miscellaneous expenses. Variable costs are typically proportional to the volume of desalinated water produced by the SWRO plant and include expenditures for: power; chemicals; replacement of

RO membranes and cartridge filters; and for waste disposal. Typically variable costs are 50–85% of the total annual O&M costs, while the fixed costs are 15–50% of these expenditures.

18.1.4. Definition of cost of water

Cost of water is an economic parameter that incorporates all project capital and annual O&M expenditures associated with water production, and is typically presented as monetary units per unit volume of desalinated water (i.e., US\$/m^3, US\$/kgal). The cost of water is calculated by dividing the sum of the amortized (annualized) capital costs (i.e., US\$/y) and the annual O&M costs (i.e., US\$/y) by the average annual desalination plant fresh water production volume (m^3/y, thousand gal/y). For a typical SWRO plant, the amortized capital costs and the O&M costs are usually in a range of 40–60% of the total cost of water.

Figure 18.1 shows an example of cost of water breakdown for a large seawater desalination project. Although, the ratio between the key cost components varies from project to project, the "largest pieces of the cost pie" are usually the plant construction expenditures (i.e., the direct capital costs), power, and the other O&M costs. The indirect capital costs, which mainly include expenditures for project engineering, development and finance, are also a significant portion (typically 10–20%) of the water production costs.

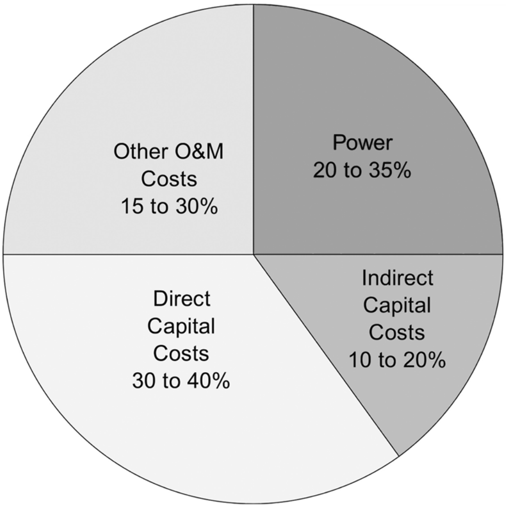

FIG. 18.1 Cost of water breakdown.

18.2. Key factors influencing project costs

Project capital, O&M, and overall water production costs depend on a number of factors, most of which are site specific to the particular project location, and technical and socio-economic circumstances. In general, there are two types of factors that strongly influence the project costs: (1) factors controlled by the decisions of the facility owner; and (2) subjective factors beyond control of the facility owner, including those that result from regulatory requirements and market forces of free economy.

18.2.1. Cost factors within the control of the plant owner

The key cost factors which are within the control of the project owner are discussed below.

Project size. Project size has a significant effect on the overall production cost of desalinated water. As illustrated on Fig. 18.2, the cost of water production by seawater desalination can be reduced significantly by building a fewer large-scale desalination plants rather than a large number of small facilities. For example, analysis of Fig. 18.2 indicates that the water production costs can be reduced by over 50% when plant capacity is increased from 4,000 m^3/d to 20,000 m^3/d (1.1–5.3 mgd). This economy of scale is mainly driven by the size of the individual treatment and pumping units, especially the RO trains. Cur-

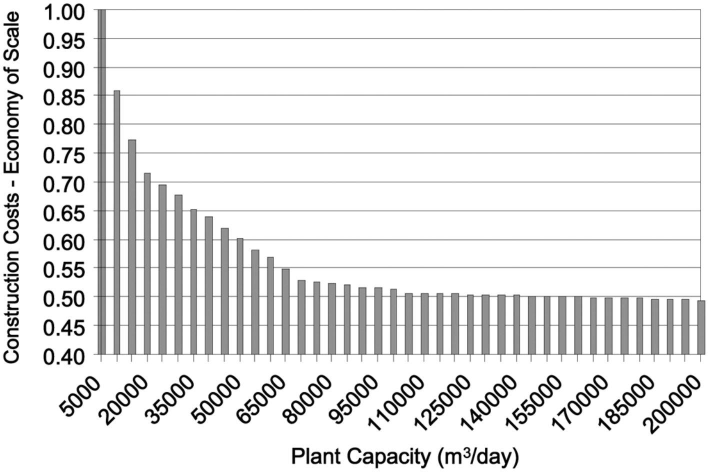

FIG. 18.2 Effect of plant size on water production costs.

rently, the largest size SWRO train that can be built using off-the shelf standard equipment (high-pressure pumps, Pelton energy recovery turbines and 8-inch membranes) has production capacity of 21,000 m^3/d (5.5 mgd). Construction of larger individual trains is possible, but usually is not as cost effective because it would require the use of custom-made RO system equipment, which is significantly more costly than the off-the-shelf standard equipment units and, as a result, some of the economy-of-scale savings are negated by the additional equipment costs.

The economy-of-scale benefits are minimal for plants larger than 200,000 m^3/d (53 mgd), mainly because the added complexity of flow distribution, treatment and operations. As the maximum unit size of commercially available desalination plant equipment (pumps, membranes, pressure vessels, energy recovery systems, etc.) increases in the future it is likely that the breakpoint plant capacity at which economy of scale would not yield measurable savings would shift to 400,000 m^3/d (106 mgd) or higher.

Capacity availability factor. The capacity availability factor is defined as the percent of the time per year during which the seawater desalination plant is producing desalinated water flow equal to or higher than the plant average annual design flow. For example, if the desalination plant design capacity is 37,850 m^3/d (10 mgd) and the plant capacity availability factor is 100%, than the total average annual fresh water flow the plant can produce is 37,850 m^3/d × 365 d/y = 13,815,250 m^3/y (3,650 million gal/y). If, for example, the actual plant flow produced by the same plant over the period of one year is only 12,450,000 m^3/yr (3,290 million gal/y), than the plant capacity availability factor is 12,450,000 m^3/y / 13,815,250 m^3/y = 90%.

When sizing reverse osmosis (RO) systems, it is commonly practiced to select the size of the individual RO trains to be 10–20% of the overall plant capacity in order to provide operational flexibility. This design approach is based on the fact that under a condition of the routine membrane cleaning of one RO train, the production capacity of the other RO trains can be increased with 10–20% for a short period of time, in order to compensate for the production of the train taken out of service for cleaning. This design approach typically allows to achieve an average annual production capacity availability of 90%, because often in years of unusual water quality (prolonged rains, red tide events or dredging of the intake area, etc.), frequently two or more RO trains may need to be taken out of service at the same time to maintain effective operations. Since often the SWRO system is not designed to accommodate more than one RO train out of service, the capacity availability of this system is limited.

Usually, designing plants of capacity availability factor of 90% is acceptable for seawater desalination plants that provide only a limited portion of the flow for a given fresh water user (municipality, community, or industry). However, when a desalination plant is projected to supply the majority or all of the fresh water used by a given water consumer, than this plant has to be designed for a capacity availability factor of 95% or higher. Higher capacity availability is typically achieved by the construction of one or more standby RO trains; and or/reducing the size and increasing the number of the individual RO trains. These design measures improve system reliability and redundancy, but at the same time increase the project construction and O&M costs.

At present, many of the existing large municipal SWRO plants worldwide are designed to supplement existing conventional water supply sources rather than to be the primary or the only source of water supply for a given area. Therefore, the operation of these plants does not need to have the flexibility to follow the actual diurnal and monthly fluctuations in product water demand and most of the plants are designed to operate at constant production capacity and availability factor of 90–95%.

As the costs of seawater desalination continue to fall in the future, SWRO plants are likely to become a prime rather than a supplemental source of water supply for many coastal communities with limited traditional local sources of fresh water supply (i.e., groundwater, river or lake water). The SWRO plants servicing such areas have to be designed to have the operational flexibility to match desalination plant production with the potable water demand patterns of the water users and to have capacity availability factor of 96% or higher.

Shift of the SWRO plant operational paradigm from constant to variable production flow and increased plant availability factor requires a change of the typical SWRO configuration from one that is most suitable for constant production output to one that is most cost-effective for delivery of varying permeate flow. A response to such shift of the desalination plant operational paradigm is the three-center RO system configuration implemented for the first time for the Ashkelon seawater desalination plant in Israel (see Fig. 18.3). Under the three-center configuration, the RO membrane vessels, the high-pressure pumps and the energy recovery equipment are no longer separated in individual RO trains, but are rather combined in three functional centers—a high-pressure RO feed pumping center, a membrane center and an energy recovery center (126). The three functional centers are interconnected via service piping.

The three-center configuration uses relatively smaller trains (RO trains with 50 pressure vessels per train vs. conventional design of 100 to 200 vessels per RO train). The RO feed pumping center includes only a few large-capacity

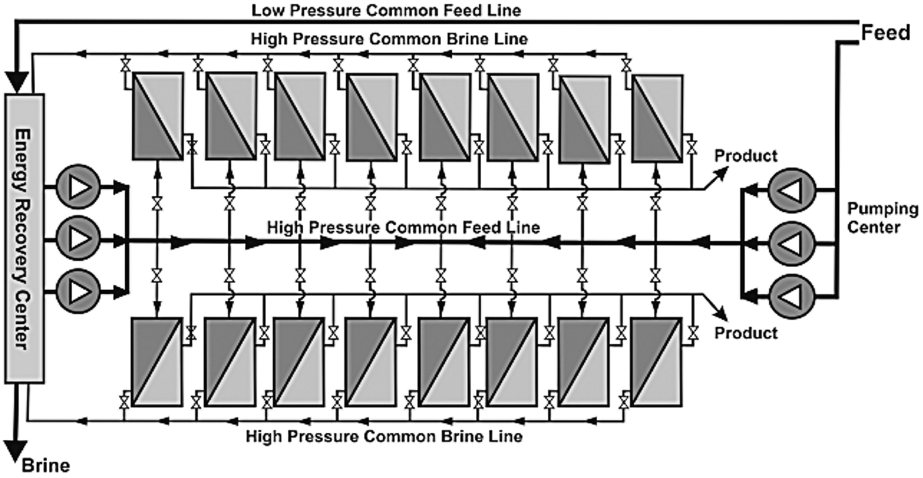

FIG. 18.3 Three-center SWRO system.

high-pressure pumps that deliver seawater to the RO membrane center. The main benefit of using few large-capacity high-pressure pumps rather than a large number of small capacity units is the gain in overall pumping efficiency. Typically, the smaller the ratio between the pressure and the flow delivered by a given pump, the better the pump efficiency and the "flatter" the pump curve (i.e., the pump efficiency is less dependent of the variation of the delivered flow). Therefore, pump efficiency can be improved by either reducing the pressure delivered by the pump or by increasing pump flow. The pump operating pressure reduction is limited by the RO system target salt separation performance and the associated osmotic pressure. Therefore, an alternative approach to improve pump efficiency is to increase unit pump flow (i.e., use larger pumps). While a conventional size high pressure RO feed pump of small capacity would typically have a maximum total energy use efficiency of 80–85%, the use of ten times larger size pump may allow to increase the pump efficiency to 88–90%, especially for large SWRO plants. This beneficial feature of the three-center design is very valuable in the case of systems delivering varying flow.

While in a conventional RO train design the membrane vessels are typically grouped in 100–200 units per train and in 2–20 RO trains, the membrane center configuration contains two to four times larger number of RO vessel groups (banks) and a smaller number of membrane vessels per bank. Under this configuration the individual vessel banks are directly connected to the high-pressure pump feed lines and can be taken off service one at a time for membrane replacement and cleaning. Although the feed water distribution piping for such membrane center configuration is more elaborate and costly than the use of individual RO trains, what is lost in capital expenditure is gained in overall system performance reliability and availability. A reliability analysis completed for a 95,000 m^3/d (25 mgd) SWRO plant (126) indicates that the optimum number of vessels per bank for this scenario is 54 and number of RO banks per plant is 20. A typical RO-train based configuration would include two to four times more (108–216) vessels per RO train and two to four times less (5–10) RO trains. According to this analysis, the use of the three center configuration instead of the conventional RO-train based approach allows to increase RO system availability from 92–98%, which is a significant benefit in terms of water supply reliability.

The centralized energy recovery system included in the three-center configuration (Fig. 18.3) uses high-efficiency pressure-exchanger based energy recovery technology. This configuration allows to improve the overall energy efficiency of the RO system and to reduce system power, equipment and construction costs. Because of the high efficiency of the pressure exchangers, the energy penalty for operation at lower recovery is small. This allows operating the SWRO plant cost-effectively at a wide range of plant recovery while delivering variable product water flow. For example, if the SWRO plant output has to be reduced by 30% to accommodate low diurnal demand, a SWRO system with RO train-based configuration has to shut down 30% of its trains and if this low demand persists, it has to flush these trains in order to prepare them for the next start up. Frequent RO train starts and stops result in increased membrane cleaning costs, in shorter membrane useful life and in higher labor expenditures. A RO system with three-center configuration would only need to lower its overall recovery in order to achieve the same reduction in diurnal water production. Although temporary operation at lower recovery would result in elevated costs for pumping and pretreatment of larger volumes of source water, these extra operational expenses are typically compensated by the lower osmotic pressure needed to operate the SWRO system at lower recovery and by the high energy recovery efficiency of the pressure-exchanger energy recovery system.

As indicated above, designing the SWRO plant around a higher capacity availability factor (90% vs. 98%) results in increase in the plant construction costs. Typically, the incremental cost of water increase to improve capacity availability from 90–95% is in a range of 3–5%. Increasing plant capacity availability factor from 95–98% usually is more costly, and would result in an additional 5–10% increase in water production costs. However, in many cases, the incremental costs associated with improved reliability can be compensated by the increased plant production capacity and reliability.

For example, let's assume that the cost of production of 1 m^3 of fresh water by a 37,850 m^3/d (10 mgd) plant with capacity availability factor of 90% is US$0.60/$m^3$ (US$2.27/kgal). The annual water sales revenue that this plant would generate is: 90% × US$0.60/$m^3$ × 37,850 m^3/d × 365 d = US$7,460,235/y. Let's now assume that this plant design is modified to improve the plant capacity availability factor from 90% to 95%, at a cost of water production increment of 3%, i.e., the new cost of water is 1.03% × US$0.60/$m^3$ = US$0.62/$m^3$ (US$2.35/kgal). The additional cost of water production that the utility will incur is = US$0.02/$m^3$ (i.e., 95% × US$0.02/$m^3$ × 37,850 m^3/d× 365 days = US$262,490/y). However, the additional revenue the water utility will gain, even if the utility sells the water at the same price (i.e., US$0.60/$m^3$), will be: 5% × US$0.60/m^3 × 37,850 m^3/d× 365 d = US$690,760/y, i.e., the utility will improve their annual revenue with US$690,760/y – US$262,490/y = US$428,270/y (i.e., 5.7% increment over the baseline revenue of US$7,460,235/y). This example, although hypothetical, illustrates the fact that in most cases in a long run it pays to build facilities of higher capacity availability factor than to construct lower cost/lower reliability plants, especially when there is high and continuous demand for desalinated water in the plant service area.

Source water quality. The key source water quality parameters that impact desalination system design, operations, and cost of water production are: TDS, temperature, turbidity, SDI, organic content, nutrients, algae, bacteria, temperature, boron, silica, barium, calcium and magnesium.

Of these parameters, seawater TDS and temperature are the two key source water quality parameters that have the most significant influence on the cost of seawater desalination. Table 18.1 presents typical TDS concentration and temperature for various seawater sources.

In general, desalination construction and O&M costs increase with the increase of source water TDS and decrease with the increase of temperature. The source water TDS concentration is directly related to the SWRO system design feed pressure, overall plant recovery and configuration. Therefore, the use of lower salinity source water (such as bay water or a mix of ocean and brackish water rather than open-ocean water) allows to reduce the costs associated with construction and operation of the RO system and at the same time to increase plant recovery.

It is important to note that the consistency of the source water TDS concentration is almost as equally important for a successful low cost SWRO design, as the level of TDS of the source water. In addition, usually fresh surface water

TABLE 18.1
Salinity and temperature of various seawater sources

Seawater source	Total dissolved solids concentration (mg/L)	Temperature (°C, °F)
Pacific/Atlantic Ocean	33,500	9–26, 48–79 (avg. 18, 64)
Caribbean	36,000	16–35, 61–95 (avg. 26, 79)
Mediterranean	38,000	18–26, 64–79 (avg. 24, 75)
Gulf of Oman, Indian Ocean	40,000	22–35, 72–95 (avg. 30, 86)
Red Sea	41,000	24–33, 75–91 (avg. 28, 82)
Arabian Gulf	45,000	18–35, 64–95 (avg. 26, 79)

Note: Seawater TDS and temperature may be outside the table ranges for a site-specific location.

TABLE 18.2
Effect of source water on desalination costs

Water source	Construction costs	O&M costs	Cost of water
Pacific/Atlantic Ocean	1.00	1.00	1.00
Caribbean	1.04–1.35	1.02–1.10	1.03–1.22
Mediterranean	1.06–1.40	1.04–1.15	1.05–1.28
Gulf of Oman, Indian Ocean	1.15–1.50	1.10–1.25	1.12–1.38
Red Sea	1.18–1.55	1.12–1.28	1.15–1.42
Arabian Gulf	1.25–1.60	1.15–1.33	1.20–1.48

sources such as river and lake water may carry an order of magnitude higher levels of turbidity, organics, nutrients, and other man-made pollutants which removal by pretreatment or post-treatment may cost more than the savings associated with lower TDS concentration. Table 18.2 presents comparison between the cost of construction, operation and water production of Pacific/Atlantic ocean water and other water sources indicated in Table 18.1.

The cost multiplier ranges in Table 18.2 account mainly for differences in source water TDS concentration and temperature, and are normalized for all other key factors that influence costs such as: product water quality; cost of capital; power cost; concentrate disposal costs; membrane useful life and costs, etc. The multipliers in Table 18.2 apply for large seawater desalination projects of capacity between 40,000 and 200,000 m^3/d (11–53 mgd). The actual costs of individual projects may vary because of site-specific project differences and conditions. A detailed analysis of the effect of source water quality on the costs for sweater desalination is provided elsewhere (127).

Plant source water temperature has a measurable effect of the SWRO design feed pressure and membrane performance. The required SWRO feed pressure typically is reduced by 5–8% on a linear scale for every 10°C (18°F) source water temperature increment in the temperature range of 12–40°C (62–104°F) . Based on tests completed at the Carlsbad seawater desalination pilot plant on cold Pacific Ocean water in the winter, when the source water temperature drops below 12°C (62°F), the temperature effect is even more dramatic–the SWRO feed pressure increases with 5–10% for every 2°C (3.6°F) of temperature drop on an exponential scale until the source water temperature reaches 4°C (39°F), below which the source water would begin to freeze and seawater desalination is dramatically hindered. The accelerated exponential increase in the operational SWRO feed pressure for source water temperatures below 12°C (62°F) is explained by similar curvilinear increase in source water density in the temperature range of 4–12°C (39 to 62°F) combined with changes in membrane material behavior. Source seawater of temperature above 40°C (104°F) typically has two negative effects on membrane performance that may negate the positive effect on membrane pressure: (1) change in membrane material behavior (membrane compaction), which could result in shorter membrane useful life; and (2) accelerated membrane biofouling due to the effect of the temperature on bacterial growth. An additional negative effect of temperature on membrane performance is the reduction in membrane salt rejection with temperature. Therefore, operation at high source water temperatures (typically, 30°C/86°F and higher) may compromise meeting product water quality goals in terms of TDS, chlorides, boron, sodium and other product water quality requirements and may require the installation of additional treatment step: partial or full second pass: to address the negative effect of temperature on product water quality. The SWRO system construction cost increase associated with the installation of partial or full second RO pass is typically in a range of 10–25% of the cost of the first-pass SWRO system. The additional O&M costs associated with the operation of second pass-system vary between 3 and 10% of the costs for operation of the first pass.

Product water quality. Product water quality has a measurable effect on SWRO plant configuration, design and costs. Typically, the higher the required product water quality the higher the desalinated water costs. Potable use of desalinated seawater is closely related to the levels of TDS, chlorides, boron and bromides in this water. Drinking water regulations worldwide usually establish levels of TDS and chlorides in the product water below 500 mg/L and 250 mg/L, respectively. However, when using desalinated seawater, the importance of these parameters is often overshadowed by the health and irrigation related water quality requirements in terms of boron and disinfection driven water quality targets in terms of bromides. The main reason why boron and bromides are of specific importance for the overall quality of the desalinated water is the fact that their concentration in seawater is usually an order of magnitude higher than that of typical freshwater sources (rivers, lakes, groundwater, etc.). For example, typical river water has boron concentration of 0.05–0.2 mg/L, while the seawater boron levels are usually between 4.0 and 6.0 mg/L. Similarly, bromide level in fresh water sources is usually between 0.05 and 0.3 mg/L, while seawater has bromide concentration of 55–85 mg/L. While SWRO membranes remove over 70% of the boron and over 99% of the bromides in the source seawater, the remaining levels of these compounds are still several times higher than these in fresh surface water sources.

Boron level in the desalinated water is often required to be reduced to less than 1 mg/l to achieve public health goals and to less than 0.75 mg/L (sometimes below 0.5 mg/L) in order to alleviate problems associated with the use of this water for irrigation of sensitive food crops (i.e., citrus trees, avocados, strawberries, etc.) or ornamental plants. In order to achieve these water quality goals, often product water TDS and chloride levels have to be reduced below 100 mg/L and 50 mg/L, respectively.

The bromide concentration of the desalinated seawater may also have a significant effect on the required level of removal of salts from the seawater, especially if this water is disinfected using chloramines rather than chlorine, or if it is ozonated. While using chlorine only creates stable chlorine residual that shows minimum decay over time, applying combination of chlorine and ammonia to desalinated water with bromide levels above 0.4 mg/L usually yields unstable total chlorine residual that decays rapidly (within several hours) to unacceptably low levels. Although effect of high level of bromide in the desalinated water can be mitigated by supper-chlorination (i.e., applying initial chlorine at dosages of 4.0 mg/L or higher), this chloramines decay effect has to be accounted for in the chemical costs for seawater desalination. Ozonation of desalinated water with bromide concentration above 0.4 mg/L may result in formation of unacceptably high concentration of bromate in the water. In addition to the potable uses discussed above, the desalinated water quality requirements for TDS, boron, chlorides, sodium, silica and other contaminants may be driven to even higher levels for some industrial applications, especially these where ultrapure water quality is necessary.

Producing higher quality desalinated water is associated with a measurable cost increment. Table 18.3 provides information on the relationship between the target product water quality and the associated costs for plant construction and operation, and for overall water production. The costs to produce water of TDS of 500 mg/L, chloride level of 250 mg/L, boron of 1.0 mg/L and bromides of 0.8 mg/L are used as a base for comparison and are assigned a value of 1. Cost increment to achieve more stringent water quality goals is presented using multiplier values.

The cost range presented in Table 18.3 reflects the fact that the cost increment would vary not only with the product water quality targets but also with the source water quality. The cost multiplier is also affected by the selected technology or combination of technologies to meet the product water quality target and the costs of various consumables (i.e., chemicals, power, ion-exchange material, etc.) used for product water quality polishing.

Concentrate disposal method. Depending on the site-specific conditions of a given project concentrate disposal expenditures may have a measurable con-

TABLE 18.3
Effect of target product water quality on water costs

Target product water quality	Construction costs	O&M costs	Cost of water
TDS = 500 mg/L chloride = 250 mg/L boron = 1 mg/L bromide = 0.8 mg/L	1.00	1.00	1.00
TDS = 250 mg/L chloride = 100 mg/L boron = 0.75 mg/L bromide = 0.5 mg/L	1.15–1.25	1.05–1.10	1.10–1.18
TDS = 100 mg/L chloride = 50 mg/L boron = 0.5 mg/L bromide = 0.2 mg/L	1.27–1.38	1.18–1.25	1.23–1.32
TDS = 30 mg/L chloride = 10 mg/L boron = 0.3 mg/L bromide = 0.1 mg/L	1.40–1.55	1.32–1.45	1.36–1.50

tribution to the total plant construction and O&M costs, and to the overall cost of water. Use of existing outfalls for concentrate disposal and more specifically co-disposal with power plant cooling water usually yields lowest concentrate disposal costs. For small seawater desalination plants with low-cost access to existing wastewater collection system, concentrate disposal may be very cost-attractive as well. On the other hand, construction of new long discharge outfalls or series of deep groundwater injection wells, although widely practiced for small desalination plants, is often costly and site-prohibiting for large projects. Section 18.5.1 of this chapter provides a more detailed discussion of concentrate disposal methods and associated costs.

Power supply and unit power cost. As indicated on Fig. 18.1, the cost of power is typically 20–35% of the total cost of production of desalinated water. Therefore, both the unit power cost and the power demand of the desalination plant have a profound effect on the project costs.

18.2.2. Project risk profile

As indicated previously indirect capital costs, including cost of project financing, development and permitting, are a significant portion (10–20%) of the overall water production costs. These costs are closely related to the risks associated with project permitting, construction and operation. Financial institutions establish the interest rate of the funds they lend to the project and the acceptable project financial structure based on thorough evaluation of the project risk profile. In order to provide low-interest rate funding for a given desalination project, financial institutions demand strong assurances that the project will be permitted and built in a timely and cost-effective manner; the power supply contract and tariff for the project will be reasonable; the operation and maintenance of the desalination plant will be professionally handled by an operations staff that has successful prior experience in seawater desalination; and that the regulatory and permitting risks of the project are minimal. In the case of build-own-operate-and-transfer (BOOT) projects, financial institutions lending funds to the project would also look for a fair and balanced water purchase agreement between the final user of the desalinated water and the BOOT contractor, which apportions risks equitably between the two parties.

The profile of the group or company providing funding for a given project may vary from project to project and could be a combination of private sector commercial lenders, banks and multilateral agencies, and international financial institutions. Increasingly, funding for desalination projects is provided from the capital markets and project bonds. Lenders differ in their approach to project risk–public sector bond underwriters/lenders and private sector (commercial) lenders often have different approaches and requirements in evaluating and mitigating project risks.

As a general rule, project fund lenders would only be willing and able to take risks that are quantifiable. Typically, lenders are not involved in the construction, operation, or insurance activities related to project implementation. Therefore, they would not take risks associated with these activities and especially risks they are not familiar with or that can be more appropriately borne by other parties involved in the project. In order to mitigate risks at early stages, the project lenders may require to be involved in the key milestones of project development and implementation, including the negotiation of project contracts, review of key project design and construction activities, as well as review and approval of certification of project completion, and project acceptance testing. Lenders would generally exercise their review rights over the project implementation with the assistance of an independent engineer.

Key project risks considered by the lending institutions when determining their interest and conditions (i.e., the cost of money) for funding desalination projects are:

- Permitting risks
- Entitlement risks
- Power supply risks
- Construction risks
- Source water risks
- Technology risks
- Regulatory risks
- Operational risks
- Desalinated water demand risks
- Financial risks

Permitting risks. Permitting risks are risks associated with obtaining and maintaining of all permits required for project implementation and for long-term plant operation including: environmental permits (such as the concentrate discharge permit); construction permits; and operations permits. Because seawater desalination projects are relatively new to many permitting agencies, the time and efforts required for permitting of this type of projects are usually more extensive than these for conventional water and wastewater treatment plants. Often the permitting of large seawater desalination projects requires long and costly environmental and engineering studies, and is influenced by environmental opposition. As a result, permitting risk is considered by lending institutions and public agencies alike as one of the main and costliest risks associated with desalination project implementation. For example, difficulties encountered with permitting of the Tampa Bay seawater desalination project was one of the key reasons why the public utility that initiated this project (Tampa Bay Water) decided to proceed with project implementation under a BOOT method of delivery, which allows this risk and associated permitting costs to be transferred to the private BOOT contractor.

Entitlement risks. Entitlement risks are mainly risks associated with control and costs of use of the existing site, and infrastructure on which the desalination plant, and associated facilities will be located. In the case where the desalination plant would share existing intake and discharge infrastructure with other facilities, such as power plants or wastewater treatment plants, entitlement risks are associated with potential changes of the technology and capacity of the existing host facilities in the future. Of particular importance are changes that may require the desalination plant to build its own intake and/or discharge facilities or to modify significantly the existing plant structures in order to accommodate the necessary changes implemented by the host power plant. For example, if a desalination plant uses existing wastewater plant outfall for concentrate discharge, and the wastewater treatment facility owner decides to expand plant capacity and therefore, revoke or decrease the allowable volume of the concentrate discharge through plant outfall, the desalination plant will face the need and expenditures to build its own outfall, unless it is contractually entitled to use a predetermined discharge throughout the useful life of the plant. In this example, if the desalination plant does not have a contractual entitlement for use of the wastewater plant's outfall under any circumstances over the period for which a lending institution would fund the project, the lender will consider this condition a risk and will penalize desalination project financing costs to provide adequate protection of lender's investment against this risk. The size of the interest rate penalty of the borrowed funds will be commensurate with the additional expenditures needed to address this risk in the future.

Power supply risks. Power supply risks are risks associated with the availability of power and the magnitude of the unit power cost change over the useful life of the desalination project. Since cost of power is usually 20–35% the total water production cost, the financial institution funding the project would require the plant operation costs to be secured with a long-term power supply contract that allows to project the changes of unit power costs in the future. Financial institutions would typically expect the power tariff adjustments allotted in the power supply agreement to be matched with the water tariff adjustments in the water purchase agreement.

Construction risks. Construction risks are risks associated with increase in the construction costs during the project implementation period due to: unusual site subsurface conditions; delay of delivery and installation of key equipment; construction cost overruns; and performance and reliability risks driven by plant commissioning and acceptance testing delays. Well-recognized construction companies with proven track record in successfully constructing seawater desalination projects in similar settings would greatly increase the confidence level of the lenders involved in the project financing. Typically, turnkey fixed price contracts which allow holding key contractors to their performance obligations are favored by project lenders. Construction contract completion guarantees with clauses which require performance and payment bonds of size of 10–30% of the turnkey construction price to be available to the lenders to rectify construction problems are preferred by the financial community as a proven mechanism to mitigate construction risks. Typically, the size of the performance and payment bond is commensurate with the probability of the contractor's default, which in turns is related to the previous track record of the contractor with similar projects and contractor experience with the technology proposed to be used for the desalination project.

Source water risks. Source water related risks are associated with the impact of the source seawater quality on desalination plant performance and with the effect of potential changes in source water quality over the useful life of the desalination project on the water production cost. For example, increase in source water turbidity, organics or other compounds that may result in accelerated fouling of the membrane elements or in the need for more elaborate pretreatment are typically of concern of the project lenders. These issues can be addressed by avoiding the location of the desalination plant intake in the vicinity of existing wastewater treatment plant discharges, industrial discharges or large industrial or commercial ports and shipping channels. In BOOT projects source water risks are contractually addressed by including a source water quality specification in the water purchase agreement, in the agreement for turnkey engineering, construction and procurement services, and in the plant operation services agreement. These agreements should also contain provisions for cost of water adjustments when the actual source water quality is outside of the contractual specifications and when deviations from the source water quality specifications have a material impact on desalination plant performance and costs.

Technology risks. Technology risks are related to the potential downsides of using new and unproven technologies with limited or no track record on large-scale desalination plants. Although use of new technologies typically has performance benefits such as reduced construction costs, and power and/or chemical consumption and costs, the potential downside is an increased plant downtime due to equipment failure or malfunction of key system components. While project engineers typically tend to focus on the cost and performance advantages of new technologies, project lenders always look at both potential upside and downside effects on a life-cycle cost basis when evaluating risks and benefits associated with using new technology for a given project. If potential project downsides outweigh cost savings over the useful life of the project, than the technology is considered higher risk and lenders may penalize rather than reward the use of new technology for a given project.

For example, if the use of a new energy recovery technology (ERT) under average conditions allows reducing power consumption by 10% as compared to a conventional ERT with well-proven track record, but the downtime of the new ERT is 10% higher than that of the conventional ERT, than the overall life-cycle cost effect of the use of the new ERT may be negative and therefore, it may be considered unfavorably by the financing community. For the specific conditions of a given project, let's assume that: (1) the desalination plant capacity is 20,000 m^3/d (5.3 mgd); (2) the savings from using new energy recovery technology are 0.66 kWh/m^3 (2.5 kWh/1,000 gal); (3) the unit cost of power is US\$0.06/kWh; (4) the new energy recovery technology has 10% higher downtime than the older technology it is compared with, and (5) the cost of desalinated water sold to the customer is: US\$0.70/m^3 (US\$2.65/Kgal). For these conditions, the annual benefit of using the new energy recovery system is: US\$0.06/kWh × 0.66kWh/m^3 × 20,000 m^3/d × 365 d × 90% of the time of availability = US\$ 260,172. However, the loss of water sales because of 10% downtime of the new energy recovery system is: US\$0.70/m^3 × 20,000 m^3/d × 365 d × 10% of downtime = US\$511,000. As shown in this example, the penalty for lower reliability of the new technology is significantly higher than the benefit of increased energy savings this technology yields. As a result, the use of new ERT for this project is not warranted and represents a risk of US\$511,000 – US\$260,172 =

US$250,828/y. Usually, the project lender would turn this risk into a cost overrun amortized over the term of lender's investment and than into an incremental increase in the interest rate of the funds which the lender commits to the project. For example, let's assume that the fair and favorable market interest rate for lending money to a low-risk project with conventional ERT is 5.0%; the term for repayment of the investment is 20 years; and the total capital cost of the project is US$34 million. The capital cost recovery (amortization) factor for this project is 12.462 (see Section 18.6.1 for calculation of the amortization factor). Therefore, the project annual amortized cost would be: US$34 MM/12.462 = US$2.73 MM/y. Because of the potential O&M cost overrun of US$250,828/y due to the use of the new ERT, the annual amortized project cost would increase from US$2,730,000 to US$2,980,828 (US$2,730,000 + US$250,828). As a result, the actual amortization factor would decrease from 12.462 to 11.413 (US$2,730,000/US$2,980,828). For a 20-year loan term, the interest rate corresponding to this amortization factor is: 6.07%. In effect, because of the risk associated with the use of the new ERT, the project lender may raise the interest rate of the invested capital from 5.0–6.07% to cover potential risk of repayment of investment and loss associated with the use of the new ERT. In this case, the cost of water production may be penalized twice: once because of the increased interest rate of the borrowed funds to build the desalination plant and second time, because the owner would loose revenue from water sales due to the increased project downtime.

This example illustrates the fact the use of new technology although attractive from an engineering point of view may not always be beneficial for reduction of the overall project costs, and in reality may often penalize the cost of water production through the increased cost of project financing.

Although, fairly simplified, this analysis illustrates the monetary value of technology risk from the point of view of the investor and of the importance of using equipment with proven track record. In general, if a new technology is introduced and the technology lacks full-scale track record of actual availability (downtime), assumption of 5–10% of downtime of the equipment is commonly used by the financial community to evaluate technology risks. This assumption stems the fact that new technology used for the first time in a large scale on a given project usually goes through two to three generations of improvements until it reaches a typical reliability of well proven and mature technology (i.e., technology with downtime of less than 1% and full-scale track record of 5 years).

The example above also illustrates the fact that use of new technology is more acceptable and warranted for projects where the potential for benefits is significantly higher than the penalties associated with equipment downtime and lost revenue form water sales. If, for the example above, the unit cost of power was US$0.15/kWh rather than US$0.06/kWh and/or the cost of water was lower, than the use of the new energy recovery equipment would be still warranted and the project financing would not be penalized. This explains the fact that the new pressure-exchanger (isobaric-chamber) based energy recovery systems have found their first full-scale applications on relatively small projects in the Caribbean where despite of the relatively high downtime of the first prototype pressure exchangers as compared to other mature energy recovery technologies, the unit power costs are so high (typically between US$0.10/kWh and US$0.25/kWh) that the overall benefits of isobaric-chamber ERT use clearly outweigh the potential downsides. Similar risk profile, but a different reason—lower cost of water rather than higher unit cost of power—among other factors, have made the use of the pressure exchanger technology an acceptable financing risk for a number of recent large desalination projects as well.

Regulatory risks. Regulatory risks are risks associated with the material impact that the change of regulations applicable to a given project may have on desalination plant construction and/or O&M costs. Change in regulations may occur during the period of desalination plant construction (for example, changes in electrical or building codes) and/or during the period of plant operations (i.e., new product water quality regulatory requirements or changes in waste stream discharge regulations). Financial community typically looks for flexibility features in the plant design that would allow accommodating future regulation-driven technology changes, and for contractual provisions that permit regulatory risks to be mitigated through cost-of-water tariff adjustments.

Operational risks. Operational risks are risks associated with the operations and maintenance of the desalination plant over the useful life of the facility. Consistent and reliable plant operation and maintenance is the key to generating adequate, predictable and steady revenue stream required to meet financial obligations associated with project financing, and to comply with water purchase agreement's penalty provisions for non-performance and regulatory non-compliance. Because the contracting of the desalination plant O&M services to a well established specialized private contractor with a proven experience usually results in lower financing costs, many plant owners are willing to contract plant O&M services to third parties, especially if they do not have in-house experience with O&M of desalination plants.

Desalinated water demand risk. Desalinated water demand risk is closely related to the need for high-quality water in the service area of the desalination plant, and to the affordability of this water as compared to existing water supply sources. Typically, in a public-private partnership, the project lender would look for a "take-or-pay" provision in the BOOT contract which assures that a predetermined minimum volume of desalinated water is purchased by the final user under all circumstances or alternatively the final user pays for this minimum amount of desalinated water independently of its use. Lending community would assign a high water demand risk to conditions where the costs of alternative traditional water supplies (groundwater and surface water) are significantly lower than these of the cost of desalinated water, and where the need for desalinated water is driven by temporary drought or shortage of fresh water. Financing community concerns associated with lack of steady desalinated water demand may be mitigated by putting in place a project funding structure that provides a temporary subsidy for the use of desalinated water, which is of size equal to the difference between the cost of desalinated water and the cost of water of other existing sources. Examples of such subsidies are the US\$0.32/m^3 (US\$1.20/1,000 gal) credit given to the Tampa Bay Water by the South West Municipal Water District for the potable water produced at the Tampa Bay Water seawater desalination plant, and the US\$0.20/m^3 (US\$250/acre-ft) credit given by the Municipal Water District of Southern California, US to its customers for the use of desalinated seawater. Similar direct or indirect mechanisms of reducing the water demand risk are used at state or local government levels through the world for almost all existing seawater desalination projects today. In many countries, the desalination cost subsidy is implicitly provided at governmental level, often by the state or local government taking upon a number of the risks presented above by providing payment guarantees, and thereby indirectly subsidizing desalinated water costs.

Financial risks. These risks are directly related to the financial strength (credit) of the entity which will be the final user or the desalinated water, and which will be responsible for all payment obligations associated with the project financing as well as of the parties involved in the project construction and operation. Project lenders would favor financial agreements with entities that have proven track record in servicing and repayment of debt and equity obligations on similar projects, and which do not carry excessive amount of previous fiscal obligations. Other financial risks are the risks associated with the political stability of the country in which the project is located and the country's currency stability (currency risk). Many of the financial risks associated with a given desalination project may be addressed cost-effectively by involvement of the private sector in project financing.

Before financial institutions commit to fund a given project they carefully quantify the risks described above and typically address the outstanding risks that are not already adequately mitigated by contractual and technical means, through the incremental increase in the interest rate of the funds they lend.

18.2.3. Project delivery and financing method

Project delivery and financing method has a significant effect on the cost of seawater desalination. Although desalination projects worldwide have been delivered under a number of different methods and financial arrangements, most of cost-of-water reduction breakthroughs to date have been achieved under a BOOT method of project delivery. A more detailed discussion of the alternative methods of project delivery and their effect on project costs are presented in Section 18.8.4 of this chapter.

18.2.4. Other project cost factors

Other factors that have a measurable impact on project costs are listed below:

- Intake system type and design: open ocean intake, co-location of intake and discharge with existing power plant, or subsurface intakes (beach wells)
- Pretreatment system type and design
- SWRO system configuration:
  - ○ Number and size of individual RO trains
  - ○ Redundant capacity
  - ○ Number of vessels per RO train
  - ○ Number of SWRO membrane elements per vessel
  - ○ Number and location of points of permeate collection from the individual vessels
- Architectural design
- Structural design:
  - ○ Buoyancy control
  - ○ Foundations (piles, slab footings, etc.)
- Electrical design:

  - Power source (electrical grid; self-generation; direct connection to existing power generation units)
  - Site power supply system configuration–location and size of power substations and connecting conduits
- Selection of key materials (piping, equipment and structures)
- Site work including:
  - Plant layout
  - Lightning
  - Roadways
  - Site drainage and stormwater management
- Corrosion control:
  - Protective coatings of structures, equipment and piping
  - Cathodic protection

18.2.5. Cost factors beyond the control of the plant owner

The key subjective and site-specific cost factors which are usually outside of the control of the plant facility owner but may have very significant impact on the overall project costs are:

- Regulatory design standards, building and fire codes
- Schedules mandated by third parties (regulatory agencies, emergency response needs, etc.)
- Conventions of engineering practice
- Comprehensiveness of construction and equipment and consumable supplier markets
- Local labor and material costs and shortages
- Construction time constraints driven by local noise and traffic related ordinances, and limitations of hours of operation of construction equipment
- Use and condition of existing facilities
- General economic climate
- Climatic conditions

- Seasonal water demand and power tariff variations
- Land costs and site subsurface conditions (i.e., soil and groundwater contamination; soil load bearing capacity and liquefaction potential; and subsurface obstacles)

The factors listed above are very site and project specific. The costs that account for these factors can contribute over 100% variation to the baseline project costs and therefore, have to be taken into account, especially when preparing budgetary and detailed cost estimates, and when comparing the costs of two projects of similar capacity, source and product water quality, and intake and discharge conditions. The effect of some of these uncontrollable and often unpredictable factors on the project costs may be mitigated by acquiring appropriate insurance policies for the project.

18.3. Types and accuracy of project cost estimates

There are several major types of cost estimates depending on the purpose for which the estimates are used, their accuracy, and the associated level of advancement of the SWRO project development. The key types of project cost estimates and their purpose, and accuracy are summarized in Table 18.4 and discussed below.

18.3.1. Conceptual cost estimate

Conceptual cost estimate is developed during the initial planning/scoping phase of a SWRO project, and its purpose is to determine an order-of-magnitude value of the project capital, O&M and water production costs. Conceptual cost estimate is typically used for preliminary screening of water supply alternatives; for general cost-of-water comparisons with other existing or planned alternative water supply sources (such as surface water, reclaimed water, brackish or groundwater, etc.); for desalination plant site screening; and for preparation of initial fatal-flow analysis for a given desalination project.

The conceptual cost estimate is usually prepared without detailed engineering data or comprehensive knowledge of the final project scope and is based on cost-capacity curves available in the literature (see references 126, 128, and 129) or on scale-up or scale-down empirical factors and cost information from existing projects of similar scope, source and product water quality, location and size. The Advanced Water Treatment Database (Desalnet) developed and

TABLE 18.4
Types and accuracy of project cost estimates

Estimate type	Cost basis	Purpose	Expected accuracy (% of actual costs)
Conceptual (incremental budgeting)	• Rudimentary Project scope and Conceptual design • Costs of similar projects • Scale factors • Cost–plant capacity curves and tables	• Conceptual planning • Fatal-flaw analysis • Project scope definition	–50% to +100%
Preliminary	• Preliminary project design • Cost models • Cost graphs, formulas and tables for individual treatment processes and equipment	• Process, technology and equipment selection • General evaluations • Guidance for future investigations	–30% to +50%
Budgetary	• Advanced project development and design • Budgetary vendor quotes for key equipment, piping and facilities • Cost estimates based on sizing and quantification of materials and labor for construction of key facilities	• Facility owner budget • Project authorization	–15% to +30%
Detailed (zero-based budgeting)	• Detailed project design • Equipment and material specifications and quantification • Firm vendor quotes/ purchase orders	• Guaranteed subcontractor prices for various activities • Preparation of project tender (bid) price • Control of project implementation	–5% to +15%

maintained by the United States Bureau of Reclamation (USBR) and the American Water Works Association (AWWA), contains numerous articles and reports that have cost-estimating curves, which could be used for the preparation of conceptual cost estimates.

Often, the preparation of the conceptual cost estimate is referred to as "incremental budgeting" (129), because this cost estimate for the new project is derived from the actual cost of an existing similar project or projects used as benchmark/s to which incremental "plus" or "minus" cost factors are applied. A key factor of critical importance in the preparation of conceptual cost estimate based on "benchmark" project costs is to have a detailed understanding of the scope of the benchmark project/s and the actual cost items incorporated in the benchmark project costs. Because SWRO technology advances almost every year and the incremental budgeting is backward looking (i.e., based on projects applying older technology), the use of this "incremental" cost estimation technique often leads to conservative results and therefore, is unsuitable for preparation of competitive or budgetary cost estimates and project bids.

A very important factor that has to be taken into consideration when preparing conceptual cost estimates based on benchmark projects is to account for the differences associated with the timeframe of project implementation and the currency in which the cost of water, and construction and O&M costs are presented. For example, the cost of construction and O&M services, and ultimately the cost of water for a given project built today, will always be higher that that of the cost of water estimated for the same project built one or more years ago. The time-related cost differences are mainly due to general inflation and fluctuations in cost of key construction materials (such as steel, lumber and fuel), labor, equipment and consumables. Therefore, if old cost curves or benchmark project costs are used to develop conceptual costs for a new project, the difference of time of construction of the two projects has to be reflected in the conceptual cost estimate.

In the USA inflation and change in costs of construction materials over time can be reflected in the construction cost estimate for a given project using the Engineering News Record (ENR) construction cost index (CCI). The ENR contains tabular information that allows comparing cost of construction in various countries, which in turn could be used to further refine the conceptual cost estimate for a new project based on cost information for an overseas benchmark project. The ENR also contains detailed information on cost of key construction materials which allows to identify mayor cost trends and to forecast potential construction cost changes over the period of project implementation.

There are two types of CCI indexes–one is a nationwide cumulative index and the other type is a regional index provided for 20 individual large US cities and the regions around them. While the nationwide ENR CCI index is suitable to be used for comparison of costs of projects in the US and these in other parts of the world, the City-specific CCI indexes allow to account for the region specific-impacts on costs such as costs of labor, chemicals, and construction materials in the US. These costs may vary by 30% or more in different part of the country and therefore it is important to reflect the location-related cost differences.

Another important factor that has to be taken under consideration when preparing conceptual cost estimates and comparing projects is the currency conversion ratio. Many countries experience fluctuations in the currency conversion ratio against the Euro, the US$, or the Yen. Such currency conversion ratio fluctuations over time have to be reflected in the preparation of conceptual cost estimates based on benchmark projects.

Usually, a conceptual cost estimate is based on desalination plant production capacity only and therefore, is very rudimentary. The level of accuracy of this estimate is relatively low and is typically between –50% and +100% of the actual costs.

18.3.2. Preliminary cost estimate

Preliminary cost estimate is typically used for: initial site-specific project cost assessment; for evaluation of alternative treatment processes and equipment; and for guidance for further investigations. This cost estimate is prepared during the planning stage of the project after the project scope has been clearly defined; the overall treatment approach and key processes and technologies have been selected; and the conceptual design of the main plant treatment facilities have been completed. Often various cost models are used to assess the preliminary costs for a given desalination project. The key cost models currently available for the preparation of preliminary cost estimates are discussed in Section 18.3.5. of this chapter.

As a minimum, the following project related information has to be available in order to develop an accurate preliminary cost estimate:

- Average annual, daily average, minimum and maximum SWRO plant production capacities
- Design plant capacity availability factor
- Source water quality specification

- Product water quality specification
- Plant intake and discharge type, size and configuration
- Selection and size of key facilities, equipment and piping for:
  - Source water pretreatment
  - SWRO desalination and energy recovery
  - Product water post-treatment
  - Concentrate disposal
  - Solids and liquid waste handling
- Process flow diagram
- Preliminary facility layout
- RO system performance projections
- Solids mass balance

Typically, preliminary cost estimates carry a significant contingency (30% or more) to address a number of site-specific unknown factors such as site soil conditions under the main project structures; plant hydraulics; site-specific costs of construction services, labor, and materials; and other site-related costs. Because the preliminary cost estimate is based on actual information for the given project rather than on the extrapolation of costs from other projects, the accuracy of this estimate is higher that of the conceptual cost estimate: –30% to +50% of the actual costs.

18.3.3. Budgetary cost estimate

Budgetary cost estimates are mainly used for authorization of project implementation and for project inclusion into owner's fiscal planning and budgeting process. In addition to the project information required for preparation of preliminary cost estimate, the budgetary cost estimate as a minimum necessitates refinement of the project design and costs based on:

- Preliminary geotechnical and hydro-geological Investigations
- Preliminary design of:
  - Key project structures and foundations
  - Electrical supply system

    - Instrumentation and control system
    - Architectural facades and appearance of key buildings;
- Preparation of plant hydraulic profile
- Development of basic specifications of key equipment and piping, equipment data sheets and budgetary quotes from vendors
- Preparation of general project implementation plan and schedule

The budgetary cost estimate typically accounts for all key site-specific factors that have measurable (over 10%) influence on the project costs. This estimate typically carries 20–25% of contingency. The accuracy of the budgetary cost estimate is usually within –15% to +30% of the actual project costs.

18.3.4. Detailed cost estimate

The detailed cost estimate is prepared during the contractor procurement phase of the project and is often used by both contractors and owners to determine the project tender (bid) contract price and the most probable project construction, O&M and water production costs. The preparation of a detailed cost estimate is often referred to as "zero base budgeting" because all key project cost components are determined based on genuine, site specific estimates rather than on cost comparisons derived from other similar projects or from empirical experience.

Detailed cost estimates are based on:

- Advanced level of project design (30–50% of design completion)
- Detailed construction survey
- Detailed geotechnical investigation and soil analysis
- Comprehensive project implementation schedule
- Detailed quantification and cost estimates of key construction activities
- Binding vendor price quotes for all equipment and prefabricated facilities of unit value in excess of US$10,000 including:
    - Source water intake, screening and pretreatment equipment
    - RO and pretreatment membranes and cartridge filters
    - Large pumps

- Energy recovery equipment
- Stainless steel and large size piping
- Construction materials and labor for equipment and facility installation, start-up and commissioning
- Automation and control systems
- Chemical feed systems
- Electrical substations and conduits
- Consumables (such as chemicals and power)

The detailed cost estimate typically includes contingency equal to 5–10% of the total project capital cost. The size of project contingency of this type of cost estimate is usually commensurate with the contractor profit margin, which is based on the fact that the contractor cannot put at risk more than its potential profit from project implementation. Typically, the upper end of the contingency range is used for establishing the not-to-exceed contractor binding project cost offer. The accuracy of this type of cost estimate is between –5% and +15% of the actual costs.

18.3.5. Cost models

Currently, there are several models and cost reference tools that can be used for the preparation of preliminary cost estimates for membrane seawater desalination projects. The most recent cost estimation program entitled WTCost©, is developed by the United States Bureau of Reclamation (USBR) in April of 2002 (131). This computer model uses Visual Basic and is based on updated cost curves developed by the United States Environmental Protection Agency (US EPA) in 1979 (132) and upgraded Excel spread sheet: Water Treatment Estimation Routine (WaTER) developed by the USBR in 1999. The WTCost© model allows to reflect the type of the selected pretreatment system (conventional gravity filters vs. microfiltration (MF) or ultrafiltration (UF) membranes); pretreatment chemicals (coagulants, lime, caustic soda, antiscalant and powdered activated carbon); and the type of the selected salt separation process: RO, ion exchange or electrodialysis), and can be used for estimating construction, capital and O&M costs for both seawater and brackish water desalination plants. This cost estimating program allows also taking into account the selected type of desalination plant intake and discharge, the post-treatment technology, and the source and product water quality.

All input data needed to run the WTCost© model such as: source water quality information; power and chemical use and prices; labor costs and rates; construction indexes; and cost of capital have default values that can be modified by the user. The cost model is membrane supplier neutral and allows to either use membranes included in the model database or to incorporate different membranes and membrane element costs.

Another cost model that is available for the preparation of preliminary cost estimates is the Reverse Osmosis Desalination Cost Planning Model©, prepared by the Water Reuse Associates (WRA) in the US (129). Similar, to the WTCost© software, this model provides itemized breakdown of the plant capital, O&M and life-cycle costs of the desalination project. The WRA model is developed and calibrated based on data from existing desalination plants, and is designed to produce results using both US and metric units. This model has the flexibility of incorporate power plant co-location configuration for the desalination plant intake and discharge.

The WRA model has two input data forms: "master" and "advanced'. The master input from for the model is designed for users with limited knowledge of seawater desalination and the advanced form is tailored to the needs of users with significant understanding of the desalination process and the economic impacts of alternative methods of project delivery and finance. The key advantage of this model, as compared to the WTCost© software, is that it provides flexibility to account for site-specific economic factors such as: local and state sales taxes, subsidies, and changing amortization rates which are used for some types of project finance. In addition, the WRA model produces a chart gallery that allows the user to visualize the relative contribution of various cost components to the total cost of the desalination project capital, O&M and water costs. The key limitation of the WRA model as compared to the WTCost© model, is that this model is designed to produce cost estimates for sea and brackish membrane desalination plants only and does not include alternative desalination technologies.

18.4. Preparation for project budgeting

The preparation of an accurate cost estimate for a given SWRO project requires comprehensive understanding and quantification of all key factors that have significant cost impact. These factors are discussed in detail Section 18.2 of this chapter. In addition, the extent of the preparation efforts is closely related to the type of the prepared cost estimate as presented in Section 18.3.

18.5. Project cost estimation and analysis

18.5.1. Capital costs

The project capital costs can be divided into two broad categories: construction costs (sometimes also referred to as “direct capital costs″) and other project related capital costs (engineering, development, financing, and contingencies), which are often referred to as “indirect capital costs”. A typical breakdown of the project capital costs for low-complexity and high-complexity desalination projects in a capacity range of 5,000 m^3/d to 200,000 m^3/d (1.3 to 52.8 mgd) is presented in Table 18.5. This table shows cost breakdown for both low-complexity and high-complexity projects. The project complexity is determined based on the size of the project; the source water quality and its variability; the type of plant intake; the method of disposal of concentrate and other plant waste streams; the complexity of permitting regulations governing project implementation; and the financing sources and structure. Usually low-complexity projects are:

- Relatively small plants (i.e., projects of production capacity of 20,000 m^3/d, 5.3 mgd or less) located in non-environmentally sensitive areas with friendly local community
- Plants with good source water quality: turbidity (NTU) and SDI of less than 1; trace levels of organics and bacterial contamination; and very low content of fouling and scaling constituents
- Plants with subsurface or open intakes which collect seawater without significant interference from contaminated surface fresh water sources, groundwater aquifers or waste discharges
- Plants with simple, low-cost concentrate disposal methods, such as direct sewer or ocean discharge with suitable environmental conditions which do not require waste stream treatment prior to discharge and construction of complex discharge structures such as diffusers
- Regulatory environment where the key regulating agencies involved in the project permitting process have experience with similar size desalination projects and adequate expertise to complete project environmental review in an expeditious and timely manner
- Projects which have simple and well developed financing and tariff structure where project costs and revenues and risks and rewards are well balanced, and where the cost of desalinated water is competitive to that of other available water sources

TABLE 18.5
SWRO project capital cost breakdown

	Percentage of total capital cost (%)	
Cost item	Low-complexity project	High-complexity project
Direct capital (construction) costs		
1. Site preparation, roads and parking	1.5–2.0	0.5–1.0
2. Intake	4.5–6.0	3.0–5.0
3. Pretreatment	8.5–9.5	6.0–8.0
4. RO system equipment	38.0–44.0	30.5–36.0
5. Post-treatment	1.5–2.5	1.0–2.0
6. Concentrate disposal	3.0–4.0	1.5–3.0
7. Waste and solids handling	2.0–2.5	1.0–1.5
8. Electrical & instrumentation systems	2.5–3.5	1.5–2.5
9. Auxiliary and service equipment and utilities	2.5–3.0	1.0–2.0
10. Buildings	4.5–5.5	3.0–5.0
11. Start up, commissioning and acceptance testing	1.5–2.5	1.0–2.0
Subtotal direct (construction) costs (% of total capital costs)	70.0–85.0	50.0–68.0
Project engineering services		
1. Preliminary engineering	0.5–1.0	0.5–1.5
2. Pilot testing	0.0–0.5	1.0–1.5
3. Detailed design	3.5–4.5	5.0–6.0
4. Construction management and oversight	1.0–2.0	2.5–3.5
Subtotal engineering services	5.0–8.0	9.0–12.5
Project development		
1. Administration, contracting and management	1.0–1.5	2.0–3.0
2. Environmental permitting	0.5–3.5	4.5–5.0
3. Legal services	0.5–1.0	1.5–2.0
Subtotal project development	2.0–6.0	8.0–10.0

continued

TABLE 18.5 *continued*
SWRO project capital cost breakdown

Cost item	Percentage of total capital cost (%)	
	Low-complexity project	High-complexity project
Project financing costs		
1. Interest during construction	0.5–2.5	1.0–4.5
2. Debt service reserve	2.0–5.5	4.5–8.5
3. Other financing costs	0.5–1.0	3.5–4.5
Subtotal project financing	3.0–9.0	9.0–17.5
Contingency	5.0–7.0	6.0–10.0
Subtotal indirect capital costs (percent of total capital costs)	15.0–30.0	32.0–50.0
Total capital costs	100%	100%

The construction portion of the capital costs varies with the size and complexity of the individual projects, and typically ranges between 50 and 85% of the total capital costs. The indirect (non-construction) portion of the capital cost is usually within 15–50% of these costs. The cost breakdown brackets presented in this table are based on data from actual seawater desalination projects completed to date. This cost breakdown encompasses most frequently encountered conditions associated with project implementation. If the site-specific conditions of a given individual project are significantly different from these encountered in desalination projects completed over the last 10 years, the actual cost breakdown for this project may be outside the of the cost brackets presented in Table 18.5. This table should be used for preparation of conceptual and preliminary cost estimates only, and is intended to reflect current “market” conditions for the individual cost items. With the advancement of membrane technology and maturing of the markets involved in funding and serving the development, construction and operation of desalination projects, the ratios between the individual cost items are expected to change over time. Gradual changes are anticipated to occur every 2–5 years. A more dramatic cost change is likely within a 10-year time frame.

18.5.1.1. Construction costs

Site related construction costs. Site related construction costs include the costs for land acquisition and site preparation for construction (plant site clearing, grubbing, fill, grading and fencing), as well as the costs for construction of access roads to the desalination plant and to all buildings, facilities, and equipment within the desalination plant. The cost of land, the expenditures for site clearing, soil contamination mitigation, and dewatering, as well as the cost and length of access roads are very site specific and could vary significantly from one location to another. In general these costs are in a range of US\$10 to US\$50/m^3/d (\$0.04–0.20/gpd) of plant production capacity. The land requirements for a typical seawater desalination plant are summarized in Table 18.6.

Intake construction costs. The intake construction costs include the costs for plant seawater intake structure and pipeline, intake pump station and screening facilities. The intake construction costs and components vary depending on the type of the intake: subsurface (vertical beach wells, horizontal directionally drilled (HDD) wells or Raney-type wells) intake; surface (open-ocean) intake; or co-located intake using existing power plant outfall or intake.

Typically, the lowest cost intake is that co-located with the discharge of an existing coastal power plant which uses seawater for cooling. The co-located desalination plant taps into the power plant discharge to collect source seawater. The co-location approach allows avoiding construction of new desalination plant intake structure, pipeline and screens, which reduces approximately 60 to 80% of the total intake construction expenditures. Other advantages of co-location are discussed elsewhere (133).

TABLE 18.6
Seawater desalination plant land requirements

Plant capacity	Typical plant site land requirements	
m^3/d (MGD)	m^2	acres
1,000/(0.3)	800–1,600	0.2–0.4
5,000/(1.3)	2,000–3,200	0.5–0.8
10,000/(2.5)	6,100–8,100	1.5–2.0
20,000/(5.3)	10,100–14,200	2.5–3.5
40,000/(10.6)	18,200–24,300	4.5–6.0
100,000/(26.4)	26,300–34,400	6.5–8.5
200,000 (52.8)	36,400–48,600	9.0–12.0

Note: Land requirements based on conventional plant layout. compact plants may require less land.

Vertical beach wells are often very cost-competitive intake facilities as compared to open ocean intakes. However their use is typically limited to small and medium size SWRO plants (134). The use of HDD wells for large-size facilities may prove beneficial for conditions where the HDD wells can collect high-quality seawater at a steady rate. The main challenge with HDD wells is maintaining their capacity and water quality over time. Cost comparisons of alternative source water intakes are provided in other sources (126, 134). Typically, the intake construction costs are in a range of between US\$50 and US\$100/m^3/d (US\$0.19–0.38/gpd).

Pretreatment construction costs. Pretreatment construction costs include the expenditures for removal of all contaminants in the seawater that may impact normal operation of the membrane separation process. The magnitude of these costs depends mostly on the content of solids (turbidity/total suspended solids); biodegradable organics; and non-organic membrane fouling compounds in the source water; and the selected type of pretreatment technologies and equipment needed for their effective removal. The pretreatment process may involve physical removal of contaminants by: coarse and fine screening and micro-screening; grit removal; sedimentation; dissolved air flotation; granular media or membrane filtration, as well as chemical conditioning of the source water to prevent non-organic scale formation (addition of antiscalants); membrane biofouling (biocides and UV irradiation); enhance boron removal (by pH adjustment); and improve solids removal by the use of coagulants and flocculants.

Because of the variety of the available technologies and equipment for solids removal and chemical conditioning, the costs associated with source water pretreatment may vary significantly from one project to another. Typically, the pretreatment costs are in a range of US\$100 and US\$300/m^3/d (US\$0.38–1.14/gpd).

RO System equipment costs. This cost item includes the expenditures associated with the procurement, purchase, installation and construction of the following facilities and equipment:

- Cartridge filters
- High pressure pumps and motors to feed the RO system
- Energy recovery system
- RO pressure membrane vessels and racks
- SWRO membrane elements
- Membrane cleaning system

- Membrane flush system
- Interconnecting piping

The construction costs of some of the key membrane SWRO system elements are summarized in Table 18.7.

Approximately 10 to 20% has to be added to the costs shown in Table 18.7 for shipping, handling, installation oversight and insurance.

The cost of the membrane RO modules (trains) is proportional to the design capacity and flux of the SWRO system. Typically, one RO module contains 50 to 200 membrane vessels and has capacity between 100 m^3/d (0.03 mgd) to 20,000 m^3/d (5.3 MGD). While there is limited economy of scale of the costs of the RO modules, the costs of the other RO system components (high pressure pumps, energy recovery devices, stainless steel piping and valves and membrane cleaning system) can benefit significantly from the use of larger size units. For example, one membrane cleaning system can be used for several RO membrane trains; the high pressure feed pump efficiency and cost improves with their size; and the economy of scale between two sizes of stainless steel pipe is usually 10–15%. Therefore, as the RO membrane module size increases, the relative cost of the SWRO system per unit volume of produced permeate decreases.

TABLE 18.7
Construction costs of key membrane SWRO system elements

Item	Construction cost (US$/item or as indicated)
8-inch SWRO membrane elements	US$400–US$550/element
16-inch SWRO membrane elements	US$2,800–US$3,200/element
8-inch BWRO membrane elements	US$250–US$350/element
SWRO pressure vessels for 8-inch elements	US$1,200–US$1,600/vessel
SWRO pressure vessels for 16-inch elements	US$4,000–US$5,000/vessel
BWRO pressure vessels for 8-inch elements	US$1,000–US$1,200/vessel
RO skid piping	US$200,000–US$600,000/RO train
RO train support frame	US$100,000–US$300,000/RO train
RO train instrumentation and controls	US$20,000–US$80,000/RO train
High pressure pumps	US$100,000–US$800,000/RO train

Note: All costs in year 2005 US$.

There are two limitations of the maximum size of the membrane module, which limit the benefits of using the largest possible size module for a given application–the system reliability and the available off-the-shelf equipment that can be used to build a very large module. The main limiting factor with using large scale modules is the loss of production capacity when the RO module is shut down for membrane cleaning, replacement or equipment and piping repairs. The larger the individual module, the larger the loss of capacity and therefore, the lower the availability factor of the SWRO plant. Since the availability factor is directly related to the cost of water, a SWRO system with lower availability factor yields higher cost of water. Another factor limiting the benefits of construction of very large SWRO trains is the need to use custom-made rather than off-the-self equipment (mainly high-pressure pumps, motors and energy recovery devices). Although the manufacturing of this equipment is possible, the one-time custom design and production of such equipment is significantly more costly than the use of the standard "off-the-shelf" equipment with well known production costs, performance parameters and proven track record. Therefore, in such applications, often the gain of the economy of scale due to the use of large custom made equipment and trains is negated by the additional expenditures for equipment production and risks associated with the lack of long-term track record of equipment performance.

The SWRO system is the most complex portion of the desalination plant and usually contributes 40–60% of the total plant construction costs. The design and construction costs of this system are mainly influenced by the source water salinity and temperature, and by the target product water quality the system has to produce. The construction cost of this system is predominantly determined by the cost of the membrane vessels and racks, high-pressure pumps and piping; cost of the energy recovery system: and the price of the SWRO membrane elements. Typically, the SWRO system construction cost varies between US\$300 and US\$1,000/m^3/d (US\$1.14–3.8/gpd).

Post-treatment costs. Post-treatment costs incorporate the costs for construction of:

- Chemical conditioning system for permeate stabilization
- Disinfection system
- Facilities for product water quality polishing

The post-treatment costs are mainly driven by the target product water quality and the final use of the desalinated water. Typically, the costs for construction of post-treatment facilities for permeate stabilization and disinfection are in a range of between US\$20 and US\$50/m^3/d (US\$0.08–0.20/gpd). However, if the permeate has to be polished to achieve high-levels of boron removal or removal of specific constituents, than these costs may increase beyond the range indicated above.

Concentrate disposal costs encompass expenditure for the conveyance and disposal of the concentrate. These costs can vary significantly depending on the concentrate disposal method. According to the results of a recent study completed by the US Bureau of Reclamation (136), the concentrate disposal methods most widely used in the US (in order of decreasing frequency) are listed in Table 18.8.

The results presented in this table are based on the survey of 203 RO, NF, and ED/EDR plants completed in year 2000. The survey included only plants larger than 200 m^3/d (0.05 mgd). Approximately 85% of the surveyed plants were nanofiltration, brackish and SWRO facilities.

The costs for the concentrate disposal methods listed above are impacted by a number of site-specific factors and therefore, a general cost estimate analysis is difficult to complete. Table 18.9 presents the typical range of the construction costs for concentrate disposal system for seawater desalination plants. Review of this table indicates that the sanitary sewer and surface water discharge are the two most cost-effective methods for concentrate disposal, which explains their popularity. Depending on the site-specific conditions, deep well injection, evaporation ponds and spray irrigation could be competitive concentrate disposal alternatives. Zero liquid discharge system typically has highest construction and operation costs. However, under specific circumstances (such as cold climate,

TABLE 18.8
Concentrate disposal methods and their frequency of use

Concentrate disposal method	Frequency of use (% of surveyed plants)
Surface water discharge	45%
Sanitary sewer discharge	42%
Deep well injection	9%
Evaporation ponds	2%
Spray irrigation	2%
Zero liquid discharge	0%

Source: (136)

TABLE 18.9
Construction costs for concentrate disposal

Concentrate disposal method	Disposal construction cost US$/m^3/d (US$/gpd)
New surface water discharge (new outfall with diffusers)	30–200 (0.11–0.76)
Co-location of desalination plant and power plant discharge	5–20 (0.02–0.08)
Co-disposal with wastewater treatment plant discharge	20–120 (0.08–0.46)
Sanitary sewer discharge	5–60 (0.02–0.23)
Deep/beach well injection	150–350 (0.57–1.33)
Evaporation ponds	200–650 (0.76–2.47)
Spray irrigation	350–850 (1.33–3.23)
Zero liquid discharge	900–2,000 (3.42–7.60)

Note: All costs in year 2005 US$.

low evaporation and soil uptake rates, high land costs and low power costs) the zero liquid discharge systems could be cost competitive to evaporation pond and spray irrigation disposal alternatives. More detailed breakdown of the construction and operation and maintenance costs for the presented concentrate disposal methods is provided elsewhere (126, 136). Key factors that have influence over the different methods for cost disposal are discussed below.

The costs for surface water discharge are influenced by a great number of site specific factors and are difficult to generalize. The key factors that determine the costs of concentrate discharge to surface water are: (1) the conveyance costs to transport the concentrate from the desalination plant to the surface water discharge outfall; (2) the costs for outfall construction and operation; and (3) the costs associated with the monitoring of the environmental effects of the concentrate discharge on the surface waters. The costs for concentrate conveyance are typically closely related to concentrate volume and the distance between the desalination plant and the discharge outfall. The outfall construction costs are very site specific and in addition to the outfall size and diffuser system configuration (which is driven by the concentrate volume and salinity) these costs are dependent on the outfall length and material, which in turn is determined by the site specific surface water body hydrodynamics conditions.

Sanitary sewer discharge is usually very site-specific and the key cost elements for this disposal method are the cost of conveyance (pump station and pipeline) and the costs and fees for connecting to the sanitary sewer and treatment/disposal of the concentrate.

The key factors that influence deep well injection construction costs are the well depth and the diameter of well tubing and casting rings. The well diameter seems to have a very limited influence on the costs. Several other key cost factors are (1) the need for concentrate pretreatment prior to disposal; (2) pump size and pressure which vary depending on the geologic conditions and depth of the injection zone; (3) environmental monitoring well system size and configuration; and (4) site preparation, mobilization and demobilization.

The costs of concentrate evaporation pond systems are mainly driven by the evaporation rate (climate); the concentrate volume; the land and earthwork costs; the liner costs and the salinity of the concentrate, which determines the useful life of the ponds. The main cost variable is the evaporative area. Typically, evaporation rates are lower than soil uptake rates and therefore, disposal of the same volume of concentrate using evaporation ponds requires more land than disposal by spray irrigation.

Spray irrigation is typically cost effective only if the concentrate is blended with a fresh water source to reduce its salinity to a level acceptable for crops/vegetation irrigation and its feasibility depends on the type of the crops/vegetation and on the soil uptake rates. The key cost factors of this disposal method are the costs of land, the storage and distribution system costs and the irrigation system installation costs, which in turn are driven by the concentrate volume and salinity.

Achieving zero liquid discharge is usually the most costly method for concentrate disposal, because it requires the use of elaborate mechanical equipment for evaporation, crystallization and concentration (dewatering) of the salts in the concentrate. Although this method has found practical application in industrial water reuse facilities, it has not yet been used for disposal of concentrate from large seawater desalination plants.

Waste and solids handling costs. These costs include expenditures for construction of facilities for collection, conveyance and disposal of solid waste (spent membranes, cartridge filters and waste solids) from the plant site as well as for the construction of solids handling system for treatment and disposal of residuals generated during the pretreatment process (screenings; residuals settled in the sedimentation tanks; solids from the filter backwash water). In addition, these expenditures also encompass costs for equipment and storage tanks for collection, conveyance and treatment (if necessary) of the waste membrane cleaning chemicals to their final disposal site (typically the sewer system).

Usually, the system for collection and disposal of waste membrane cleaning chemicals consists of storage tank, and pumps and piping used to convey the spent cleaning chemicals to the storage tank and from the tank to the nearby sewer system. The cost of this system is typically in a range of US$10–US$40/m^3/d (US$0.04–0.15/gpd).

SWRO systems which use open ocean intakes would generate a large amount of solids as well. If regulatory constraints limit the disposal of these solids back into the ocean, the filter backwash solids will have to be settled, dewatered and disposed off to a landfill. The expenditures for construction of solids handling facility for backwash residuals are typically in a range of US$10 and US$50/m^3/d (US$0.04–0.19/gpd).

Costs of electrical and instrumentation systems. These costs include all expenditures for: the desalination plant's electrical supply system (electrical substation, equipment and conduits connecting the desalination plant to the electrical grid or to a power self-generating facility); the equipment transformers and motor control centers; and all electrical conduits and equipment connecting the plant electrical system to the individual electrically-driven equipment). The electrical system construction costs incorporate the expenditures for emergency power generation equipment as well. These costs also include the expenditures for plant instrumentation and controls. The plant electrical and instrumentation costs are usually 5–15% of the total construction costs and range between US$30 and US$90/m^3/d (US$0.11–0.34/gpd).

Costs of auxiliary and service equipment and utilities. The facilities in this category are the plant chemical storage and feed systems; process air and water supply facilities; the plant fire protection system; sanitary wastewater collection system; stormwater management system; and all utilities needed for the normal plant operation (potable and utility water; telephone, etc.). These costs also incorporate the expenditures for an initial set of spare parts for the desalination plant facilities. The expenditures for construction of auxiliary and service equipment and utilities are usually between US$20 and US$60/m^3/d (US$0.08–0.23/gpd).

Building costs. Typically, the desalination plant has one or more buildings that house: plant administration and management; laboratory; operator locker and shower facilities; maintenance shop; equipment and chemical storage area; and the key equipment of the SWRO system (high pressure pumps; membrane vessels and racks; etc.). Depending on the complexity and size of the desalination plant, as well as its location, appearance and ambient environment, the construction costs for the desalination plant buildings may vary from US$5–15/m^3 of the building footprint and range between US$50 and US$150/m^3/d (US$0.19–0.57/gpd).

Start-up, commissioning and acceptance testing costs. These costs include all expenditures for labor, consumables (electricity, chemicals, etc.) and equipment used during the plant commissioning, start up and acceptance testing process. These expenditures typically also incorporate the costs for construction related permitting and insurance, for preparation of plant operation and maintenance manuals; for initial training of the permanent desalination plant O&M staff; and for equipment and other items that are required for normal plant operations (tools for the workshop; service vehicles for plant operations staff and management; furnishings and equipment for the plant laboratory and administration building; etc.). These costs also incorporate all expenditures associated with the use of outside services, such as lab analysis of all source and product water quality parameters that cannot be completed in-house. Depending on the complexity of the project these costs can vary in a wide range—US$25 and US$75/m^3/d (US$0.10–0.29/gpd).

18.5.1.2. Costs for project engineering services

Preliminary engineering. Preliminary engineering costs encompass all expenditures associated with initial assessment of project feasibility, definition of project scope and size, as well as studies required to determine the project location; the type of project intake and discharge and the configuration of key project facilities and equipment: i.e., intake, pretreatment, RO separation, concentrate disposal; permeate post-treatment, and product water conveyance and delivery. The preliminary engineering costs are very dependent on the project size and complexity. These costs range from US$10 to US$30/m^3/d of project's product water capacity (US$0.04–US$0.11/gpd).

Pilot testing. Pilot testing is highly recommended for large desalination projects (i.e., projects of production capacity of 40,000 m^3/d (11 mgd) or higher, especially when the source water is an open ocean intake or a series of large beach wells, such as HDD or Ranney-type wells. Typically the purpose of pilot testing is to:

- Assess feasibility of seawater desalination plant intake and concentrate discharge technologies and configuration

- Generate technical data required for project permitting, such as the characteristics of the plant waste streams (concentrate, filter backwash, spent chemical cleaning solution, solids residuals, etc.)
- Assess feasibility of alternative pretreatment technologies and energy recovery technologies for the site-specific project conditions
- Provide information needed for detailed project design and implementation
- Create opportunities for public outreach and education regarding the benefits of seawater desalination, and advantages of desalinated seawater as compared to existing water supply sources

Although pilot testing costs are relatively high—US\$10/m^3/d to US\$30/m^3/d (US\$0.08–0.11/gpd), they usually are a good investment towards the successful implementation of large desalination projects. In addition to the costs for constructing a pilot plant, additional operational costs of US\$10,000 to US\$20,000 per month for pilot operations and maintenance have to be budgeted. In order to be truly beneficial for a given project, pilot testing has to be completed for a period of at least 6 to 12 months and has to encompass conditions of challenging source water quality such as: frequent rain events; red-tide blooms; large but infrequent waste discharges, and dredging or intense boat traffic in the plant intake area.

Detailed design. Development of detailed project drawings and specifications typically encompasses expenditure in a range of US\$50/m^3/d to US\$150/m^3/d (US\$0.19–0.57/gpd). Detailed project design also includes the development of as-built drawings and specifications that document the actual project implementation and deviations from the original design during construction.

Construction management and oversight. Construction management and oversight include all engineering activities associated with project construction as well as the management of the construction contractors and suppliers involved in project implementation. The construction management and oversight costs range between US\$30 and US\$100/m^3/d (US\$0.11–0.38/gpd).

18.5.1.3. Project development costs

Project development costs comprise of all desalination plant owner's expenditures associated with the implementation of the project from its inception, planning, administrative review and budgeting, to environmental permitting, retaining contractors for project construction and implementation, obtaining funds for project construction, and staffing of desalination plant operations.

Project administration, contracting and management. Project administration, contracting and management are owner's responsibilities which usually involve in-house expenditures for plant staff and overhead associated with project implementation as well as expenditures for contracting outside engineering and other advisors to provide specialized support services as needed. Expenditures associated with these efforts depend on the owner's in-house capabilities and experience with the implementation of seawater desalination projects and vary between US\$30 and US\$100/m^3/d (US\$0.11–0.38/gpd).

Environmental permitting. Expenditures associated with environmental permitting include two key elements: (1) costs for preparation of environmental studies and engineering analysis needed to obtain environmental permits and (2) fees associated with environmental permit filing and processing.

Environmental permitting efforts and associated costs depend on the size and complexity of the desalination project, on the methods planned to be used for disposal of the desalination plant concentrate and on the site-specific environmental conditions of the area of plant discharge. Extensive waste discharge modeling studies are often necessary to ascertain the environmental viability of large ocean outfalls and beach well discharge wells. Usually, the completion of these studies is a multi-year effort and involves significant cost expenditures, expert reviews and a multi-step evaluation process. Environmental permitting costs and efforts also depend on the experience of the regulatory agencies with permitting similar seawater desalination projects and the advancement of the regulatory law addressing concentrate discharge permitting and monitoring. Because of the significant differences in desalination plant discharge permitting experience in various countries, the cost of environmental permitting may vary significantly from project to project and from country to country. Overall, the costs associated with project permitting may vary from US\$50 and US\$500/m^3/d (US\$0.19–1.9/gpd).

Legal services. Costs for legal services include the expenditures associated with legal review and processing of environmental permits, and with the preparation and negotiation of contracts for water supply, engineering, operations and construction services. In addition, these expenditures encompass costs for review and processing of contractual agreements for acquisition of land for the desalination plant site; obtaining easements for source water and product water pipelines and electrical supply lines to and from the site; for negotiation of power supply contracts; and any for preparation of other contracts for services, equipment and goods needed for construction and operation of the desalination plant. The cost of legal services is directly related to the complexity of the project and usually varies between US$20 and US$80/m^3/d (US$0.08–US$0.30/gpd).

18.5.1.4. Project financing costs

Project financing costs include expenditures for obtaining all funds and insurance needed for project implementation, from its conception and development through construction, startup and commissioning. The most common methods of financing desalination projects are:

- *Government financing*—where the public sector or the local or state government directly lends funds or provides grants, subsidies or guarantees for repayment of the funds required to build the desalination plant
- *Conventional (bond or construction loan) financing*—where long-term funds are raised by issuing bonds or by providing a long-term construction loan by a private lender to public agency, private utility or business enterprise against a proven credit risk rating and/or ongoing revenues from water sales or other assets
- *Private project financing*—where private lenders fund the project via a special project company, and rely only on future cash flow from the project for repayment of their investment

Government financing of an entire seawater desalination project is not very common today and is usually implemented for construction of small projects and under emergency conditions. However, in many countries such as the USA, Israel, Spain, and many Caribbean and Middle Eastern states, the government directly or indirectly subsidies costs associated with seawater desalination in order to close the "gap" between the cost of water of the traditionally available surface and/or groundwater sources and the cost of desalinated water. Often, the state government provides sovereign guarantees for payment for water supply services under a BOOT contract with private company, especially in circumstances where the direct purchaser of desalinated water is a public agency under the fiscal and administrative control of the government. Sovereign government guarantee is critical for privately financed projects when the contracting public agency does not have fiscal autonomy and/or is not credit risk-rated.

Conventional (bond or construction loan) financing is based on issuing long-term debt in the form of general obligation or revenue bonds or a commercial bank loan for a given project. General obligation bonds are used for financing of publicly owned projects and are secured by the full faith and credit of the issuing entity. In order to issue this type of bonds the entity seeking funding (government, public utility, municipality, etc.) has to have taxing powers to support payments of debt obligations. The key advantage of the general obligation bonds is that they are backed by the full taxing capacity of the governmental entity/public agency, and consequently this credit is considered the strongest security pledge available to a lender, and therefore comes at the lowest available net interest rate. In addition, issuance of general obligation bonds is usually simpler and frequently less costly than raising other types of debt. However, the use of general obligation bonds for funding of desalination projects has a number of constrains. In order to issue such bonds, the legislation of most countries requires prior legislative or voter approval of the bond issue and limits the amount of tax-supported debt that can be issued by a legal administrative entity (utility, municipality, authority, etc.). As a result, financing large seawater desalination projects with general obligation bonds may reduce government agency's ability to issue debt for future projects and may have a negative impact of the issuer's credit rating. This type of bonds cannot be issued by private entities/businesses. The interest rates for general obligation bonds typically vary from 2.5% to 4%.

The second option for conventional project financing is the use of public or private activity revenue bonds. The interest and principal of the long term debt raised through revenue bonds are payable solely through the revenue generated from the specific utility and/or the specific project. Revenue bonds are generally tax-exempt and are typically issued at interest rates lower than taxable debt/bonds and construction loans but higher than general obligation bonds. Typically, tax-exempt revenue bonds have interest rates of 3.5–6%. Taxable debt/bonds usually have interest rate of 4.5–8%.

Since debt service on revenue bonds and commercial loans is typically secured by the revenue stream generated by a particular project, reduction or discontinuance of this revenue could result in a default on these bonds or loans. In order to protect against default, lenders issuing revenue bonds or commercial construction loans require the establishment of several reserve funds that provide security to the investors that adequate amount of funds is available for repayment of the debt, as well as for normal plant operations and for ongoing capital improvements. The typical reserve funds required to be included in project capital costs when revenue bonds or commercial construction loans are issued are:

- *Interest during construction*—fund established for payment of debt service obligations during the period of construction. Usually, during the construction phase of the project the owner pays interest only on the money that are used for construction rather than on the entire principal of the bond issue/loan. Therefore, this interest is often referred to as interest during construction. Since the project does not generate revenue during the construction period, the money required to pay interest and the construction-related portion of the principal on the loan/bond during construction usually have to be borrowed as well, and therefore they become a part of the project capital costs.

- *Debt service reserve fund*—This reserve account is intended to protect the lenders against inability to repay debt where the project revenue is insufficient during the operations period. Similar to the interest during construction, this reserve fund has to be borrowed as part of the loan/bond issue and also is a capital cost item.
- *Working capital (operating fund)*—typically 15–20% of the annual O&M costs in size. This fund provides routine working capital during project construction, start up and commissioning, as well as during normal plant operations. The working capital fund is usually included as a portion of the bond value (i.e., capitalized) when new seawater desalination projects are constructed.
- *Insurance reserve*—funds reserved for self-insurance or for supplementing existing insurance coverage for items not covered by traditional insurance policies of the owner or contractors involved in desalination project implementation.

All reserve funds described above are typically funded as part of the commercial construction loan/bond proceeds and therefore, have to be accounted for in the plant capital costs. As a result, the reserve fund requirements can cause the loan/bond size to increase with 5–10% or more.

Bonds are typically used to finance medium and large size projects (i.e., projects of 20,000 m^3/d (5.3 mgd) or higher). Smaller projects are often financed by construction loans issued by commercial banks/lenders specialized in such financing. Fixed-rate commercial loans are widely used for this purpose and these loans have constant interest rate and payment for the full term of the loan. The term of such loans depends on the project size and risk profile, and typically is between 5 and 20 years. Interest rate for commercial loans is usually set at a spread ranging from 150 to 275 points (i.e., 1.5–2.75%) over internationally accepted and established inter-bank interest rates such as the London inter-bank offered rate (LIBOR). LIBOR is a rate that most creditworthy international banks charge each other for large loans.

Private project financing is widely used in the development and implementation of large BOOT seawater desalination projects. Under this method of financing the source of funds are private lenders–most often the BOOT project developer, private banks, and institutional investors, such as pension and insurance funds. The private project financing is usually a non-recourse financing. In this type of financing, the purchaser and consumer of desalinated water (the public or private water supply entity and its customers) does not have any direct liability for repayment of the funds used for project development and construction and therefore, does not need to pledge any of is assets for fulfillment of the project funding related obligations. The water user only pays for water services. The sole source of repayment of the funds invested in the project is the revenue generated from the sale of desalinated water. Responsibility for repayment of the funds of the privately financed project lies within the special project company established by the private BOOT contractor the assets of which are owned by the project investors providing equity for the project.

Privately financed projects are usually funded by a combination of debt and equity. In some cases, funding can be obtained from Multilateral Lending Agencies (i.e., the World Bank, the Asian Development Bank, the European Bank for Reconstruction and Development) or national "export promoting" agencies. Debt may be in the form of bonds, commercial construction loans, and/or other financial instruments with a long-term or short-term repayment period. Equity is typically provided at the request and according to the conditions of the financial institution issuing the project debt and is usually in a range of 10 to 50% of the total project capital costs.

Commercial banks, financial corporations and project finance funds are typical sources of debt for seawater desalination projects. Equity for the project is usually provided by the BOOT contractor and/or outside equity fund (i.e., private equity fund, insurance or pension fund). If the BOOT project is properly structured and priced, the BOOT contractor's equity could be either direct cash payment and/or indirect contribution of the funds the BOOT contractor actually expends for project development ("sweat equity"). For example, if the capital cost for a given project is US$20 million and the BOOT contractor's gross profit in the project is 5%, or US$1 million, this 5% can be "invested" as a portion of the equity contribution required for the project. If the lender of debt to the project requires a minimum of 10% of equity contribution from other sources and is willing to lend debt for 90% of the project capital cost, than the BOOT contractor can use its 5% of "sweat equity" against the 10% equity requirement and therefore, has to rise the remaining 5% of the required equity from outside investors.

Revenue-based (non-recourse) project financing typically is more complicated and costly to structure than an asset-based debt. Transaction costs normally include financial advisory fees, bank fees, legal fees, and independent engineering fees. As a result, private non-recourse financing may not be practical or cost-competitive for smaller desalination projects (projects with capital costs of less than US$10 million), unless the transaction costs can be streamlined or multiple projects can be combined in one financial package.

When the project is operational, the revenue generated from the desalinated water sales is used to: (1) pay for plant's O&M expenditures; (2) repay debt obligations; and (3) pay return on equity investment. The payment seniority usually follows the sequence described above. The operational expenses get paid first followed by repayment of the debt, and followed by payment of return to the equity investors. Because the equity investors get paid last, after all other project-related payment obligations are met, and because the plant revenue is the only source of repayment or all project fiscal obligations, the equity investors are exposed to highest risk of realizing their return-of-investment goals. Typically, the debt investors are protected by a "take-or-pay" clause of the water sales agreement between the BOOT contractor and the entity purchasing the desalinated water. The equity investors usually do not have such protection of their investment and therefore, their return-on-investment expectations are higher than these of the debt lenders. In general, equity investors have expectations of returns commensurate with the returns yielded by financial stock markets trading securities of comparable risk profile.

Annual interest at a preset rate is charged for the use of the funds which lenders provide under any of the forms of project financing described above. For a given public utility, the cost of funds required to finance desalination project would depend mainly on the credit rating of this utility and on the restrictions that apply to the public utility in relation to assuming new debt obligations. Public utilities with relatively low credit rating and/or limited capacity to borrow adequate amount of funds/issue bonds may often be able to obtain a more favorable financing terms by using private sources of financing. In addition to lowering the overall cost of project funding and the project risk profile, involvement of the private sector in the project financing also has the benefits of keeping such financing off the balance sheet of the public utility which is embarking on a desalination project and of sharing project implementation, and performance risks, and costs with the private sector.

Interest during construction. Debt/bond obligations are typically repaid using the revenues from the sale of the desalinated water to the consumers of this water. However, during the period of time when project is under construction no revenue is available to repay debt obligations. Therefore, typically the owner of the project borrows additional funds for the payment of the interest on the money used for construction.

Typically, interest during construction is calculated by multiplying the construction cost of the project by the annual interest rate of the loan and by 50% of the length of the construction period in years. This estimate assumes that 50% of the loan on average will be outstanding. Depending on the type of financing used for funding of the desalination project, interest during construction is usually between 0.5 and 4.5% of the total capital costs.

Debt service reserve. As indicated previously, the debt service reserve account is intended to protect the lenders against inability to repay debt where the project revenue is insufficient. Depending on the type of financing, the complexity of the project and the revenues of the water sales as compared to the debt obligations, the debt service reserve is typically set as the least of the following three values: (1) maximum annual debt service; (2) 125% of the average debt service, or (3) 10% of the principal. The debt service reserve typically ranges between 2.0% and 8.5% to the project capital costs.

Other financing costs. Other project financing costs comprise of expenditures associated with the funding of other reserve funds in addition to the debt service reserve fund, if needed to satisfy lender requirements; administrative and legal costs related to issuing project bonds or arranging project loans and administering payments; and costs associated with arranging project equity, if equity contributions are used for project financing. Other financing costs also include expenditures associated with purchasing insurance and obtaining performance and payment bonds to protect the owner and contractors against construction failures and problems; and payment of various taxes associated with project implementation as well as shipping costs for delivering plant components to the site. Theses costs range between 0.5% and 4.5% of the total capital costs.

Contingency. Contingency provisions in the project cost estimate reflect the fact that even when a detailed cost estimate is completed, there are a number of unknown factors that may influence the actual expenditures associated with project implementation. As indicated previously, the level of contingency used for a given cost estimate depends on the level of accuracy of this estimate. The various levels of accuracy of the project cost estimate and the associated contingencies are discussed in detail in Section 1.3 of this chapter. A detailed cost estimate usually carries a contingency factor of 5–10% depending on the complexity and size of the project. Higher contingency factors are used for lower-level accuracy cost estimates.

18.5.2. Operation and maintenance costs

As indicated previously, the plant O&M costs incorporate all expenditures associated with plant operations for the entire useful life of the desalination plant. The key O&M cost elements are energy (power), chemicals, labor, membrane replacement and maintenance. In total these costs typically encompass over 80% of the annual O&M expenditures. Table 18.10 presents a cost breakdown for the key annual O&M cost elements.

Power. The power costs for the desalination plant are dependent on two key components: the power tariff and associated unit cost of power, usually expressed in monetary units per kWh, and the amount of power used to produce desalinated water, typically presented in kWh per m^3 or 1000 gallons of product water.

The power costs are directly related to the source water salinity and associated osmotic pressure that has to be overcome to produce fresh water. The SWRO system usually uses over 85% of the power required to operate the de-

TABLE 18.10
Annual O&M cost breakdown

	Percentage of total O&M cost (%)	
Cost item	Low-complexity project	High-complexity project
Variable O&M costs		
1. Power	45.0–61.0	35.0–58.0
2. Chemicals	3.0–6.5	5.5–9.0
3. Replacement of membranes and cartridge filters	5.0–9.0	6.5–11.0
4. Waste stream disposal	2.5–5.5	3.5–7.0
Subtotal—variable O&M costs	55.5–82.0	50.5–85.0
Fixed O&M costs		
1. Labor	5.0–9.5	4.0–11.0
2. Maintenance	6.5–12.5	3.0–13.0
3. Environmental and performance monitoring	0.5–4.0	1.0–5.0
4. Indirect O&M costs	7.5–18.5	7.0–20.5
Subtotal—fixed O&M costs	19.5–44.5	15.0–49.5
Total O&M costs	100%	100%

salination plant. The rest of the power is consumed mainly by the plant's intake and pretreatment systems and product water pumps. An example detailed breakdown of the power use of various facilities in a 200,000 m^3/d (53 mgd) seawater desalination plant treating seawater of TDS concentration of 33,500 mg/L and average annual temperature of 23°C is presented in Table 18.11. This example includes the use of energy recovery turbine (Pelton wheel) for recovery of a portion of the energy in the desalination plant concentrate. The efficiency of energy recovery varies with the type of the equipment used (Pelton wheel, turbocharger, reverse running pump or pressure exchanger) and with the overall recovery and configuration of the SWRO system.

When electricity is purchased from an independent power generation supplier, the cost-of-power tariff is typically outside of the control of the owner of the desalination plant. In this case, the desalination plant could be designed to take advantage of the cost reduction associated with the off-peak power rate, which usually is lower than this rate during the peak hours of power consumption. Usually, the peak power rate timeframe coincides with the periods of peak of water demand, during which the desalination plant often has to operate at maximum rather than minimum capacity. Therefore, provision of adequate amount of product water storage would be essential to take advantage of the benefits of maximum off-peak power tariff operation of the desalination plant. Construction of additional plant product water storage capacity to accommodate off-peak power tariff benefits would increase plant construction costs and therefore, its viability has to be accessed on a life-cycle cost basis.

Some power generation utilities provide additional power tariff incentives if the desalination plant owner is willing to significantly curtail or completely discontinue plant operations during periods of the year when the power generation utility can sell this power at very high prices to other users. Power curtailment conditions, if offered by the electrical company, would vary from one power supplier to another, but in general would involve a requirement for reduction of over 90% of the desalination plant power use for a period of 6–12 hours for at least two times per month. In order to accommodate such curtailment schedule, the desalination plant design, operations and water supply delivery commitments have to have built-in flexibility and extra product water storage capacity, which usually come with an increased capital expense.

Another potential alternative for reduction of the unit power rate is to co-locate the desalination plant with an existing power plant and to connect the desalination plant's electrical system directly to the power plant generation units thereby completely avoiding the use of the power grid for electrical supply.

TABLE 18.11
Example average and maximum power use breakdown for 200,000 m^3/d seawater desalination plant.

Desalination plant facility	No. of units (duty/standby)	Unit size (Hp)	Average power use Total Hp	Average power use (kWh/m^3)
Intake pump station	3/1	750	2,138	0.191
Pretreatment filters				
– Filter backwash blowers	2/1	150	150	0.013
– Filter backwash pumps	2/1	180	180	0.016
Reverse osmosis system				
– Filter effluent transfer pumps	12/1	350	3,990	0.357
– High pressure RO feed pumps	12/1	3,500	37,800	3.383
– Energy recovery turbine (reduction)	12/1	–875	(9,450)	(0.845)
Product water pump station	4/1	550	1,980	0.177
Membrane cleaning system	—	—	299	0.027
Solids handling equipment	—	—	388	0.035
Chemical feed equipment	—	—	985	0.088
Service facilities	—	—	1,450	0.130
Total desalination plant power use			39,910	3.572

Often, the power tariff consists of two components–a power generation and a power grid distribution charge component. Depending on the exiting government regulations governing power generation, supply and distribution, the direct connection to the power plant's generation units may allow to avoid the payment of the power grid component of the tariff. Since this component may be as large as half of the power rate, the co-location approach could allow a substantial reduction in the unit power costs and therefore, of the total costs for seawater desalination.

Another alternative to avoid power grid associated charges and to reduce the unit cost of power is to self-generate electricity at the desalination plant site. This approach is usually viable for very large plants (example is the Ashkelon seawater desalination facility), because the generation of small quantities of electricity is typically not as cost effective as power generation on a large commercial scale by an experienced power generation company. Power self-generation may be cost-effective for small-size desalination plants in cases when there is no easy access to a nearby electrical power grid and/or when the commercially available power rate is very high and self generation of electricity is cost-competitive. Another important issue associated with power self generation is the risk the desalination plant owner and investors take with the increase in the unit cost of fuel (usually natural gas) used for power generation over time and the sustained availability of a particular type of fuel over the useful life of the desalination project. Taking these risks is usually prudent only if they can be shared with the water consumer, mitigated by the government or taken by a major supplier of this fuel product via long-term fuel supply contract that expands over the useful life of the project.

The use of power for production of desalinated water can be reduced by a number of ways. One widely used approach is to decrease the total energy use of the SWRO desalination system by the implementation of pressure-exchanger (isobaric chamber) type of energy recovery system. This technology is described in detail in the previous chapters.

An alternative approach for reduction of the overall power demand for seawater desalination is the use of warmer source seawater (i.e., cooling water from a power plant discharge). As indicated in Section 18.2.1 of this Chapter, the use of 5–10°C warmer seawater results in similar reduction in the RO feed pressure required for desalination and therefore, in lower overall power demand.

Seawater desalination power costs could be decreased by reduction of the source water salinity as well. The source water salinity could be lowered by blending of the plant source seawater with brackish water from intake wells or with concentrate (brine) from existing brackish water desalination plant (desalter). The second option, i.e., the use of concentrate from brackish water desalter as feed water to a seawater desalination plant, is mutually beneficial for both the desalter and the seawater plant. Usually, in-land desalter capacity is limited by the lack of suitable discharge location of the plant concentrate. If the seawater desalination plant can accept the desalter concentrate and process it, the brackish desalter capacity could be increased beyond the threshold driven by brine discharge limitations, and the desalination plant source salinity could be reduced at the same time. While the seawater TDS concentration is usually in a range of 30 to 40 ppt, the TDS of the concentrate from brackish water desalters is typically several times lower (i.e., typically, 2–15 ppt). Therefore, when blended with the source seawater it would reduce the overall desalination plant feed water salinity, which in turn would have a positive effect on the overall desalination plant power use, recovery factor and cost of water.

Cost of power is a variable annual expenditure and usually ranges between US\$0.13/m^3 and US\$0.25/m^3. The cost variation may be wider for site specific conditions where power supply is difficult or power-self generation is applied.

Chemicals. Chemical costs these costs are highly variable from one location to another and are mainly dependent on the source water quality, the selected pretreatment processes and the target product water quality. Table 18.12 presents unit costs for various chemicals frequently used in seawater desalination plants. The actual chemical cost values for a given project have to be established based on quotes from local suppliers of the site-specific chemicals.

Cost of chemicals is a variable expenditure and typically is in a range of US\$0.015/m^3 to US\$0.045/m^3 (\$0.06–0.17 kgal) of product water.

Labor. Plant operation labor costs are closely related to plant size, complexity and number of treatment processes, and equipment, and to the overall level of plant automation. Typically, desalination plants are highly automated and reliable facilities which use limited amount of specialized staff for overall plant performance monitoring and control, equipment maintenance, preparation of chemical batches for various treatment processes and collection and analysis

TABLE 18.12
Unit chemical costs

Chemical	Unit cost, US\$/kg (\$/lb)
Chlorine gas	0.4–0.8 (0.2–0.4)
Sodium hypochlorite	2.0–2.5 (0.9–1.1)
Ferric sulfate and ferric chloride	0.3–0.8 (0.1–0.4)
Sulfuric acid (93% H_2SO_4)	0.1–0.2 (0.05 – 0.10)
Citric acid	1.5–2.0 (0.7–1.4)
Biocide	2.5–4.5 (1.1–2.0)
Sodium hydroxide (50% NaOH)	0.5–0.9 (0.2–0.4)
Sodium bisulfite	0.3–0.4 (0.15–0.20)
Antiscalant (scale-inhibitor)	1.5–4.0 (0.7–1.8)
Ammonium hydroxide	0.5–0.8 (0.2–0.4)
Lime	0.1–0.3 (0.05–0.15)
Sodium tripolyphosphate (corrosion inhibitor)	1.5–3.0 (0.7–1.4)
Other cleaning chemicals (US\$/m^3 of permeate)	0.004–0.006 (0.002–0.003)

Note: All costs in 2005 US\$.

of water quality samples. Usually, every desalination plant has a plant manager, shift supervisors, operators, one or more mechanics and electricians and laboratory and administrative staff. Several smaller facilities are often supervised by one regional plant manager and serviced by a central laboratory, and instrumentation and control group.

Table 18.13 summarizes the typical plant staffing requirements for a seawater desalination plant as a function of the level of plant automation, treatment process complexity and labor skills. As seen in this table, the number of plant staff varies with plant capacity and is strongly influenced by economy of scale.

Usually, desalination plant staff is organized in one to three shifts and in some smaller and fully automated plants plant operations are unmanned at night. Large plants are typically staffed 24-hours per day, 365 days per year, with at least two operators on duty at all times. The labor costs are fixed for a given plant and are typically in a range of US\$0.015–0.035/m^3 (US\$0.057–0.132/kgal) of treated water.

Maintenance. Maintenance costs are one of the larger cost elements of the annual O&M expenditures. This cost item includes all expenditures associated with the routine operations, and preventive and emergency maintenance of all plant equipment and piping. Typically, the useful life of most of the key desalination plant equipment is between 25 and 30 years. Therefore, the average annual maintenance expenditure is approximately 0.33% (100%/30 years) to 0.40% (100%/25 years) of the cost of the installed equipment. The maintenance costs vary from year to year because the key high-cost desalination equipment such as high-pressure pumps and other large capacity pumps undergo routine

TABLE 18.13
Seawater desalination plant staffing requirements

Plant capacity, m^3/d (mgd)	Plant automation, complexity and labor skill level	
	High	Low
1,000 (0.26)	2–3	4–6
5,000 (1.32)	4–6	8–10
10,000 (2.6)	7–10	12–14
20,000 (5.3)	9–12	16–18
40,000 (10.6)	12–16	18–20
100,000 (26.4)	14–18	20–24
200,000 (52.8)	18–28	30–40

equipment rebuilt every 5 to 10 years in order to maintain their high efficiency and consistent performance.

Since most of the plant equipment is maintained routinely on a preset maintenance schedule independent of the actual water production, some (typically 40–60%) of the routine equipment maintenance costs, are considered and often budgeted as a fixed O&M component. The remaining portion of the maintenance costs is accounted as a variable component and is related to the actual equipment run time. Plant total maintenance costs are typically in a range of US\$0.025–0.065/m^3 (US\$0.094–0.245/kgal) of desalinated water.

Membrane and cartridge filter replacement. These costs incorporate expenditures for replacement of pretreatment and RO membranes as well as cartridge filters. These costs are closely related to the replacement frequency of these filters, which in turn depends on the source water quality and plant design. The typical useful life of pretreatment membranes is 3–5 years. Therefore, their annual replacement costs range between 20% and 33.3%. The seawater membrane useful life is between 5 and 7 years, and in the case of high quality source water may extend up to 10 years. As a result, the typical annual average SWRO membrane replacement rate is 14.3–20%. Both pretreatment and SWRO membranes are replaced when the membrane media fouls irreversibly to levels that require excessive power use for their operation. Membranes as also replaced when they loose their integrity and their performance declines irreversibly. The unit costs for various SWRO elements are shown in Table 18.7.

Cartridge filters for SWRO plants have a typical minimum useful life of 6–8 weeks. However, in many applications where the source water is of high quality, cartridge filter replacement is less frequent (once every 6 to 12 months). Depending on the cartridge filter size, the unit cartridge filter cost is between US\$10 and US\$30 per filter. The total membrane and cartridge filter replacement costs are typically in a range of US\$0.020–0.060/m^3 (US\$0.075–0.226/kgal).

18.5.3. Waste stream disposal

The main waste stream of every membrane seawater desalination plant is the RO system concentrate. Depending on the concentrate disposal practices described in detail in Section 18.5.1 of this chapter, the total waste stream disposal costs are typically in a range between US\$0.01/m^3 to US\$0.03/m^3 (US\$0.038–0.113/kgal). In most applications, the ocean discharge of concentrate form desalination plants with open ocean intake is acceptable without any additional treatment and minimal or no costs. For open-ocean discharges from seawater desalination plants with beach well intakes, the concentrate disposal costs are dependent on the need to aerate the concentrate before its discharge to the ocean or to otherwise treat it if the concentrate exhibits toxicity or has other measurable environmental impacts. Concentrate disposal costs include expenditures associated with operation and maintenance of concentrate injection wells, evaporation ponds, or mechanical evaporation equipment if such disposal methods are used. These costs may very widely depending on the disposal method and project size.

For sanitary sewer discharge of the concentrate while the conveyance costs are mainly driven by the volume of the concentrate, the sewer connection fees can vary significantly for a given location from none to several orders of magnitude larger than the conveyance costs. The sewer connection fees usually are related to the available capacity of the sewer facilities and the effect of the concentrate discharge on the operational costs of the wastewater treatment plant which would provide ultimate treatment and disposal of the concentrate.

Seawater desalination plants generate a number of other waste streams in addition to the plant concentrate. The main waste streams are the pretreatment waste filter backwash and the spent RO membrane cleaning solution. In the case of membrane seawater pretreatment, the desalination plant generates two additional waste streams: chemically enhanced backwash (CEB) and spend pretreatment membrane cleaning solution.

The cost of waste disposal depends on the method of waste disposal planned to be used at the desalination plant and the size of the waste streams. In many applications worldwide all waste streams are returned to the ocean for disposal. Therefore, under best-case scenario no additional costs for waste disposal are incurred. Frequently, the waste filter backwash along with the plant concentrate are the only two desalination plant process streams allowed to be discharged to the ocean, and the rest of the waste streams have to be conveyed for the sanitary sewer for disposal and further treatment. In this case, the expense of waste stream disposal is usually the sewer discharge fee established by the local wastewater collection and treatment agency. This cost may be between US\$0.005/m^3 to US\$0.010/m^3 (US\$0.019–0.038/kgal).

In some large seawater desalination plants the spent filter backwash water has to be treated (typically by sedimentation) before discharge to the ocean. The solid residuals generated during the filter backwash treatment are usually dewatered to solids content of 20% or higher via mechanical dewatering equipment (belt filter presses, centrifuges or plate-and-frame presses) and disposed off to a sanitary landfill. Depending on the capacity and distance of the available landfills in the area of the desalination plant, the residuals disposal costs could reach up to US\$0.010/m^3 (US\$0.038/kgal).

18.5.4. Environmental and performance monitoring

Every desalination plant has discharge water quality monitoring requirements. These requirements may be applicable to the entire discharge and/or to the individual plant waste streams. In addition, in many environmentally sensitive areas the monitoring requirements encompass not only the discharge but the receiving water body (ocean, or groundwater aquifer or estuary) as well. Depending on the complexity and frequency of the environmental monitoring required for permit compliance, the discharge monitoring costs could be substantial and have to be taken under consideration in determining the overall plant O&M costs. Plant discharge monitoring costs may vary between US\$0.001/m^3 to US\$0.010/m^3 (US\$0.004–0.038/kgal).

Plant performance monitoring costs are expenses needed to measure and analyze key process performance parameters (i.e., SDI, temperature, pH, salinity of plant feed water, etc.). These O&M costs depend on the level of automation and plant complexity. Product water monitoring costs are expenditures associated with sample collection, laboratory analysis, and data management and reporting, which are required to be completed in order to comply with all applicable regulatory requirements associated with the product water supply. Typically, plant performance and product water quality monitoring costs are between US\$0.002/m^3 to US\$0.020/m^3 (US\$0.007–0.075/kgal).

18.5.5. Indirect O&M costs

Indirect O&M costs include annual expenditures for staff training, professional development and certification; expenditures for consumables and maintenance of plant service vehicles; administrative and utility/service (water, sewer, telephone, etc.) expenses; taxes associated with plant operations; operations insurance, contingency and other O&M reserve funds. These costs also incorporate the fees for plant operation, if the desalination plant is operated by a private operations company. Typically plant indirect O&M costs vary in a range of 7.0–20.5% of the total O&M costs (US\$0.03/m^3 to US\$0.09/m^3, US\$0.113 to US\$0.340/kgal).

18.6. Cost of water

The cost of water production encompasses all expenditures associated with project implementation and operation and consists of fixed and variable components. The fixed water costs are all expenditures associated with plant construction and with repayment of the capital investment in the plant (i.e., capital cost recovery); and of the portion of the annual O&M expenditures that are independent of the actual volume of water produced by the desalination plant (labor, maintenance, environmental and performance monitoring, and indirect O&M costs). The variable cost of water component incorporates O&M expenditures that are directly related and usually proportional to the actual volume of the produced desalinated water (power, chemicals, replacement of membranes and cartridge filters, and waste stream disposal).

When the desalination plant is delivered under a BOOT contract between a public agency and a private contractor, the water tariff structure under which the water utility purchases water is typically reflective of the cost of water structure described above. The tariff usually includes a capacity payment component which compensates the private contractor for the fixed cost associated with water production, and a commodity (output) tariff payment component, which provides compensation for contractor's variable O&M expenditures. Example capital, O&M and cost of water estimates for a fictional 40,000 m^3/d (10.6 mgd) seawater desalination project are provided in Appendix B.

18.6.1. Fixed components of water cost

Capital cost recovery. The capital cost for construction of seawater desalination plant is usually amortized over the term of repayment of the capital used to build the desalination plant (typically a period of 5–30 years). To determine the amortized value of the capital costs, these costs are divided by a capital recovery factor (CRF) and by plant's design capacity availability factor. The CRF is a function of the interest rate of the capital and the number of years over which the investment is recovered (i.e., the plant capital expenditures are repaid). The CRF can be calculated using the following relationship:

$$\mathrm{CRF} = \{(1+i)^n - 1\}/\{i(1+i)^n\}$$

where: n is the period of repayment of capital expenditures; i is the interest rate of the amortized investment.

For example, the CRF for a 100,000 m^3/d (26.4 mgd) desalination project that has capital costs of US$100 million, repayment period of 20 years, and amortization rate of 5%, is 12.462. Therefore, the project's annual amortized (annualized) capital cost is US$100 MM/12.462 = US$8,802,440/y. The capital cost recovery portion of the cost of water for this example project is: US$8,802,440/y/(100,000 m^3/d× 365 d) = US$0.24/$m^3$ (US$0.91/kgal). If for example, the plant has a design capacity availability factor of 95%, than the capital cost recovery charge will be increased accordingly to US$0.24/$m^3$/(95%) = US$0.25/m^3 (US$0.94/kgal).

In many projects, the capital investment is a combination of equity and debt, which have different interest rates of return on investment. In addition, the interest rates may vary over the repayment period. As a result the calculation of the capital cost recovery for such project may not be as straightforward as shown above, and typically requires the development of financial model that reflects all specific features of the various investments used for the project. Development of financial model for the project is usually a responsibility of the project developer/owner. If the project developer does not have adequate in-house capabilities to develop a financial model of level of sophistication needed for obtaining competitive financing, typically the developer/owner retains specialized company to provide the necessary expertise.

18.6.2. Other fixed costs

As indicated previously, the other fixed components of the cost of water are:

- Labor costs
- Maintenance costs
- Plant environmental and performance monitoring costs
- Indirect O&M costs

These costs are typically calculated by dividing the annual fixed O&M expenditures by the design average annual production capacity of the desalination plant and by the plant design capacity availability factor. For example, if for the 100,000 m^3/d (26 mgd) plant referenced above, the annual labor costs are US$700,000/y; the maintenance costs are US$1,100,000/y; the plant environmental and performance monitoring costs are US$50,000/y; and the indirect O&M costs are $1,500,000/y, than the other fixed water costs are estimated at: US$3,350,000/y/ (100,000 m^3/d× 365 d × 95%) = $0.10/$m^3$ (US$0.38/kgal). As a result, the total fixed water costs for this example are: $\$0.25/m^3$ + $\$0.10/m^3$ = US$\$0.35/m^3$ (US$1.32/kgal).

The fixed cost of water component is independent of the actual amount of water that is produced by the desalination plant. Therefore, these costs have to be minimized as much as possible. Labor costs are typically reduced by using high level of automation. Maintenance costs are minimized by selecting high-quality materials, equipment and piping, and implementing proactive and systematic preventive maintenance program. Plant environmental and monitoring costs are reduced by using environmentally safe, low-cost concentrate disposal methods and by automation of most plant performance monitoring functions. Indirect O&M costs are typically reduced by using highly qualified operations staff or subcontracting plant operations to a private contract operation company specialized in seawater desalination. Since the reduction of the other fixed costs requires higher capital expenditures and therefore, increases the capital recovery costs, the total fixed costs have to be optimized to find the right balance between the key two fixed cost elements.

18.6.3. Variable components of water cost

A variable component of water cost typically includes the following O&M expenditures:

- Power
- Chemicals
- Replacement of membranes and cartridges
- Waste stream disposal

Power expenditure is the largest variable cost component and usually accounts for 20–35% of the total cost of water. Depending of the power tariff structure, the fixed portion of the power costs, such as the electrical grid connection charges, may sometimes be accounted for as a portion of the fixed cost of water component. On the other hand, some of the maintenance costs, which traditionally are considered fixed costs, may be accounted for as variable costs. This holds true especially for equipment which routine maintenance/replacement schedule is based on the actual number of operating hours.

As indicated previously, chemical costs are related not only to the desalination plant source water and production flows, but to the source water quality as well. Usually, treatment of source water of good quality (low SDI, turbidity and organic content) requires lower amount of pretreatment chemicals and less frequent membrane cleaning, which in turn reduces plant chemical costs as well. The difference between the chemical pretreatment and membrane cleaning costs for good and worst-than-average source water quality could be significant–often two to four times lower. This difference, however, has to be put in prospective. Since the chemical costs are usually less than 8% of the total water production costs, a two-fold chemical cost reduction due to improved source water quality may not amount to a very large overall cost of water reduction.

Where, the source water quality makes a measurable cost difference however, is the frequency and extent of SWRO membrane fouling and the associated increase in plant downtime. If the source water quality is poor and it requires very frequent membrane cleaning and replacement due to fouling, the excessive membrane maintenance needs typically result in plant production interruptions and ultimately in reduced overall plant capacity availability factor. In addition, accelerated membrane fouling increases the average plant power use.

Waste stream disposal costs usually are relatively small. However, in some cases operations of the concentrate disposal facilities could constitute a significant portion of the plant water production costs and malfunctioning of these facilities could reduce significantly plant capacity factor (i.e., increase downtime). Therefore, the use of simple and environmentally sound methods of concentrate disposal such as co-discharge with power plant cooling water, sanitary sewer discharge or direct open ocean discharge, when viable, is recommended over deep well discharge, evaporation pond disposal or zero-discharge.

The variable water costs are typically calculated by dividing the annual variable O&M costs listed above by the actual average annual production capacity of the desalination plant. For the purposes of budgetary cost estimates and determination of the water tariff, of new desalination projects, the variable water costs are calculated by dividing the projected annual variable O&M expenditures by the design average annual plant water production flow and availability factor. For example, if a 100,000 m^3/d (26 mgd) seawater desalination plant has an actual availability factor of 95%; the actual annual plant power costs are US$5,700,000/y; the chemical costs are US$800,000/y; the annual costs for membrane and cartridge filter replacement are US$850,000; and the waste stream disposal costs are US$300,000/y, than the variable water costs are estimated at US$7,650,000/y/(100,000 m^3/d × 365 d × 95%) = US$0.22/$m^3$ (US$0.83/kgal). As a result, the total (fixed and variable) cost for production of desalinated water for this example will be: US$0.35/$m^3$ + US$0.22/$m^3$ = US$0.57/$m^3$ (US$2.15/kgal).

18.7. Trends of water cost

Historically, one of the key obstacles limiting the use of seawater desalination in a large scale has been the high cost of water production. A number of cost-saving innovations in seawater desalination technology over the last ten years are transforming this once costly option of last resort into a viable water supply alternative.

The "engine" of every desalination plant that turns seawater into fresh potable water is the RO membrane element. The most widely used type of RO membrane elements consist of two membrane sheets glued together and spirally wound around a perforated central tube through which the desalinated water exits the membrane element. A large seawater desalination facility usually has thousands of membrane elements connected into a highly automated and efficient water treatment system, which typically produces 1 cubic meter of fresh water from approximately 2 cubic meters of seawater. The membrane productivity, power use, salt separation efficiency, cost of production and durability of the membrane elements by large determine the cost of the desalinated water. Technological and production improvements in all of these areas in the last two decades are now rendering water supply from the ocean affordable. Membrane productivity–i.e., the amount of water that can be produced by one membrane element, has increased over two times in the last 20 years.

Recent introduction of spiral wound membrane elements with a larger number of membrane "leaves" and denser packing offer increased efficiency as compared to older designs. Today's most efficient elements have more than twice as many membrane leaves compared to older designs. Higher productivity means that the same amount of water can be produced with significantly less membrane elements, which has a profound effect of the size of the membrane equipment, treatment plant buildings, and the footprint of the desalination facility—all of which ultimately reduce the cost of water production.

In seawater desalination facilities salts are separated from the fresh water applying pressure to the seawater, which is 60–70 times higher than the atmospheric pressure. After the salt/water separation is complete, a great portion of this energy stays with the more concentrated seawater and can be recovered, and reused to minimize the overall power cost for seawater desalination. Dramatic improvements of the membrane element materials and energy recovery equipment over the last 20 years coupled with enhancements in the efficiency of RO feed pumps, and reduction of the pressure losses through the membrane elements have allowed to reduce the use of power to desalinate seawater to less than 3.5 kWh/m^3 (13.2 Kwh/kgal) of produced fresh water today. Taking under consideration that the cost of power is typically 20–35% of the total cost of desalinated water, these technological innovations contributed greatly to the reduction of the overall cost of seawater desalination.

Novel energy recovery systems working on the pressure exchange principle (isobaric chambers) are currently available in the market and use of these systems is expected to further reduce the desalination power costs with approximately 10–15%. The pressure exchangers transfer the high pressure of the concentrated seawater directly into the RO feed water with an efficiency exceeding 95%. Future lower-energy RO membrane elements are expected to operate at even lower pressures and to continue to yield further reduction in cost of desalinated water.

Membrane performance tends to naturally deteriorate over time due to combination of material wear-and-tear and irreversible fouling of the membrane elements. Typically membrane elements have to be replaced every five years to maintain their performance in terms of water quality and power demand for salt separation. Improvements of membrane element polymer chemistry and production process over the last 10 years have made the membranes more durable and have extended their useful life. Use of elaborate conventional media pretreatment technologies and ultra and micro-filtration membrane pretreatment systems prior to RO desalination is expected to allow extending the membrane useful life to seven years and beyond, thereby reducing the costs for their replacement and the overall cost of water.

Today, the RO membrane technology and elements are highly standardized in terms of size, productivity, durability and useful life. There are number of manufacturers of high-quality seawater RO membrane elements which provide interchangeable products of excellent quality, proven track record and performance. All of the leading membrane manufacturers are dedicated to supporting the water desalination market and advancing membrane technology, and science at a pace no other water technology can compare with. The desalination facility of today is a highly automated water production factory with a number of build-in protection and safety systems allowing reduction of staffing requirements to a minimum and thereby reducing the costs of plant operation.

The developments in seawater desalination technology during the past two decades, combined with transition to construction of large capacity plants, colocation with power plant generation facilities and enhanced competition by using Build-Own-Operate-Transfer method of project delivery have resulted in

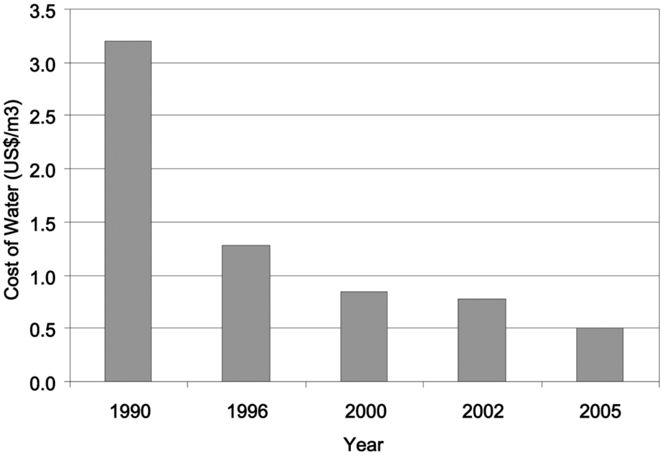

FIG. 18.4 Historic trend of desalinated water cost reduction from large seawater RO plants.

a dramatic decrease of the cost of desalinated water. Fig. 18.4 illustrates the trend of decreasing cost of water produced by seawater desalination based on recent large RO desalination projects in the US, Israel, Cyprus, Singapore and the Middle East. The actual cost values presented in this figure represent specific large desalination projects which had the lowest recorded costs for the particular year. The figure is only an illustration of the overall cost-of-water reduction trend observed over the last 15 years rather than a cost curve suitable for project cost projections.

The advance of the reverse osmosis desalination technology is closest in dynamics to that of the computer technology. While conventional technologies, such as sedimentation and filtration have seen modest advancement since their initial use for potable water treatment several centuries ago, new more efficient seawater desalination membranes and membrane technologies, and equipment improvements are released every several years. Similar to computers, the RO membranes of today are many times smaller, more productive and cheaper than the first working prototypes. The future improvements of the SWRO membrane technology are forecasted to encompass:

- Development of membranes of higher salt and pathogen rejection, and productivity; and reduced trans-membrane pressure, and fouling potential

- Improvement of membrane resistance to oxidants, elevated temperature and compaction
- Extension of membrane useful life beyond 10 years
- Integration of membrane pretreatment, advanced energy recovery and SWRO systems
- Integration of brackish and seawater desalination systems
- Development of new generation of high-efficiency pumps and energy recovery systems for SWRO applications
- Replacement of key stainless steel desalination plant components with plastic components to increase plant longevity and decrease overall cost of water production
- Reduction of membrane element costs by complete automation of the entire production and testing process
- Development of methods for low-cost continuous membrane cleaning which allow to reduce downtime and chemical cleaning costs
- Development for methods for low-cost membrane concentrate treatment, in-plant and off-site reuse, and disposal

These technology advances are expected to ascertain the position of SWRO treatment as viable and cost–competitive processes for potable water production and to reduce the cost of desalinated water by 20% in the next five years and by up to 50% by year 2020.

18.8. Project implementation

18.8.1. Project delivery alternatives

Seawater desalination projects can be implemented using a number of contracting methods, which can be summarized into three key categories: design-bid-build (DBB); design-build-operate (DBO); and build-own-operate-transfer (BOOT). To date, the DBB method has been commonly used for procurement of small and medium size seawater desalination plants in Europe, the US and Israel, and for large-scale desalination projects in the Middle East. Large seawater desalination projects in Europe, Israel, Asia, the Caribbean, and the US are typically implemented using the BOOT method of delivery. Table 18.14 presents a list of the large-scale seawater desalination plants build in the last 10 years along with the method of project delivery for each plant.

TABLE 18.14
Large SWRO plants constructed in the last ten years

Plant name/ location	Capacity, m^3/d (mgd)	In operation since	Project delivery method
Ashkelon, Israel	325,000 (86)	2005	BOOT 25-yr term
Tuas, Singapore	136,000 (36)	2005	BOOT 20-yr term
Cartagena/Mauricia, Spain	65,000 (17)	2004	BOOT 15-yr term
Fujairah, UAE	170,000 (45)	2003	BOOT 25-yr term
Tampa Bay, USA	95,000 (25)	2003	BOOT/DBO
Alikante, Spain	50,000 (13)	2003	Design-Bid-Build
Carboneras—Almeria, Spain	120,000 (32)	2003	BOOT 15-yr term
Point Lisas, Trinidad	110,000 (29)	2002	BOOT 30-yr term
Las Palmas—Telde	35,000 (9)	2002	Design-Bid-Build Private O&M Contractor
Larnaca, Cyprus	54,000 (14)	2001	BOOT 10-yr term
Al Jubail III, Saudi Arabia	91,000 (24)	2000	Design-Bid-Build
Muricia, Spain Private O&M Contractor	65,000 (17)	1999	Design-Bid-Build
The Bay of Palma Palma de Mallorca	63,000 (17)	1998	Design-Bid-Build Private O&M contractor
Dhekelia, Cyprus	40,000 (11)	1997	BOOT 10-yr term
Marbella—Malaga, Spain	55,000 (15)	1997	BOOT 25-yr term
Okinawa, Japan	40,000 (11)	1996	Design-Bid-Build

The type of the selected contracting method mainly depends on the type of owner (public agency or private entity); the project risk profile and owner's experience with similar projects; and the source of project funding–loans, grants, bonds, equity or a mixture of these funding sources. The type of the selected project delivery method often has a significant influence on project costs and therefore deserves considerable attention.

18.8.2. Design-bid-build

Key project parties and their roles. Under this traditional method of project delivery, the desalination plant owner is typically a public entity (municipality or utility), which is responsible for the overall project implementation as well as for the project financing and long-term plant operation and maintenance. Typically under this method of project delivery, the owner retains a consulting engineer to prepare detailed technical specifications for the desalination project, which are used to procure construction contractor or contractors to build the project. The construction contractors complete their work under the supervision of the owner and the consulting engineer and their main responsibility is to implement the requirements indicated in the specifications.

Key advantages and disadvantages. The key advantage of this delivery method for the owner is that the owner retains complete control over the plant ownership, design and implementation. Because the owner operates the desalination plant with an in-house staff, it also retains all opportunities to take advantage of cost-savings that membrane technology advancements could yield in a long term. The key disadvantages are that the owner takes practically all risks associated with project development (permitting and permit compliance, site availability and underground conditions, future power tariff changes, potential environmental damages and associated mitigation efforts); project implementation (faulty design, technology and equipment selection blunders, construction contractor deviations from engineering specifications; start up and commissioning risks and delays); and project financing. The owner also takes all risks associated with the long-term project operations and performance–such as the risks that the desalination plant may not be capable of: producing desalinated water at or above the design capacity; operating at or below the projected power, cartridge filter, membrane and chemical use; and of meeting all applicable product water quality and concentrate discharge regulations. Since the owner is responsible for the project financing, it also caries the financial burden associated with the project, including reduction of the owner's available bonding capacity for implementation of future projects.

This project delivery method is most suitable for owners that have prior experience with the permitting and implementation of seawater desalination projects and operation of desalination plants. For owners lacking such experience, the use of the design-bid-method of delivery is advisable for the implementation of small desalination projects with low-risk profile, which would allow them to gain the necessary experience and develop in-house desalination plant O&M capabilities.

18.8.3. Design-Build-Operate (DBO)

Key project parties and their roles. Similarly to the DBB method of project delivery, the DBO approach also involves asset ownership by a public entity (utility or municipality). Under this method of delivery the owner is responsible for project development, permitting and financing as well. The owner's consulting engineer typically develops detailed performance specifications and preliminary project design, which are than used to prepare tender and retain a DBO contractor that is responsible for the final process design, and for the detailed design, construction, startup and commissioning, as well as for the long-term operation of the desalination plant. Usually, the DBO contracting team consists of an Engineer, a Contractor and a private operations company (Operator).

Key advantages and disadvantages of DBO. Key advantage of DBO method of delivery as compared to DBB project implementation approach is that the early coordination of the facility planning and design with key construction activities and plant O&M requirements allows optimizing desalination plant design and reducing life-cycle water production costs. Another advantage for the public entity (utility or municipality) which would use the desalinated water is that it retains the ultimate ownership of the desalination plant. In addition, under this method of delivery the owner transfers most of the plant O&M risks to a private operator that has the experience and skills to manage these risks more cost effectively.

A modified DBO approach used in Australia is the "alliance" contracting concept. Under this delivery method, the owner (the public partner) and the private DBO contractor share responsibilities, risks and rewards for project delivery and performance. For comparison, under the traditional DBO approach, used in the US and elsewhere, the risks are clearly allocated to the respective parties responsible for project delivery, and commercial and legal penalties apply for failure to deliver on the contractual commitments of either party of the public-private partnership. The project alliance agreement establishes predetermined cost, schedule and performance targets which both the public and the private partner collectively agree to meet at the beginning of the project. For targets that are not met, both parties share the risks and the losses associated with project implementation. For project areas where the actual performance and costs are lower than the initially set targets, both parties share the monetary benefits. This approach allows the private contractor to reduce its contingency component of the costs, thereby reducing the initial cost of services and to trade some of the project savings, which the contractor would otherwise keep, for a lower overall risk exposure. The "alliance" project delivery method gives an opportunity to the public agency to be more actively involved throughout the project implementation and to exercise more control over the final product. These benefits are traded for taking upon some of the project design and construction risks that are traditionally apportioned to the private DBO or BOOT contractor.

Key disadvantage of the DBO method of delivery is that the public agency carries engineering and construction risks similar to these typical for the DBB approach of project delivery. Some of these risks are somewhat reduced because a number of the design concerns and potential construction deviations are diminished by the fact that the Contractor is closely involved in the project design. Another disadvantage is that the public agency carries the project's fiscal (i.e., financing), and permit compliance responsibilities and associated cost burdens.

18.8.4. Build-Own-Operate-Transfer (BOOT)

Key project parties. The main difference between this and the other two methods of delivery is that the public entity purchases water (a commodity) rather than a physical asset (the desalination plant). The project ownership is retained by the BOOT contractor.

Under this project delivery method the BOOT contractor is responsible for all aspects of project implementation, including: environmental and construction permitting; design; equipment procurement; construction, startup, and commissioning; long-term operations and permit compliance; and project finance. As indicated previously, the BOOT projects are usually financed with a combination of equity and debt. The debt bond/commercial construction loan repayment obligations for this type of projects are typically revenue-based and are "non-recourse" to the private project company that delivers the project and the public agency purchasing the desalinated water, because the net worth of the owners of the project company and the public agency does not have to be used to provide security for debt repayment.

The public or private entity that is the final user of the desalinated water procures a turnkey BOOT contractor based on a performance specification developed by the owner's engineer. The BOOT contractor sells product water at a guaranteed price, quality and quantity, and point of delivery under a water purchase agreement (WPA). The key terms of a typical WPA are discussed in the following section.

Once the terms for payment of services are set by the WPA, the BOOT project owner/developer usually retains a turnkey contractor to provide all engineering, procurement and construction (EPC) services needed to build and commission the desalination plant, and a private O&M contractor to operate the plant over the entire term of the WPA. Often, the BOOT project owner/developer may also serve as an EPC and/or O&M contractor and may provide a portion of or the entire amount of equity needed to finance the project.

The WPA, EPC and O&M contracts in combination with other entitlements, such as environmental and construction permits; land purchase or lease agreement; power purchase agreement; agreement for access to source water; and agreement for concentrate and waste disposal services, are used as a proof of control of the BOOT contractor over the project cash flow, which is necessary to secure private financing for the BOOT project. The financing costs associated with the project are a direct function of the strength of the BOOT project's contracts and the financial and operating strength of the entity purchasing the water and the EPC and O&M contractors. A well structured BOOT project with good WPA, EPC and O&M contracts and willing participants typically can be financed with 80% debt and 20% equity. If the project structure is strong and the project risk profile is favorable, a lower percentage of equity may be required. More detailed discussion of BOOT project financing is provided in Section 18.5.1.4 of this Chapter.

Water Purchase Agreement. The WPA guarantees water delivery to the user of the desalinated water (public or private entity) at pre-determined quantity, quality and availability over the entire term of the agreement. On the other hand this agreement guarantees a pre-determined payment for the delivered water to the BOOT contractor and thereby, secures a revenue stream that the BOOT contractor can pledge to obtain project financing. The key provisions recommended to be incorporated in a well structured water purchase agreement in order to minimize the project financing cost and therefore, the overall cost of water production are:

- *"Take or pay" clause*—by which the water purchasing entity agrees to purchase a minimum amount of water at any given time and/or to pay for the fixed costs of water incurred by the BOOT contractor, if the desalination facility is put on "standby. "

- *Firm water purchase obligations*—the contract should not contain provisions that allow the purchasing entity to unilaterally terminate or substantially revise the contract in the future.
- *Provisions to assign water contract to lenders*—the financial institutions that will provide equity and debt funds for project implementation should have the right and ample opportunity to cure project default if the BOOT contractor fails to perform its obligations under the WPA.
- *Firm and clear water tariff structure*—the WPA should have a water tariff structure that provides adequate coverage of the fixed water production costs and includes water cost escalation factors tied to third-party commodity (power, chemicals, labor, etc.) price adjustment indexes and foreign exchange fluctuations.
- *Change in law clause*—which allows the BOOT contractor to adjust the water tariff in order to reflect the additional costs which the BOOT contractor will incur in order to comply with future environmental and/or other regulations than have material impact on the water production costs.
- *Unambiguous water quality standards*—the WPA should contain clear specifications of the product water quality and quality; the plant capacity availability factor; the location/s of water delivery; and the procedures for measurement of the delivered water flow and monitoring of the quality of the desalinated water.
- *Liability for third-party claims*—the WPA should have provisions protecting equally both the BOOT contractor and the water purchaser from claims from the ultimate water consumers. In most cases, the BOOT contractor sells the water to a wholesale water supply agency, which in turn conveys and distributes the product water to the actual consumers. The BOOT contractor can only be hold liable for the product water quality at the point of delivery to the wholesale agency and cannot take the responsibility for changes in water quality caused by malfunction of the wholesale supplier's distribution system and conveyance facilities. On the other hand, the BOOT contractor should carry liability for impacts on the wholesale supplier's distribution system, if the BOOT contractor supplies inferior out-of-spec product water quality which is the cause such impacts.

Water purchase agreements have a number of other provisions which aim to define contractual division of responsibilities and risks between the BOOT contractor and the water purchaser. These provisions may vary from project to project, but in general have to be such that the project risks are apportioned between the BOOT contractor and the water purchaser commensurate with their ability to control and mitigate the risks and to deliver water to the ultimate consumer at lowest overall cost and competitive market price.

Key advantages and disadvantages of BOOT projects. As shown in Table 18.14, most of the large seawater desalination facilities built over the past 10 years, or currently undergoing construction, are delivered under public-private partnership arrangement using BOOT method of project implementation. The BOOT project delivery is preferred by municipalities and public utilities worldwide because it allows cost-effective transfer to the private sector of the risks associated with the number of variables affecting the cost of desalinated water, such as: intake water quality and its sometimes difficult to predict effects on plant performance; permitting challenges; startup and commissioning difficulties; fast-changing membrane technology and equipment market; and limited public sector experience with the operation of large seawater desalination facilities.

18.9. Project schedule

A detailed project implementation schedule has to be developed during the design phase of the seawater desalination project. The desalination plant construction schedule should as a minimum include the following information:

- The total suration of the project implementation
- Duration and start date of contractor mobilization and site preparation
- Duration and start date of the project engineering and design
- Duration and start date of procurement and installation of high pressure RO pumps, energy recovery equipment; high-pressure stainless steel piping; RO membrane elements and any other significant long lead time items, which procurement, installation or start up requires over three months
- Duration and start date of construction of intake facilities, intake and discharge interconnecting piping; pretreatment system; RO system and post-treatment facilities
- Duration and start date of plant commissioning and start up
- Duration and start date of desalination plant acceptance testing

TABLE 18.15
Typical length of desalination project implementation

Plant capacity m^3/d (MGD)	Design period (months)	Construction period (months)	Start-up and commissioning (months)	Total (months)
1,000/(0.3)	1–2	2–3	1–2	4–7
5,000/(1.3)	2–3	4–6	1–2	7–11
10,000/(2.5)	2–4	6–8	1–2	9–14
20,000/(5.3)	3–5	8–10	2–3	13–18
40,000/(10.6)	3–6	14–16	2–3	19–25
100,000/(26.4)	5–8	18–20	3–4	26–32
200,000 (52.8)	6–10	20–24	3–4	29–38

Note: Accelerated implementation of some of the activities is possible but is likely to result in cost increase.

Table 18.15 presents a typical length of desalination project design and construction as a function of the plant size.

The total length of the desalination plant project design and construction may vary from the indicative periods indicated in Table 18.15 depending on the site specific project scope and conditions. Some construction activities may take longer than the duration indicated in the table, especially if most of the construction has to be completed in adverse weather conditions; if the plant footprint is too compact; if the construction staging area is very limited; and/or the access to the site and hours of the day and days of the week during which construction is allowed are burdened with significant constraints due to noise, traffic, air pollution or other regulatory requirements. Some of the construction activities may be accelerated by working in multiple shifts and pre-purchasing some of the long-lead equipment and piping. However, such project acceleration activities usually result in an increase in the overall plant construction costs.

19

Concentrate management

*Mike Mickley**

19.1. Introduction

The ultimate goal of concentrate management is to dispose/use concentrate in a cost-effective and environmentally safe manner that does not simply transfer fate concerns to another situation. RO concentrate management is playing an increasingly important role in the determination of desalination plant feasibility due to the growing challenges of meeting this goal.

In areas such as the arid southwestern United States, desalination plants are not being built because there are no cost-effective concentrate management options. In locations where traditional concentrate disposal options are feasible, the time, effort, and cost of implementing these options is increasing. While water production costs have been decreasing, concentrate management costs have grown. Concentrate management costs are becoming a larger fraction of total plant cost. With these growing challenges it is imperative that concentrate management be considered early in desalination project planning to avoid later project delays.

In this chapter the issues, practices, and trends in concentrate management are reviewed. The intent is to provide an understanding of the growing challenges and to provide a basis for considering how concentrate management options may be screened in an initial feasibility determination at a given desalination site. Final

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feasibility determination rests on detailed analysis of the management options surviving the screening phase–considerations beyond the scope of this chapter.

An appendix C provides examples of how traditional disposal options may be screened to determine their feasibility.

19.2. Nature of concentrate

Some common characteristics of desalination concentrates are:

- They are of higher salinity than the feed water
- They contain higher concentrations of most feedwater constituents
- Concentrate water quality is site-specific because raw water and feed water are site specific
- Concentrate is composed mainly of raw water constituents with some process-added chemicals

Constituents added to concentrate may include:

- Acid (sulfuric or hydrochloric)
- Antiscalant (synthetic chemicals)
- Residual chemicals from other pretreatment steps
- Chlorine
- Dechlorinating species (such as sodium bisulfite)

The reader is referred to Chapter 8.2.2. for a detailed description of pretreatment chemicals and their use. Due to the relatively low levels of these chemicals in concentrate, concentrate reflects the makeup of the raw water and has been described as ‘raw water concentrated’. Environmental concerns, however, may be associated with the increased salinity of concentrate, the increased concentration of species present in the raw water, and the concentration of process-added chemicals.

The concentration factor, CF, relates the concentration of a given constituent in the concentrate to the feed concentration by the relationship (138):

$$\mathrm{CF} = C_c/C_{f0} = (1 - R)^{-r} \tag{1}$$

where CF is the concentration factor

C_c is the concentrate concentration

C_{f0} is the feed concentration
R is the water recovery
r is the average rejection of the constituent in question

For SWRO membranes where ion rejections are close to unity ($r = 1$), the concentration factor relationship simplifies to the more familiar form:

$$\mathrm{CF} = (C_c/C_{f0}) = [1/(1 - R)] \tag{2}$$

This expression can be used to estimate concentrate levels for ions and TDS for SWRO processes. Due to the slightly lower but still very high ion rejections of BWRO membranes, the expression is also useful for BWRO processes.

For nanofiltration membranes where rejection of monovalent ions is considerably less than unity, individual ion and TDS concentration factors cannot be simply estimated in this way without the risk of considerable error. Readily available membrane system software should be used to predict NF concentrate characteristics (and, of course, can be used to predict SWRO and BWRO concentrate characteristics).

Table 19.1 defines typical feed TDS range and recoveries for NF, BWRO, and SWRO systems.

From this table the differences between NF, BWRO, and SWRO concentrates become more apparent. Specifically:

- The volume of concentrate relative to feed volume is much greater for SWRO than for BWRO and NF.
- The concentrate salinity is typically much greater for SWRO than for BWRO, with NF concentrate typically being of lower or similar salinity relative to BWRO concentrate.

Concentrate management possibilities and practices differ according to concentrate characteristics being somewhat different for SWRO, BWRO, and NF concentrates. In the following sections discussion focuses on BWRO concentrate. Seawater and NF concentrate are discussed in a later section.

19.3. Concentrate management options

The term ‘concentrate management’ has come to replace the more limiting term ‘concentrate disposal’. Management includes disposal and also allows consideration of concentrate as a resource for beneficial use. Beneficial use of

TABLE 19.1
Typical operating ranges for NF and RO processes

Membrane system (mg/L)	Feed TDS (mg/L) range	Recovery (%)	Minimum concentrate TDS (mg/L)	Maximum concentrate TDS
NF	< 1000	90+	< 1,000	~12,000[1]
BWRO	1,000–20,000	60–85	2,500	~70,000[2]
SWRO	30,000 and above	30–60	43,000	~70,000[3]

Note 1: assuming 97% recovery and TDS rejection of 70%

Note 2: limited by osmotic force considerations; not possible without pretreatment to remove limitation of scaling by sparingly soluble salts and silica

Note 3: limited by osmotic force considerations

concentrate means the same drop of water will be used more than once and this may free up higher quality water resources for other uses. The term 'management' is also more palatable to public perceptions than 'disposal'.

Concentrate management options include:

- traditional disposal options
- less conventional disposal options
- beneficial use options
- volume reduction treatment of concentrate
- zero liquid discharge treatment of concentrate

The role of the 'newer' concentrate management options is discussed following review of traditional disposal options.

19.4. Traditional concentrate disposal options

The practices and regulations of concentrate disposal in the U.S. have been well documented and characterized (139) and serve as a basis for illustrating both practices and challenges.

As of 2005 over 230 municipal membrane desalination plants have been built in the United States for the purpose of producing potable water (140). About 95% of these plants are inland brackish water plants. Five traditional disposal options account for over 98% of the disposal situations. The traditional disposal options and their frequency of use for plants built by 2003 (139) are:

• Discharge to surface water	45%
• Discharge to sewer	27%
• Deep well injection	16%
• Land application	8%
• Evaporation pond	4%

Figure 19.1 shows the relative use of these disposal options as a function of desalination plant size.

Several characteristics and trends are reflected in this figure and include:

- Relatively high use of surface discharge for all sized desalination plants
- Decreasing use of disposal to sewer with increasing plant size (due to the effect of higher salinity concentrate on wastewater treatment plant biologicals and effluent salinity)
- Increasing use of deep well disposal with plant size (reflecting a large economy of scale of deep injection wells but a large cost for small capacity wells)
- Low frequency use of land applications (spray irrigation, percolation ponds) and evaporation ponds with larger plants (due to lack of economy of scale and relatively high per acre cost)

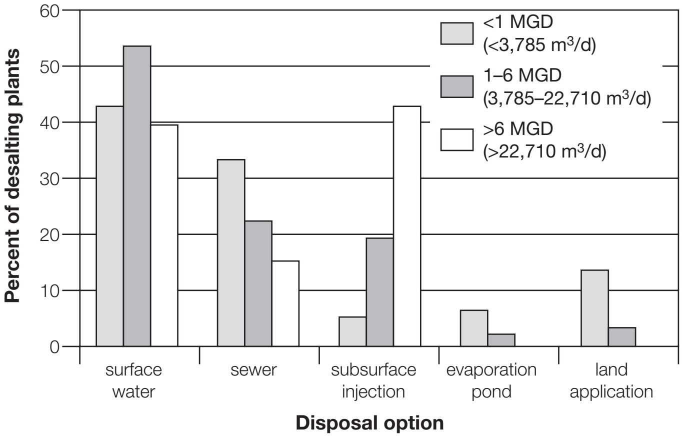

FIG. 19.1 Use of traditional disposal options by plant size.

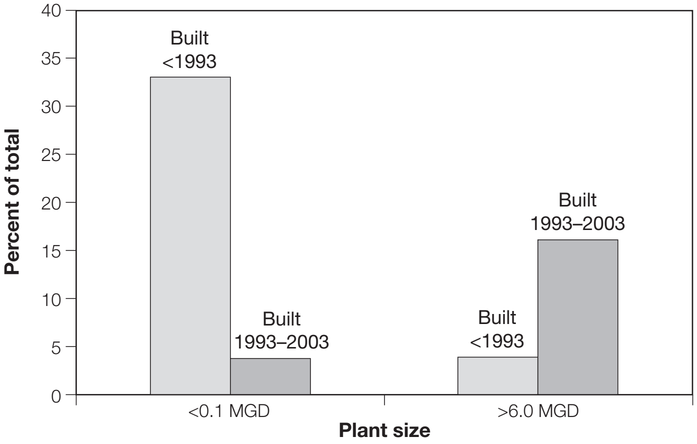

FIG. 19.2 Increase in plant sizes after 1993.

- Low frequency of use of land applications and evaporation ponds in general (since year-round operation is not possible in many parts of the United States)

The use of land applications and evaporation ponds has decreased in the United States due to the growing size of membrane desalination plants that is reflected in Fig. 19.2.

Characteristics, cost factors, environmental concerns, and regulatory basis are highlighted in Table 19.2 for these 5 traditional disposal options.

The above composite statistics on the five disposal options are somewhat misleading in that while discharge to a surface water or the sewer account for 72% of the cases in the United States, they account for 100% of the cases for 21 of the 25 states having municipal desalination plants. The other 3 traditional disposal options (deep well injection, land application, and evaporation ponds) have limited widespread application, primarily due to hydro-geological and climate requirements. Thus the five traditional disposal options are not widely available; local availability of options is typically very limited.

Limiting factors affecting availability of the traditional disposal options include:

- *Climate*: year-round use of land applications and evaporation ponds is not possible in colder climates
- *Adequate hydro-geological conditions*: deep well injection requires adequate aquifer isolation, capacity, injection rate, and freedom from earthquakes–conditions not often found

(text continues on p. 385)

TABLE 19.2

Characteristics, cost factors, environmental concerns, and regulatory basis of conventional disposal options

Surface water discharge	
Characteristics	• Includes disposal to river, creek, lagoon, ocean, etc. • Need relatively nearby moving water of sufficient volume year round • Not usually available in arid regions • Used with all size plants • Requires a discharge permit
Major cost factors	• Some pretreatment may be required for groundwater based concentrates ◦ Aeration to increase dissolved oxygen ◦ Degasification for H_2S, CO_2 • Piping and pumping • Post-treatment such as aeration, degasification, pH adjustment • Outfall structure • Monitoring • Frequently a lower cost disposal option
Environmental concerns	• Regulation based on compatibility of concentrate with receiving water (salinity and individual constituents) • Low dissolved oxygen levels and possible high dissolved gas (H_2S, NH_3) levels in concentrate from groundwater sources • Concerns of salt and mineral loading • Potential detrimental impacts on environmentally sensitive areas • Risk of CO_2 emission from discharged alkaline waters • Increasing globally disregarded as a sustainable inland saline effluent management option • Mitigation: enhanced mixing and dispersion through discharge site choice, use of diffuser, outfall design; pre-dilution
Regulation basis	• Based on comparison of concentrate flow and water quality with receiving water flow and water quality • Typically tied to a worst case scenario: a low receiving water flow condition (based on historical records) and maximum concentrate flow • Receiving water standards based on its use classification • May include whole effluent toxicity test requirements

continued

TABLE 19.2 *continued*

Characteristics, cost factors, environmental concerns, and regulatory basis of conventional disposal options

Surface water discharge, *continued*	
	◦ For concentrate from groundwater sources there is a concern with major ion toxicity • Mixing zones may be granted for establishing compatibility conditions • Monitoring requirements ◦ Typically need to periodically monitor TDS, designated specific constituents, dissolved oxygen, pH, others
Sewer disposal	
Characteristics	• Offers advantage of lowering the BOD of domestic sewage effluent • Used less frequently with increasing concentrate volume
Major cost factors	• Piping and pumping • Fee charged by wastewater treatment plant
Environmental concerns	• Compatibility issues ◦ TDS ◦ Specific constituents ◦ other • Salt loading • Effect of concentrate salinity (and constituents) on biologicals, effect of concentrate salinity on effluent salinity
Regulation basis	• Generally no permit is required but need permission of wastewater treatment plant and they may enforce treatment requirements • Based on comparison of concentrate volume and water quality with wastewater treatment plant influent volume and water quality • Comes down to a decision on the part of the wastewater treatment plant; they do not have to grant permission
Deep well injection	
Characteristics	• Wastes the resource • Need adequate aquifer characteristics ◦ Structurally isolated from and below drinking water aquifers ◦ Sufficient capacity ◦ Sufficient permeability, porosity for adequate injection rate

	• High cost for small volume but good economy of scale • Used primarily with larger volumes • Adequate aquifer characteristics are not often found • Expensive feasibility determination • Possibilities include reworking abandoned wells
Major cost factors	• Piping and pumping • Possible pretreatment • Land preparation • Mobilization • Testing • Backup disposal system for during system integrity tests (periodic) • Monitoring wells
Environmental concerns	• Potential for aquifer water movement and contamination of other aquifers • Potential for leakage from the well into other aquifers • Long-term sustainability under question due to doubts with the results from fate and compatibility studies • Concern with earthquakes • Concern with overpressure causing fracture and earthquakes • Concern with mixing of concentrate and receiving water
Regulation basis	• Typically a multi-step process involving test wells • Construction requirements based on well type (Class I well in the United States) • Periodic well integrity tests are required • Requires a permitted second disposal option for use during periodic integrity tests
Land application	
Characteristics	• Includes spray irrigation of lawns, parks, golf courses, crop lands • Also includes percolation ponds, raid infiltration basins • Land intensive • May require dilution water • Relatively level land required • Climate dependent • Low economy of scale • Used with smaller concentrate volumes

continued

TABLE 19.2 *continued*

Characteristics, cost factors, environmental concerns, and regulatory basis of conventional disposal options

Land application, *continued*	
Cost factors	• Piping and pumping • Land preparation • Storage for use during rain days • Distribution system with associated valves and control • Costs associated with possible dilution requirements • Costs associated with drainage system (if required) • Costs associated with possible surface runoff control system
Environmental concerns	• Possible surface runoff and downstream impacts • Compatibility concerns ○ with underlying ground water ○ with vegetation (SAR is important) ○ with soil • Concern with salt loading
Regulation basis	• Based on concentrate characteristics and land use classification and resulting groundwater standards • Possible drainage system requirements
Evaporation pond	
Characteristics	• Land-intensive • Requires level land • Low economy of scale • Climate dependent • most suitable for areas with high evaporation rates • Used for smaller concentrate volumes • Evaporation rate decreases as solids level/salinity increases
Cost factors	• Piping and pumping • Land • Land preparation • Possible distribution system with associated valving and control • Synthetic or clay liner • monitoring • Solids disposal

Environmental concerns	• Environmental concern of wildlife effects • Leakage risk • Produces a concentrated brine or solid which by itself requires safe disposal means • Risk of CO_2 emission from high alkaline waters • Source of salt spray to the surrounding environment
Regulation basis	• Regulatory requirements for natural or synthetic liners • Monitoring requirements

- *Land availability, cost, and topography*: land applications and evaporation ponds require large amounts of inexpensive land per volume unit of concentrate, and the land needs to be relatively level; these conditions can be limiting; many of the cities in the arid southwest United States are areas of high population growth and have high land costs.
- *Compatibility of concentrate with receiving water*: The acceptability of direct discharge to a receiving water, of discharge to the sewer, and of land application all depend on the salinity and water quality of the concentrate relative to the receiving water (surface water, sewer water, and groundwater, respectively). The greater the difference between concentrate levels and receiving water levels, the less likely the application will be feasible.
- *Economy of scale*: Economy of scale favors use for larger concentrate volumes; deep well injection has a good economy of scale; land applications and evaporation ponds do not.
- *Availability of dilution water*: Frequently concentrate has higher TDS and individual constituent levels than are allowed for land application. Availability and use of dilution water may make land application possible–at the expense of the resulting larger volume of diluted concentrate that requires disposal; a situation which is typically cost-effective only at smaller volumes.

Figure 19.3 depicts the relative cost of the different disposal options for most situations (141). The costs of conveyance of concentrate to the site of disposal are not included. There are many exceptions to this representation as concentrate itself is site-specific, available disposal options are site specific, and the above limitations while not eliminating feasibility may play a role in determining the site specific cost of that disposal option.

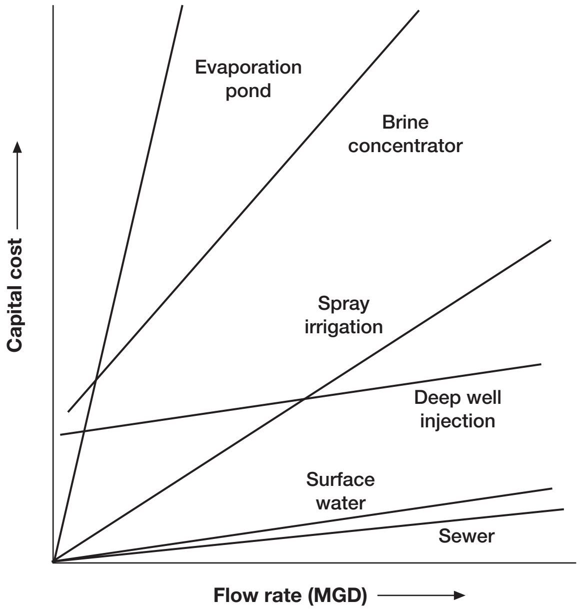

FIG. 19.3 Relative capital costs of traditional disposal options and zero liquid discharge.

Design parameters and preliminary cost models for the traditional disposal options are provided in a Bureau of Reclamation publication (139).

Appendix C provides examples of how traditional disposal options may be screened to determine their feasibility.

Less conventional disposal options.

In a few situations non-traditional disposal options have been found because of site-specific conditions. They are not necessarily of wide applicability. One example is the use of BWRO concentrate as partial feed for SWRO plants (152). The feed to the Israeli Eliat SWRO plant is 80% seawater and 20% BWRO concentrate. In addition to providing for disposal of the BWRO concentrate, cost savings were realized due to reduced feed pressure and threfore reduced energy consumption at the SWRO plant.

19.5. Concentrate disposal challenges

As stated at the beginning of the chapter, the ultimate goal of concentrate management is to dispose of the concentrate in a cost-effective and environmentally safe manner.

Cost reduction of traditional disposal options is unlikely. Their feasibility is less dependent on technology and equipment than on the non-technical factors and limitations discussed above. While the cost of desalting water using membrane technologies has decreased, the cost of disposal has not–giving rise to the disposal costs becoming an increasing percentage of total plant cost.

From the previous section it is evident there is a limitation in local availability of options (regardless of concentrate volume), and the increasingly larger volumes of concentrate, in general, further reduce the feasibility of the traditional disposal options.

The challenges of finding a suitable disposal option are further complicated by more stringent regulation and increased public concern, both of which impact the regulatory and permitting process.

While traditional disposal options will continue to play an important role in many locations and be cost-effective, there is a growing number of locations where the time, effort, and cost of implementing them will increase, and there are some locations, such as in the arid southwestern United States, where traditional disposal options are not possible or cost-effective.

In recent years there has been increased concern over concentrate representing a lost water resource for the desalination utility. In some evaluations of concentrate disposal options a cost has been assigned to the concentrate equal to the cost of obtaining water rights to replace the same volume of this 'lost resource.' The concern has also increased attention on volume reduction of concentrate, discussed in the next section.

Disposal of concentrate to surface waters, to the sewer, and to land applications may result in salt loading of the receiving water. Although discharge may be acceptable from the perspective of not violating receiving water standards and being able to secure disposal permits, various constituent concentrations and salinity of these waters may increase. Future discharges become more limited, as there is less capacity of the receiving water to uptake the salt and still be within receiving water standards. At some point, future discharges will not be possible. Disposal via these options is not sustainable.

19.6. Consideration of concentrate management options

Increased disposal challenges have spurred research into other concentrate management options. These options include beneficial use of concentrate, volume reduction of concentrate, and zero liquid discharge processing of concentrate.

TABLE 19.3
Beneficial uses of concentrate

Oil well field injection	Subsurface storage
Solar ponds	Feedstock for hypochlorite generation
Aquaculture	Cooling water
Wetlands	Dust control and deicing
Transport of mineral resources	Scrubber water

19.6.1. Beneficial use

Table 19.3 lists several possible beneficial uses of concentrate. While such possibilities exist, most if not all have considerable limitations (142). Many beneficial uses do not provide for disposal of the concentrate–they use the concentrate and leave a residual stream to be disposed. An example is a wetlands where concentrate may serve an important environmental function but still requires disposal of the effluent from the wetlands. Another example is an aquafarm where concentrate can support a fish industry but results in an effluent, now with an increased organic component, to be disposed. Most beneficial uses do not have widespread applicability, most are unproven, and most do not address the concentrate disposal challenge. However, given the growing challenges of concentrate disposal, beneficial use options should be evaluated at an early screening stage of management options.

19.6.2. Volume reduction

With the concern for concentrate representing a lost water resource, and given the growing general challenge of disposing concentrate, there has been an interest in volume reduction of concentrate. The usual means of accomplishing this is a second stage RO system. To avoid the limitation due to sparingly soluble salts and silica, either treatment of the concentrate to remove these species or combination of treatment and high pH operation of the second stage are typically employed. The second stage concentrate is now limited, as in seawater RO, by osmotic forces, which, after scalant removal, limit final concentrate to a TDS of from 65,000 to 80,000 mg/L. While this volume reduction is possible, it does not, in general, facilitate concentrate disposal. The resulting more saline concentrate (now a brine) is more incompatible with most receiving waters (surface water, sewer, groundwater), thus eliminating these disposal options from consideration. Given the limitations on availability of deep well injection and evaporation ponds, this eliminates all traditional disposal options in many locations.

19.6.3. Zero liquid discharge

While zero liquid discharge (ZLD) has not been used in municipal drinking water plants in the United States, it is a widely used commercial technology applied in many industries, including the power industry. Mechanical vapor recompression evaporators (brine concentrators) are most frequently used for taking the feed solution up to a brine concentration of 180,000 to 280,000 mg/L depending on the water quality. Brine from the brine concentrator is typically sent to evaporation ponds or further processed to dry salts by a thermal crystallizer or, if the volume is small, by a spray dryer. A variation of this processing scheme involves the use of a second stage RO unit to reduce the volume going to the brine concentrator. A third variation is to eliminate the brine concentrator altogether and to use the final processing steps of evaporation pond, crystallizer, or spray dryer on the brine from the second stage RO.

ZLD is presently prohibitively expensive for most municipal settings. A representative relative cost of ZLD is shown in Fig. 19.3. Analysis of these options (143, 144) has shown that, while volume reduction prior to the brine concentrator can reduce energy requirements significantly, these costs are in large part replaced by chemical costs for treatment and increased solids disposal costs. This is particularly true of high hardness waters.

There are few uses of concentrate as concentrate, of brine derived from concentrate through volume reduction, and of mixed solids produced from ZLD operations (from evaporation ponds or crystallizers, or spray dryers). One reason for this is the growing requirement for concentrate, brine, or mixed solids to meet environmental or application specific standards–whether for soil remediation, de-icing salts, dust suppression, or the variety of other possible applications.

Historically, concentrate management has been concentrate disposal; and in the vast majority of cases it will be so in the future. This is not surprising when concentrate is viewed as a ‘waste product.’ Concentrate shares some characteristics with other ‘wastes’ in that disposal options are limited and in large part have been well defined for years. More recent trends have been to treat wastes to develop marketable products through ‘recycle’ programs. This possibility holds some promise for concentrate in terms of selectively and sequentially removing individual salts from concentrate in a manner meeting salt use specifications. Beneficial use makes most sense in this context, transforming concentrate into useful products. Volume reduction makes most sense when it is part of a ZLD process. This drive towards 'sustainability' for concentrate is the same giving rise to recycling of other wastes: resources are limited and the 'wastes' can be recycled and made into useful products. The challenges of doing this for concentrate are considerable. The commercial technology exists to accomplish this (145); however, production of marketable salts transcends the traditional role of a water treatment plant. Another challenge is to develop local markets for the salts, such as value-added building materials, to avoid saturation of local existing salt markets and dependence on them (146, 147).

Volume reduction and ZLD processing are expensive and should, in general, be considered only when traditional disposal options are not feasible. In the short-term they may provide a feasible, though costly, solution where otherwise there are none. In time, with cost reductions in technology and rising water values, the costs of volume reduction and ZLD processing may permit their more widespread use.

19.7. Seawater desalination discharge

Seawater concentrate is much different from brackish RO concentrate in terms of salinity and volume. Due to the coastal location, the obvious and traditional disposal option for seawater concentrate is back to the source water. The high salinity and large volume of the concentrate effectively preclude use of the other traditional disposal options.

Critical factors involved in potential environmental impacts from brine discharges include (148, 149):

- the nature of the local eco-systems
- the extent of mitigation measures taken
- the size of the discharge
- the local currents—or more broadly the mixing and flushing conditions of the receiving water

The environmental concerns include those due to both the salinity of the concentrate and the constituents present in the concentrate. Potential impacts from undiluted concentrate may be from:

- constituents present in the feed water
  some raw water components are modified or removed as a result of pre-treatment
- higher concentration of these components than the receiving water; thus higher salinity
- residual chemicals from the pre-treatment process
  most pretreatment chemicals are removed as a result of sedimentation and filtering, but residuals remain
- heavy metals
  from intermittently used cleaning solutions and from equipment, pipe, and pump materials
- other components from cleaning solutions
  acid, base, detergents, complexing agents, enzymes, etc.
- chlorine from disinfection
- dechlorination chemicals
  such as sodium bisulfite from dechlorination
- lower dissolved oxygen
  as a result of dechlorination chemical use
- organohalogen compounds
  formed from interaction of chlorine with naturally occurring organic material
- pH difference from receiving water
- antiscalant
- temperature difference from receiving water
- variations in the above

While most organisms can adapt to minor changes in salinity (and other conditions) or temporarily manage higher deviations, the continuous discharge of higher saline effluent will be harmful to marine life. The individual concentrate properties also have potential effects on the marine environment and their effects may be additive or synergistic.

Most components of concentrate have a limited dispersal range so that the environmental effects are restricted to the discharge site (near-field) and its more immediate vicinity (far-field). The environmental fates include chemical changes (e.g., chlorine), transport into sediments (e.g. heavy metals), ingestion /

uptake (by flora and fauna), and dispersion / dilution. Most residual chemical concentrations in the concentrate are relatively low but may eventually amount to heavy loads due to the large concentrate volumes produced (148).

A major factor in determining the level of impact is the receiving water condition. From both simulation of and measurements in receiving waters for discharges in waters of limited mixing, the concentrate discharge forms a distinct mass characterized as a plume that originates at the discharge outlet and grows and disperses away from the outlet in the direction of net receiving water movement. The concentrate is of higher salinity and higher density and thus negatively buoyant. The plume sinks and spreads along the sea floor affecting the less mobile benthic organisms. The extent to which this will occur is dependent on the depth of the sea floor relative to the sea surface and the mixing/dispersion conditions of the receiving water. Some receiving waters are high energy, providing much mixing and dispersion; others are low energy such as in more confined bays that have long time periods for flushing. In a high energy receiving water of larger depth impacts will be minimized. In a low energy receiving water of shallow depth, impacts will be heightened. Use of diffusers at the end of the discharge pipe can increase mixing and dispersion and help to mitigate impacts.

In order to minimize the compatibility issues, many new seawater desalination plants are utilizing cooling water from power plants or effluent from wastewater treatment plants for dilution of the concentrate prior to discharge. Other advantages of this co-siting arrangement include shared use of intake and outfall structures and consequent easier permitting path.

19.8. Nanofiltration concentrate disposal

Typically at the other salinity extreme, NF concentrate is of low enough salinity and low enough volume for several traditional disposal options to be feasible. In particular, surface water disposal, disposal to sewer, and land applications usually have lower compatibility issues than are associated with discharge of brackish RO concentrate.

19.9. Other topics related to concentrate disposal

19.9.1. Presence of contaminants

Care must be taken in early feasibility studies to address the presence of contaminants in concentrate. A particular challenge that may increasingly occur is where concentrate contains a level of contaminant (arsenic, selenium, radionuclides, pesticides, etc.), such that treatment and removal are necessary prior to concentrate disposal. Contaminants may occur naturally or from the activities of man. The presence of contaminants could significantly increase the cost of disposal.

Naturally occurring radionuclides in groundwater in southwest Florida locations occur at levels that would prevent discharge of the concentrate to surface water. The availability of deep well injection has mitigated this problem.

Contaminants present at acceptable levels in groundwater and concentrate may result in a hazardous waste when they are processed to solids. Naturally occurring radionuclides are of concern in this regard (144).

19.9.2. Major ion toxicity

Major ion toxicity is possible in concentrates originating from groundwater. As with other toxicities it is determined in whole effluent toxicity tests that may be required for surface water discharge. Occurrence of the toxicity is dependent both on the test organism used (some organisms are more sensitive than others) and the concentrate water quality. This toxicity (150) may occur when common ions are present in high (or very low) amounts relative to those in a seawater diluted to the same salinity. Seawater may be considered to be ‘balanced’ in terms of major ion makeup with respect to most biological systems and major ion toxicity appears to occur when a water has a makeup ‘imbalanced’ relative to seawater. The toxicity has occurred most frequently in Florida with the mysid shrimp test organism and high calcium and or high fluoride levels. A case of very low potassium amount was also identified. Fortunately, the toxicity has sharp thresholds and dilution of the concentrate by a factor of 4 or 5 typically negates the toxicity.

19.10. Summary

Five traditional disposal options (disposal to surface water, disposal to sewer, deep well injection, land application, and evaporation ponds) account for nearly all disposal options for BWRO, NF, and SWRO processes. Characteristics, cost factors, environmental concerns, and regulatory basis of these traditional disposal options are presented in Table 19.2.

Rarely are more than a few of these options available for a given plant, due to limitations related to climate, size, hydro-geological conditions, and other factors. The trend of increasing plant size and concentrate volume further challenges the task of finding a feasible disposal option. While the traditional disposal options will continue to be feasible in many locations, there are some inland locations in which brackish water desalination plants are not being built due to the lack of cost-effective disposal options.

In an effort to define more concentrate management solutions recent trends include consideration of beneficial use of concentrate, volume reduction of concentrate, and zero liquid discharge processing of concentrate.

While all concentrate management options should be considered at an early stage of desalination plant planning, in general there are few beneficial uses of concentrate, brine resulting from volume reduction of concentrate, or mixed solids produced in zero liquid discharge processing of concentrate. One factor influencing this is the growing trend of concentrate, brine, and mixed solids needing to meet environmental requirements and application-related specifications. In addition, most beneficial uses are unproven, not widely available, and do not necessarily provide for disposal of the concentrate. Disposal to surface water, to sewer, and to land is usually not possible with reduced volume, higher salinity concentrate (now a brine). This leaves only deep well injection and evaporation ponds as disposal options–options that are not often available. Zero liquid discharge costs are currently prohibitive for municipal settings.

One promising possibility is to selectively and sequentially remove salts from concentrate as part of zero liquid discharge processing. While the technology exists, large-scale success will depend substantially on finding sufficient markets for the salts. Development of value-added products from the salts that could be locally used would be a strong step toward making this approach a greater reality (142).

Seawater concentrate will continue to be disposed back to the sea. Increasing environmental concerns dictate increased scrutiny of and care in discharge planning and design, as well as more extensive monitoring of the discharge situations.

Concentrate disposal is less equipment/process oriented than water production and is more dependent and limited by other factors. Unlike with water production costs that have been decreasing in recent years, BWRO, SWRO, and NF concentrate disposal costs will increase.

The challenges and cost of concentrate management are major considerations affecting the feasibility of municipal membrane desalination plants. These challenges need to be addressed early in the project life to assure cost-effective and environmentally safe solutions.

20

Hybrid systems and technology

*Leon Awerbuch**

20.1. Introduction

The hybrid desalting concept is the combination of two or more processes in order to provide a better and lower cost product than either alone can provide. In desalination, there are distillation and membrane processes which under hybrid conditions can be combined to produce a more economic process. Thus, two or three elements that are integrated to make hybrid desalination are:

- distillation: multi-stage flash (MSF), multi-effect distillation (MED), vapor compression (VC)
- membrane desalination: reverse osmosis (RO), nanofiltration (NF)
- power: steam power plants, combined cycle power plants

Large dual-purpose power-desalination plants are built to reduce the cost of production of electricity and water. Over 30,000 MW of power is combined with desalination plants in the largest use of cogeneration concepts.

In many countries, particularly in the Middle East, peak power demand occurs in summer and then drops dramatically to 30–40%. In contrast the demand for desalinated water is almost constant around the year. Therefore, the design of

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future plants requires careful consideration of the power (MW) to water (mIgd) ratio PWR.

An examination is made of hybrid systems and hybrid technology for both a simple and integrated approach in order to take full advantage of both thermal and electrical energy.

Two comprehensive studies were carried out on hybrid desalination systems by Daniel Hoffman and Amnon Zfati (155), Sherman May (156), and Awerbuch review (212).

20.2. Distillation desalination methods

Distillation methods were the first processes applied to seawater desalination and are currently very well established in the desalination market. Over 50% of all commercial desalination plants worldwide use distillation technology for water production. In the distillation process a high salinity feed water is boiled in an evaporator, using outside energy usually provided by high temperature steam, to increase its temperature to the boiling point. During this step, the heating steam condenses and returns to the boiler and seawater feed partially evaporates producing vapor with practically no dissolved ion content. Heated seawater flows through successive evaporating stages (or effects) of the system. In each subsequent stage, which is maintained at decreasing temperature and pressure, there is partial evaporation of the seawater and condensation of water vapor which forms a distillate. The distillate is collected together as a combined product water. Remaining seawater is discharged back to the sea.

In the evaporation desalination systems some of the heat energy is lost through transmission to the surrounding air and through discharged streams of cooling seawater, blowdown and product water. In order for the thermal desalination process to continue, these energy losses have to be compensated from external energy sources such as high temperature steam or electric heaters and compressors.

During water vapor condensation, the latent heat released is transferred through the internal heat exchange surfaces to the seawater feed, providing energy for boiling at a given temperature and pressure. The stage wise distillation process with repeated heat transfer from vapor to seawater, greatly increases thermal efficiency of the process. It enables production of number of units of distillate per each unit of steam, which was used initially to heat seawater. The common indicator of efficiency of the process is the Gained Output Ratio (GOR), which is defined as a mass of distillate per unit mass of heating steam provided from the boiler. Well designed systems have a GOR value of 6–11. The value of GOR increases with the number of evaporation stages and with the difference between the top temperature in the system and the initial temperature of seawater.

The most common commercial distillation methods used for seawater desalination are: multiple-effect distillation (MED), multi-stage flash (MSF) and vapor compression (VC). MED and MSF use steam to increase temperature of the seawater feed. VC utilizes mainly electrical energy to power the compressor to increase energy of the steam in the system. In all above distillation processes additional electrical energy is required for water pumping to and from the system and for internal recirculation.

Multi-effect distillation (MED)

A schematic configuration of the MED process flow diagram is shown in Fig. 20.1. The basic principle is straightforward. The feedwater flowing over a heat transfer surface in the first chamber (effect) is heated by prime steam, resulting in evaporation of a fraction of the water content of the feed. The most accepted process and widely used MED plant is of the horizontal-tube type in which the prime steam and all the downstream vapors flow inside the horizontal tubes, where they condense and contribute to the product water stream. The feed, meanwhile, is sprayed on the outside of the tubes producing vapor. The water

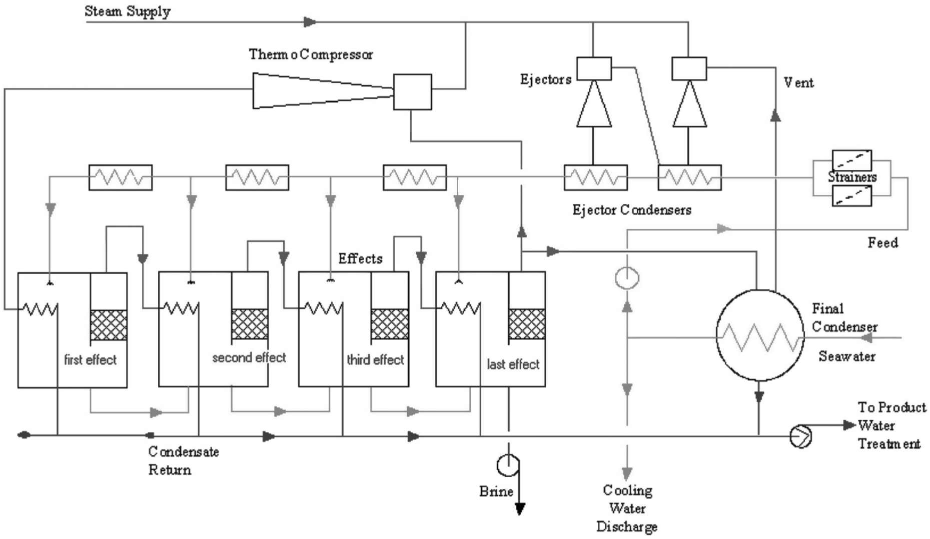


FIG. 20.1 Schematic diagram of the MED process.

vapor generated by brine evaporation in each effect of the horizontal-tube evaporator flows to the next effect where it supplies heat for additional evaporation at a lower temperature.

The partially concentrated brine is delivered to a second chamber (effect), maintained at a slightly lower pressure than the first effect. Likewise, the vapor liberated from the first effect feed is sent to the second effect. There the vapor condenses on the heat transfer tubes, giving up its latent heat to evaporate an additional fraction of water from the brine flowing on the opposite wall of the tube. The process of evaporation-plus-condensation is repeated from effect to effect each at successively lower pressure and temperature. The combined condensed vapor constitutes the product water.

The seawater feed, needs to be partially preheated by a fraction of the vapor from effects. The feed forward pattern has the advantage of minimizing scale formation, since the most concentrated brine is exposed to the lowest temperature. In each effect the feed has to be preheated to boiling temperature in the plenum, which distributes it to the tubes for further evaporation. It is critical to assure that the heat transfer surface is uniformly wetted, thus avoiding deposition of solids on dry spots. In the unlikely event of a leaky tube wall, the vapor (which is at a higher pressure than the brine) would leak into the brine chamber, thereby avoiding contamination of the product water.

The temperature and pressure in each subsequent effect is stepwise lower, providing equilibrium conditions for heat transfer between water vapors and seawater and resulting in seawater evaporation.

Each effect serves as a condenser for the vapor from the preceding effect; however, the vapor generated in the last effect is condensed in a final condenser, where the heat is rejected to a stream of cooling water.

MED plants can be made more capacity and energy-efficient by increasing the number of effects and the heat transfer area, or by increasing the maximum operating temperature. The MED specific power consumption is below 1.8 kWh/ton of distillate, significantly lower than MSF typical 4 kWh/ton. Thermal energy required by MED systems is provided by very low pressure steam (0.2–0.4 atm) or hot water sources above 55°C (>126°F). In the case of MED units with thermocompressor they utilize steam at 3.5–16 ata. The MED systems consist of between 10 to 16 evaporation effects and operate at a top temperature of about 70°C (158°F). MED-TVC units operating in the Persian Gulf operate at top temperature of only 63°C and use 5–8 effects. This relatively low top temperature results in reduced scaling and corrosion tendency of seawater in the MED process

MED units are available with capacities of up to 40,000 m^3/day (10.6 mgd) in a single unit with larger capacity plants being composed of multiple units. The future of MED technology is presented by Canaton (166).

The size of MED units has been growing rapidly in recent years, the largest operational plant of 22,700 m^3/d (6.0 mgd) operating in Sharjah. Recently awards in Sharjah increased the unit capacity to 36,400 m^3/d (9.6 mgd) and in Bahrain the Hidd project will build 72 mgd composed of 10 units each of 27,300 m^3/d (7.2 mgd).

Multi-stage flash (MSF)

A schematic configuration of the MSF process is shown in Fig. 20.2. The MSF Distillation desalination process accounts for the major portion of fresh water currently produced and is used primarily for the desalting of seawater. The principles involved are simple. In the MSF system seawater is pumped into heat rejection section of the plant where it exchanges energy with the lowest temperature water vapors. Most of the cooling seawater is returned to the sea. The remaining part is being used as a makeup. Before being used as a makeup this fraction of seawater is treated to remove dissolved gases. Also scale inhibitor or acid is added to reduce scaling tendency of seawater at high temperature.

After mixing with the recycle stream the seawater feed is pressurized and heated to the maximum top brine temperature (TBT). When the heated brine flows into a chamber maintained at slightly below the saturation vapour pressure of the

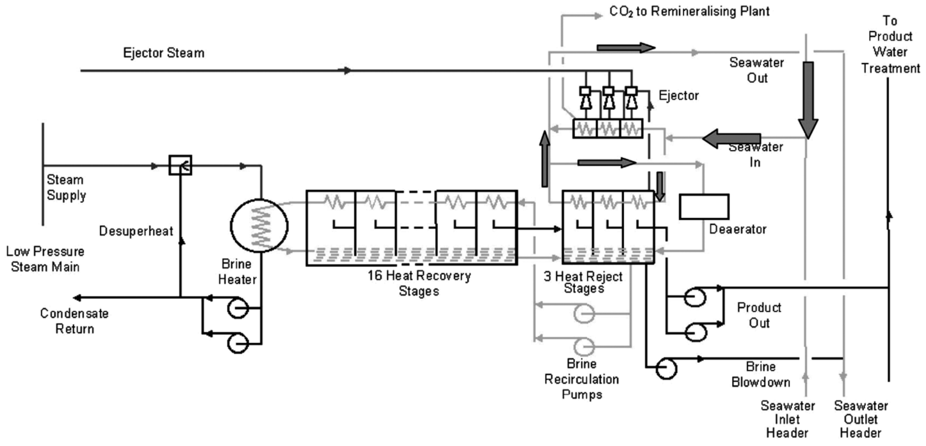


FIG. 20.2 Schematic diagram of the MSF process.

water, a fraction of its water content flashes into steam. The flashed steam is stripped of suspended brine droplets as it passes through a mist eliminator and condenses on the exterior surface of heat transfer tubing. The condensed liquid drips into trays as hot product water.

The unflashed brine enters a second chamber, or stage, where it flashes to steam at a lower temperature, producing a further quantity of product water. Simultaneously, the distillate from the first stage passes to the distillate tray in the second stage, giving up some of its heat and thereby lowering its temperature. The flashing-cooling process is repeated from stage to stage until both the cooled brine and the cooled distillate are finally discharged from the plant as blowdown brine and product water, respectively.

It is common practice to recycle a fraction of the blowdown water, combined with feedwater, through the entire circuit in order to extract an additional fraction of its water content. The recirculating stream, flowing through the interior of the tubes that condense the vapour in each stage, serves to remove the latent heat of condensation. In so doing, the circulating brine is preheated to almost the maximum operating temperature of the process, simultaneously recovering the energy of the condensing vapor. This portion of the MSF plant is called the "heat recovery" section.

The preheated brine is finally brought up to maximum operating temperature TBT in a brine heater supplied with steam from an external source.

At the cool end of the plant, a separate set of tubes is installed in several of the stages in a "heat rejection" section to remove the waste heat. The coolant there is generally not recycled brine, but the feedwater (in this example, seawater), of which the greater portion is discharged to waste. A small fraction of this coolant becomes preheated makeup water after proper pre-treatment and Deaeration. Thermal energy required by the MSF process is provided in the form of low pressure steam (2.5–3.5 atm). The low pressure steam is usually extracted from the end stages of electricity generating turbine.

Among the advantages of the MSF and other distillation processes is the fact that the composition of the feedwater to the plant has an almost negligible effect on the energy consumption per pound of product water delivered. This contrasts with the performance of other desalination processes in which the energy consumed is a direct function of feed composition.

Another advantage is that this process, in common with all distillation processes can produce comparatively pure water. Manufacturers typically submit bids containing warranties that the TDS will be less than 25 mg/L.

The major advantage of MSF is the achievement of large scale unit size. Designs of unit size of 16.7 million imperial gallons per day (mIgd) were achieved for the Shuweihat desalination plant in Abu Dhabi. A review of the future of MSF beyond the large size presented by Borsani et al. (192).

Vapor compression (VC)

The schematic diagram of the VC process is shown in Fig. 20.3. In the VC process seawater enters the pre-heater unit where it exchanges thermal energy with concentrate and product water effluents. Preheated seawater is sprayed over bundles of heat exchange tubing, which are at a higher temperature then seawater.

Seawater partially evaporates. Water vapors are compressed using a mechanical or thermal compressor and they are pumped inside the tubing of heat exchange bundles. The compressed vapor condenses inside the tubing, releasing energy that is transferred to seawater sprayed over the tubing and causes its evaporation. The condensed distillate and hot concentrated seawater flows through the pre-heater unit, exchanging its energy with seawater pumped in. After the pre-heater unit the product water is pumped to the post treatment. Part of the concentrate is discharged as a blowdown, and the rest is blended with fresh seawater feed and returned to the VC vessel.

The distillation desalination methods, MED and MSF, use steam to increase temperature of seawater feed. In addition, electrical energy is required for pumping feed seawater, brine and distillate. VC operation can be based entirely

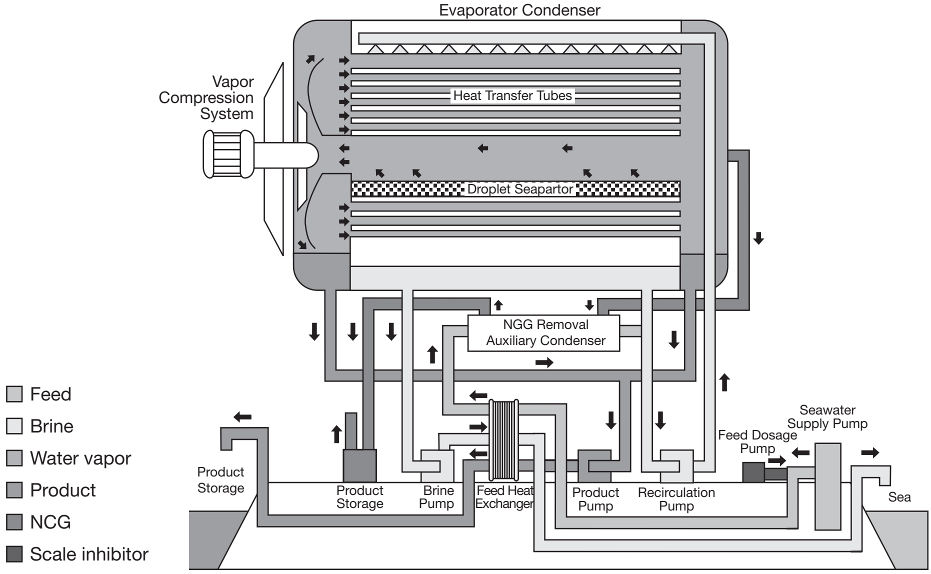


FIG. 20.3 Schematic diagram of the VC process.

TABLE 20.1
Representative energy values of thermal desalination processes.

Process/energy type	MED	MSF	VC
Steam pressure, atm	0.2–0.4	2.0–3.5	_
Electric energy equivalent, kwhr/m^3 (1)	4.5	14.0	
Electric consumption, kWh/m^3	1.2–2.0	3.0–4.0	8.5
Total electric energy equivalent, kWh/m^3	5.7–6.5	17–18	8.5

(1) Low pressure steam energy is less costly than the electric energy, about \$0.025 vs \$0.06 per kWh.

on electric energy. Some variations of the VC method utilize steam in the thermocompressor unit to increase energy of the water vapor. The quantity of fuel or electrical energy required for evaporation of a unit mass of water can be calculated based on combustion heat of various fuels and latent heat of vaporization at a given pressure (156,175). In practice the energy to run the distillation desalination process is usually provided either as a low pressure steam discharged from electricity generating turbine or in the form of electrical energy, or both. Table 20.1 shows representative values of energy required by MED, MSF and VC in a conventional systems configuration (155,171).

20.3. Description of hybrid systems

Simple hybrid. In the simple hybrid MSF/RO desalination power process, a seawater RO plant is combined with either a new or existing dual-purpose MSF/power plant to offer some advantages. Several plants currently installed are using some of these advantages. Examples are in Jubail and Madina-Yanbu II in Saudi Arabia and Fujairah in UAE.

Integrated hybrid. The fully integrated MSF/RO desalination power process, which is particularly suitable for new seawater desalting complexes, takes additional advantage of integration features.

Power/water hybrid. Integration of the power and water cycle aims to obtain the optimum cost for both water and power. Important parameters in the design of these systems include:

- seasonal demands for electricity and water
- power-to-water ratio
- minimization of fuel consumption and increase in the power plant efficiency

- minimization of the environmental impact of carbon dioxide including potential consideration of CO_2 tax credit

Some of the earlier analyses in the references showed that when seasonal and daily variations occur; electrically driven technology can provide an excellent choice for hybridization with more conventional dual-purpose plants. The hybrid approach could achieve the lowest cost of total investment, flexibility in production and the lowest cost of power and water production.

Water can be stored while electricity storage is not practical. In this case excess electricity can be diverted to water production incorporating electrically driven seawater reverse osmosis (SWRO) and/or vapor compression (VC) and combined with low pressure steam driven technology of MSF or MED, making it advantageous to design an integrated hybrid plant. One method of making use of idle power capacity is the use of electrically driven RO or VC plants in combination with Desalination Aquifer Storage Recovery (DASR) both for averaging the desalination capacity, for strategic fresh ground water storage or improving quality of the basin.

The increase in the unit size of MSF, MED, VC and RO will lead to reduction of capital costs, but combined with unique application of hybrid ideas will offer reduction in water cost.

Effective integration of membrane/thermal desalination and power technology can reduce the cost of desalination and electrical power production (hybrid desalination). In the early 1980s MSF was the most prevalent seawater desalination process and seawater reverse osmosis was in its early stages of development. Because of RO's development status, two-pass RO was often required in desalination applications treating high salinity seawater feed, common in the Middle East and North Africa (MENA) areas. Early suggestions for hybrid desalination were based upon elimination of the requirement for a second pass to the RO process so that the higher-salinity RO product could be combined with the better quality product from an MSF plant. This is the simplest application of hybrid desalination. Since then, other concepts have been proposed for hybrid desalination. Today although RO can produce product of potable TDS in one pass, blending allows reducing the requirements for second and third partial pass to solve the critical boron issue.

The dual purpose power-desalination plants make use of thermal energy extracted or exhausted from power plants in the form of low pressure steam to provide heat input to thermal desalination plants for MSF or MED distillation processes. The electrical energy can be also effectively used in the electrically driven desalination processes like RO and VC processes (Fig. 20.4).

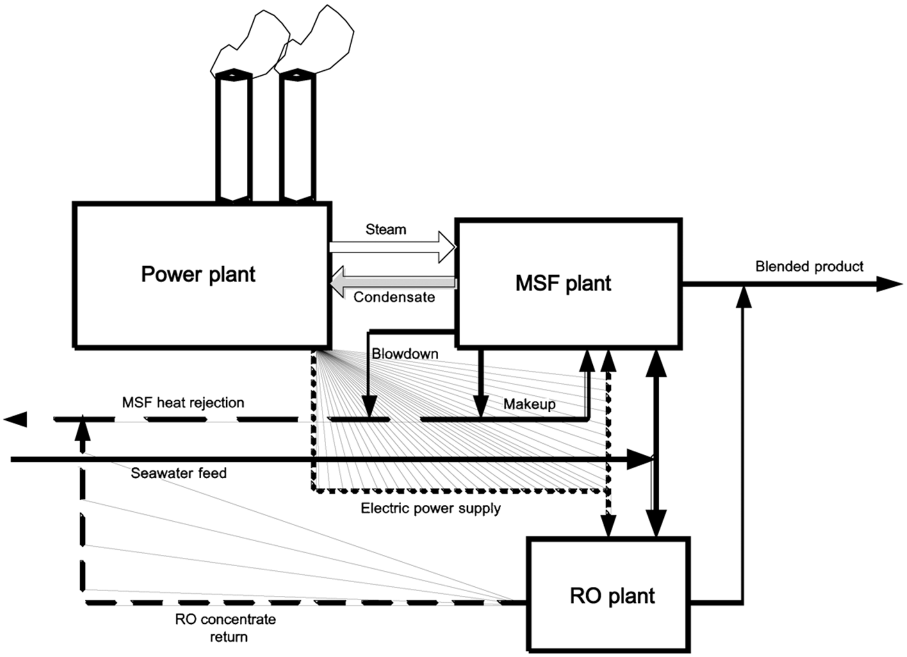


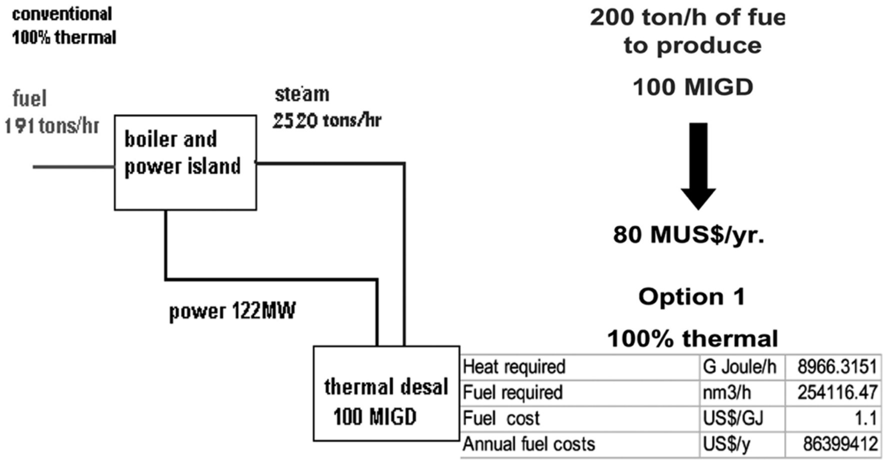
FIG. 20.4 Example of hybrid system components and their relations.

The energy conservation using a hybrid system

In view of dramatic rise in fuel prices in excess of US$ 60/barrel which is equivalent to US$10.3/MMBTU hybrid (RO + distillation) system offers significant saving in fuel cost in comparison with only distillation option (Fig. 20.2). This well demonstrated by simple presentation provided by Dr. Corrado Sommariva in his course on Thermal Desalination Processes and Economics.

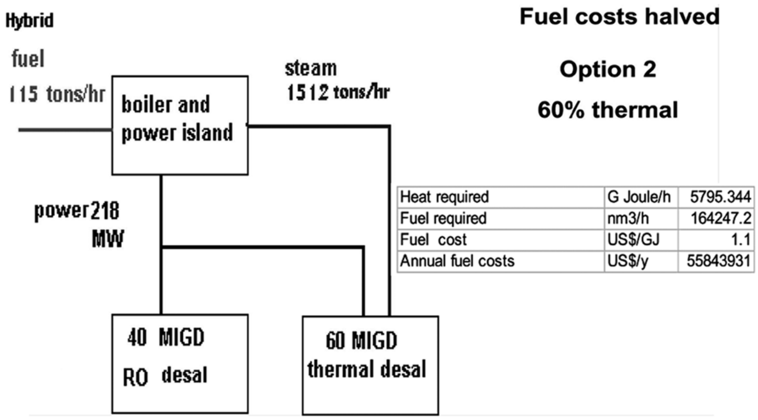
In this case (Fig. 20.5) for 100 mIgd (455,000 m^3/d) MSF desalination and 400 MW of electric power generation plant the annual fuel cost requirement will exceed 86 million US $ based on historic fuel cost of only 1.1 US$/GJ. By comparison a hybrid 100 mIgd (455,000 m^3/d) desalination plant based on 60% thermal and 40 % RO will operate at reduced fuel consumption of only 55 million US$ per year (Fig. 20.6).

This annual fuel cost difference of over 30 million US$ per year is based on 1.1 $/GJ, considering the impact of today's fuel price of 10. $/GJ the annual cost differential will exceed 300 million dollars and will pay back for the total Capex in less then 3 years. Of course in base case we produce more power and to some extent this compensates the additional cost, but this assumes that we need the power.



Heat required	G Joule/h	8966.3151
Fuel required	nm3/h	254116.47
Fuel cost	US$/GJ	1.1
Annual fuel costs	US$/y	86399412

FIG. 20.5 A case study of a hybrid system. The thermal plant configuration. 400 MW power + 100 mIgd: PWR = 4 (courtesy of Dr. Corrado Sommariva)



Heat required	G Joule/h	5795.344
Fuel required	nm3/h	164247.2
Fuel cost	US$/GJ	1.1
Annual fuel costs	US$/y	55843931

FIG. 20.6 A case study of a hybrid system. The thermal + RO plant configuration. Impact of fuel prices at 60$/BBL = 10.3$/MMBTU (courtesy of Dr. Corrado Sommariva)

Awerbuch (195) reviewed impact of high energy and material cost on desalination technology.

There are unique conditions in the Gulf countries where peak demand for electricity rises significantly during summer mainly because of the use of air-conditioning, and than drops dramatically to 30–40% of summer capacity. This creates situation that over 50% of power generation is idled. In contrast, the demand for desalinated water is almost constant. This inequality of demand between electricity and water can be corrected by diverting excess of available electricity to water production incorporating electrical driven technology of SWRO and/or VC and combined with low pressure steam driven technology of MSF or MED, making it advantageous to design an integrated hybrid plant.

20.4. Hybrid—the new alternative

The idea of combining electrical power, MSF, and SWRO has been reported in a number of publications. Initial publications were in the early 1980s. The Hybrid Desalting Systems idea of combining power, MSF distillation plant and a membrane SWRO plant was previously reported to offer significant advantages (157–165, 167, 193).

In the simple hybrid MSF/RO desalination power process, a SWRO plant is combined with either a new or existing dual-purpose MSF/power plant with the following advantages:

- A common, considerably smaller seawater intake can be used.
- Product waters from the RO and MSF plants are blended to obtain suitable water quality.
- Product waters from the RO and MSF plants are blended, therefore allowing higher temperature of distillate.
- A single pass RO process can be used.
- Blending distillation with membrane products reduces strict requirements on boron removal by RO.
- The useful RO membrane life can be extended.
- Excess power production from the desalting complex can be reduced significantly, or power to water ratio can be significantly reduced.

The fully integrated MSF/RO desalination power process which is particularly suitable for new seawater desalting complexes, takes additional advantage of integration features, such as:

- The feedwater temperature to the RO plant is optimized and controlled by using cooling water from the heat-reject section of the MSF/MED or power plant condenser.
- The low-pressure steam from the MSF/MED plant is used to de-aerate or use de-aerated brine as a feedwater to the RO plant to minimize corrosion and reduce residual chlorine.

- Some components of seawater pretreatment process can be integrated.
- One post-treatment system is used for the product water from both plants.
- The brine discharged-reject from the RO plant is combined with the brine recycle in the MSF or is used as a feed to MED.
- The hybridization of nanofiltration as softening membrane process for feed of distillation plants MSF and MED could lead to significant improvement in productivity of desalination plants.

20.4.1. The classic scheme

This is the most common and straightforward hybrid plant scheme (Fig. 20.7). It has been adopted in Jeddah to blend higher TDS RO permeate with distillate from existing MSF plants, and is described in detail by Awerbuch et al. (159, 196) and by many other papers. In general in this scheme part of the MSF plant's heated coolant reject is de-aerated, using low-pressure steam from the MSF plant (to reduce corrosion and residual chlorine), and used as the feed to the SWRO plant. The higher temperature of the feed improves membrane

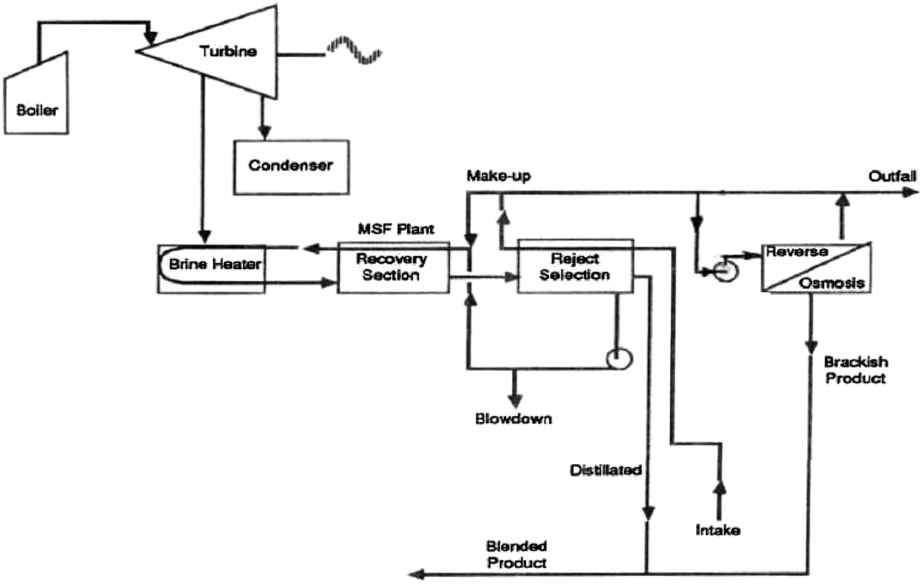


FIG. 20.7 The basic classic scheme hybrid system configuration.

performance (flux, at constant pressure, increases by 1.5–3% for each °C). This is particularly important during the winter, when seawater temperatures can drop to as low as 15°C (59°F). The MSF plant's distillate, at less than 20 ppm TDS, is blended with the SWRO plant's product, making it possible to meet potable water standards for maximum TDS and chloride concentrations with higher SWRO plant product salinity. This, in turn, means that the SWRO plants can be operated at higher conversion ratios, thereby reducing consumption of energy and chemicals and extending membrane useful life.

20.4.2. The classic scheme variant

In one variant of the classic scheme, the SWRO plant's reject brine becomes the feed to the MSF plant, utilizing its high pressure, with a special turbocharger, to boost the MSF plant's recirculation pump (Fig. 20.8). The conversion ratio of the SWRO plant is then limited by the maximum brine recirculation concentration possible. With a once-through MSF plant this limitation is avoided.

20.4.3. The once-through MSF scheme

In this scheme, described by Kamal et al. (161) Al-Sofi et al. (162), Awerbuch et al. (159, 160, 167) and others, a once-through MSF plant is specified, and it's preheated and de-aerated reject, at about 47,000 ppm TDS (with Gulf 42,000 ppm TDS seawater), is used as SWRO plant feed. This scheme has the same advantages as the "classic scheme", but benefits also from the continued de-aeration of the feed by the Non Condensable Gases (NCG) removal system, as the seawater flows through the MSF plant's heat recovery section, and from the reduction of the seawater's bio-fouling potential due to the high temperature sterilization effect at the MSF plant's heat input section.

20.4.4. The duo-cycle ROMED scheme

This is Hornburg's earlier mentioned duo-cycle ROMED hybrid system (180). The main feature of this scheme is the use of a high-GOR TVC (or MVC) plant in lieu of an MSF plant, but the flow scheme is also different from that of the above variant schemes. The seawater is first fed to the SWRO plant, i.e., without preheating and de-aeration in the distillation plant (TVC plants normally do not include feed de-aerators). The SWRO plant's reject is directed,

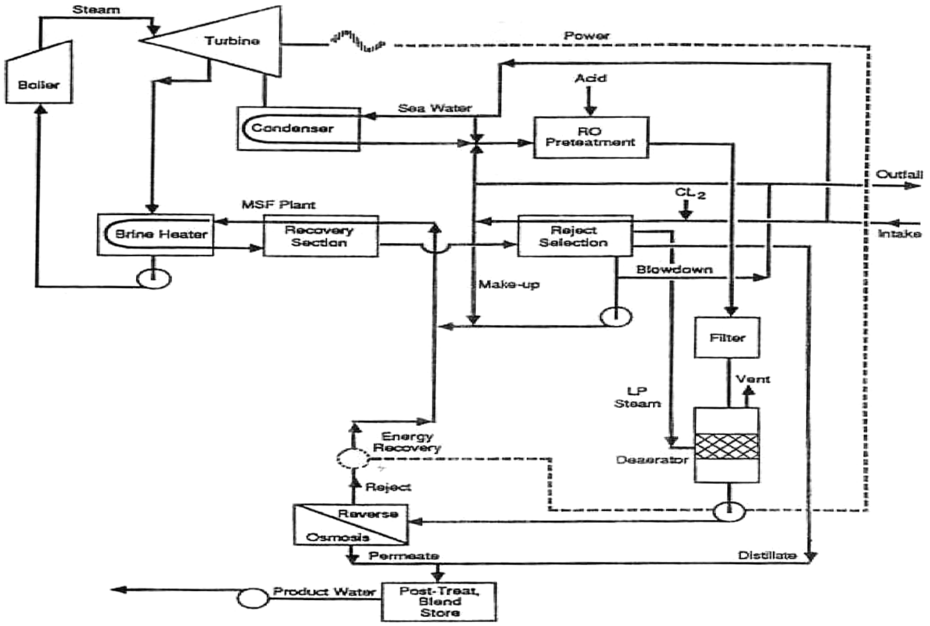


FIG. 20.8 The alternative configuration of "classic scheme" hybrid system.

after passing through an energy recovery turbine, into the TVC plant's heat discharge section, serving as its coolant (TVC plants' heat rejection sections normally utilize falling-film, heat-transfer surfaces, whereas MSF plants utilize pressurized, forced circulation-flow shell and tube condensers). Part or all of this coolant is then used as the feed to the TVC plant's heat recovery section.

20.4.5. The direct-drive steam turbine scheme

The fifth scheme was is the one proposed by Hazen E. Nelson in US Patent 3,632,505 "Evaporation-Reverse Osmosis Water Desalination System" assigned to Stone and Webster Engineering Corporation as early as 1972. It is based on an MSF-SWRO plant combination; with motive steam directed first to back-pressure steam turbines that drive directly the SWRO plant's high-pressure pumps. The steam exhausted from the turbines is then fed to the MSF plant's brine heater. The SWRO plant's brine discharge energy-recovery turbines generate the electric power required for all other pumps and the system's auxiliaries.

20.5. R&D related to improving hybrid systems

The R&D activities pursued today that are most relevant to cogeneration and/or hybrid systems are those relating to the creation of a wider range of nanofiltration and SWRO membranes and the pilot-plant testing and prototype plant designing of low-cost high-GOR high-temperature MED plants. As Awerbuch (167) suggested, an optimal hybrid system would benefit from SWRO membranes with higher fluxes and lower rejections than currently being offered commercially. The minimal accepted membrane rejection will be that which will give permeate with a salinity sufficient to provide, after dilution with the MSF plant's distillate and permeate post-treatment, a combined product salinity of 500 ppm TDS. Some membrane manufacturers have been investigating the potential performance and markets for such high-flux SWRO membranes. The ongoing work on nanofiltration membrane softening technology combined with distillation and hybrid options of NF-MSF-RO or NF-MED-RO offer new potential for improving hybrid systems.

20.6. Quantifying the benefits of the hybrid SWRO/thermal plant scheme in cogeneration stations.

The magnitudes of these potential savings are a function of the relative outputs of the SWRO and distillation plants. They are quantified below for preferred hybrid scheme developed by Hoffman et al. (155)

20.6.1. Savings due to reduced seawater requirements

The use of distillation plant coolant reject as feed to a SWRO plant within selected hybrid plant scheme reduces both seawater supply and brine and coolant rejection requirements vis-à-vis non-hybrid, separate and independent (stand-alone) thermal and SWRO plants.

The cost savings are derived from four sources:

- reduced investment in the seawater intake and supply system
- reduced investment in the brine and cooling water discharge system
- reduced seawater pretreatment costs
- reduced seawater-pumping energy

The actual savings per m^3 of product will depend on the SWRO and distillation plants' output ratio and, since the amount of available coolant reject limits the amount of product that can be generated from SWRO plants without increasing seawater supply, also on the GOR's of the distillation plants. Plants with higher GOR's consume and reject less heat and, for a given coolant temperature rise (normally about 7°C, 12.6°F) requires less coolant flow per m^3 of product. Where seawater supply and pretreatment and brine rejection costs are high (long intake and reject lines, large pumping power requirements, high seawater turbidity, etc.), they add an important cost element to the energy vs. capital cost trade-off equation for deriving the optimal distillation plant GOR.

Using 0.6 exponent for scale-ups and scale-downs Hoffman has computed the savings in intake and outfall system investments and annual operating costs (including pumping and treatment, i.e., chlorination) for a 150,000 m^3/d hybrid plant, for the full range of SWRO to thermal plant outputs ratios. Fig. 20.9 presents these results.

Unit costs (in US ¢/m^3) relate to total intake and discharge costs per 1 m^3 of seawater supply, and are based on a 20-year depreciation period, 7% interest, 8,000 hr/year utilization, seawater pumping power at 0.05 kWh/m^3 min to 0.15 kWh/m^3 max (0.19–0.57 kWh/kgal), coolant and brine discharge by gravity (i.e., no pumping energy), an OPEC country electricity cost of US ¢3.0/kWh

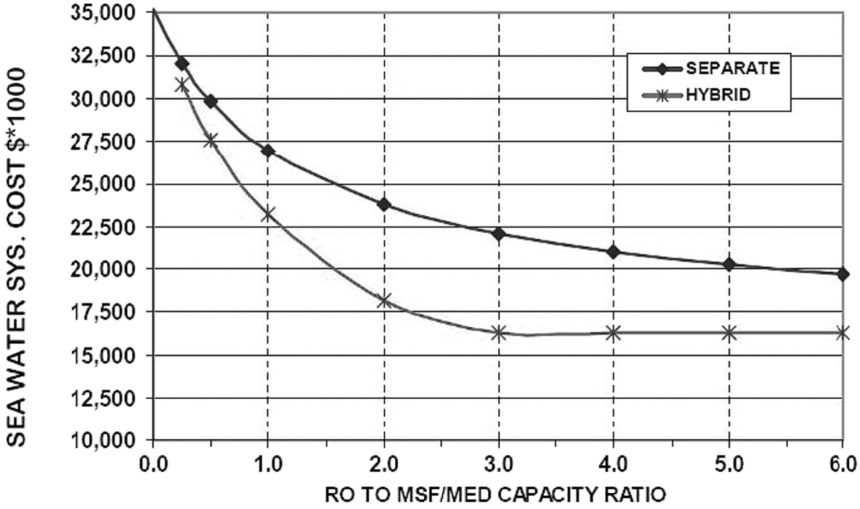


FIG. 20.9 Effect of distillation to RO capacity ratio on combined system cost.

and costs for parts and chemicals (chlorination, etc.) of US ¢0.4–0.7/m^3 (US ¢1.5–2.7/kgal)

20.6.2. RO membrane life

For all membranes, water permeability (i.e., permeate production) declines with operating time while product salinity and chloride concentration increases. The drop in production with time can be compensated by installing extra membrane rack space and installing additional membranes as required. The increase in product salinity cannot be compensated for except with large scale membrane replacement (see discussion on this subject in chapter 10.10).

Therefore, to maintain the product water quality within WHO standards, the designer of stand alone seawater RO plants has the option to replace membranes more frequently or install a two pass (seawater RO and brackish water RO) system.

In the case of hybrid systems (RO + distillation), a single pass RO system can be specified while maintaining a long membrane life. This is made possible by blending the RO product water with the high purity distilled water produced by the thermal desalination unit.

20.6.3. Membrane performance as a function of seawater temperature

The use of all or some of the preheated cooling water discharge from a thermal desalination plant as feed to a SWRO plant enables elevating and controlling the SWRO plant's operating temperature at its optimal or any other higher desired value.

Feed water temperature affects the two main performance characteristics of a membrane: flux and salt rejection. Higher feed water temperatures increase not only flux but also salt passage. Operation at higher temperature may also reduce membrane life (due to membrane compaction), but, as there are no definite *quantitative* figures relating to this effect, we will not include it in our considerations.

For all membranes, water production is a function of temperature, at constant feed pressure. Production will go up with temperature increasing by 1.5% to 3% per degree Celsius for nearly all membranes, thereby enabling reduction of the number of RO membrane modules required for a given permeate capacity. This is of course condition to that feed water quality is sufficiently good that membrane fouling rate will not increase during operation at higher flux.

For the fully integrated hybrid process, the above advantage can be utilized by operating the RO plant at optimum temperature and pressure conditions by using cooling water from the reject section of the MSF plant. El-Sayed et al. (169) conducted pilot study of MSF/RO hybrid systems in Kuwait and observed a significant increase in RO product water flow rate. It was demonstrated on basis of experimental data that 42–48% gain in the product water flow could be achieved for a temperature of 33°C (91.4°F), over that of an isolated RO plant operating at 15°C (59°F) during winter season. The results imply that the energy consumption of RO can be reduced without involving any form of energy recovery, to the level of 5.2 kWh/m^3 (19.7 kWh/kgal) using a simple integration of MSF/RO hybrid arrangement in which the RO plant is fed the preheated seawater rejected from the MSF heat rejection section.

A very interesting study was conducted recently by Nisan et al. (204). It summarizes an investigation on conceptual studies with preheating of feedwater, which is expected to lead to lower specific power consumption, higher water production, thus further reducing the cost of desalination. The results were based on Dow- FilmTec ROSA software and performance of membrane SW 30 HR 380.

The results obtained by the author based on simulation work with the ROSA program are presented in Figs. 20.10–20.15 below for feed TDS values of 28,127, 32,163, 39,086 and 47,400 mg/l. The feed temperature was varied from 10°C to 44°C (50–111°F). The results included in these figures show the variation of the permeate production and recovery ratio as a function of feed temperature at different feed TDS values and at constant design parameters of feed flow, number elements and pressure vessels and at a constant feed pressure.

The design parameters used for calculations of the presented results were:

Feed salinity	28,100–47,400 ppm TDS
Feed temperature	10–44°C (50–111°F)
Recovery rate	29–63%
Feed flow per vessel	10 m^3/h (44 gpm)
Permeate flux	11.7–25.5 l/m^2/h (6.9–15 gfd)
Feed pressure	68 bar (986 psi)

Fig. 20.10 indicates the possibility of increased desalted water production with increased feed water temperature applying constant feed pressure. The rate of capacity increase levels off at the higher end of the temperature range evaluated. The calculation were based on a constant feed flow rate to reflect the usual design conditions of RO pumping and pretreatment equipment. Therefore, higher permeate flows with increased temperature are associated with increased recovery rate (Fig. 20.11).

It is quite obvious that higher recovery can be obtained with lower salinity feed, which has clear process implication when we consider Nanofiltration in front

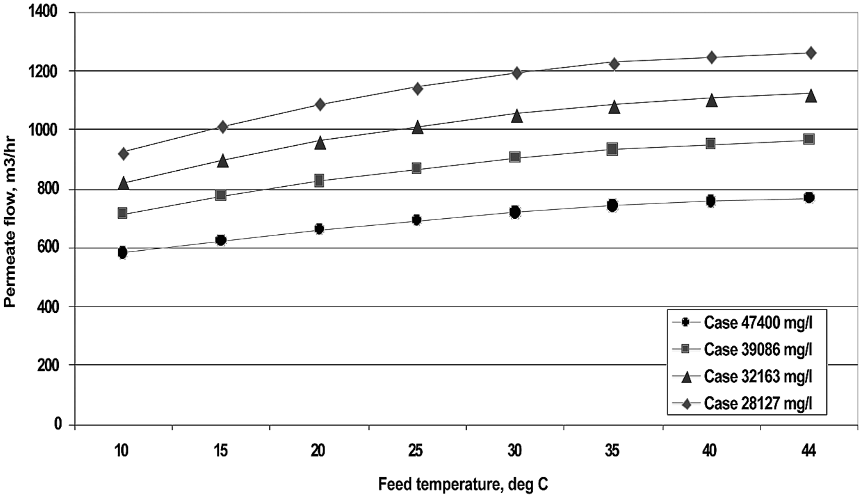


FIG. 20.10 Permeate flow rate (for constant feed rate) as a function of temperature.

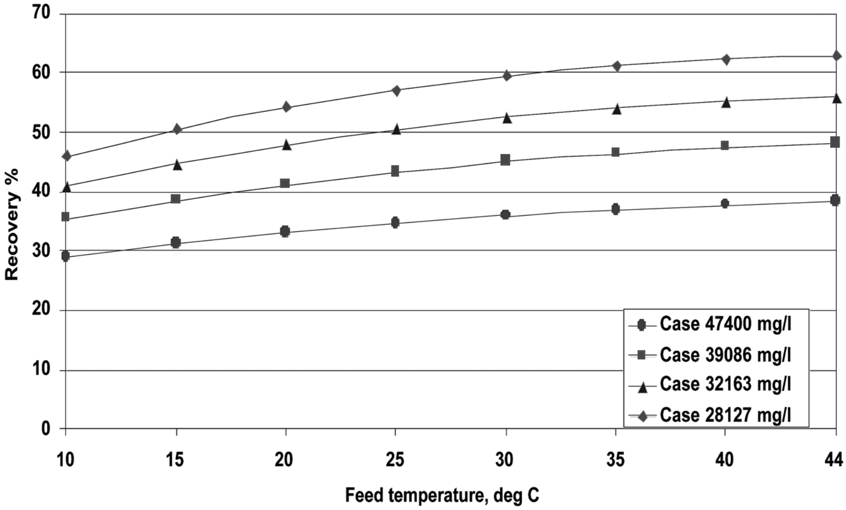


FIG. 20.11 Recovery ratio as function of feed temperature and TDS.

of RO system or use of blending seawater feed with lower salinity water (concentrate of brackish RO for example) to lower the feed salinity to RO system.

Higher membrane permeability at elevated temperature may also result in higher recovery rate. However, higher feed water temperature and recovery rate is associated with an increase of osmotic pressure (Fig. 20.12).

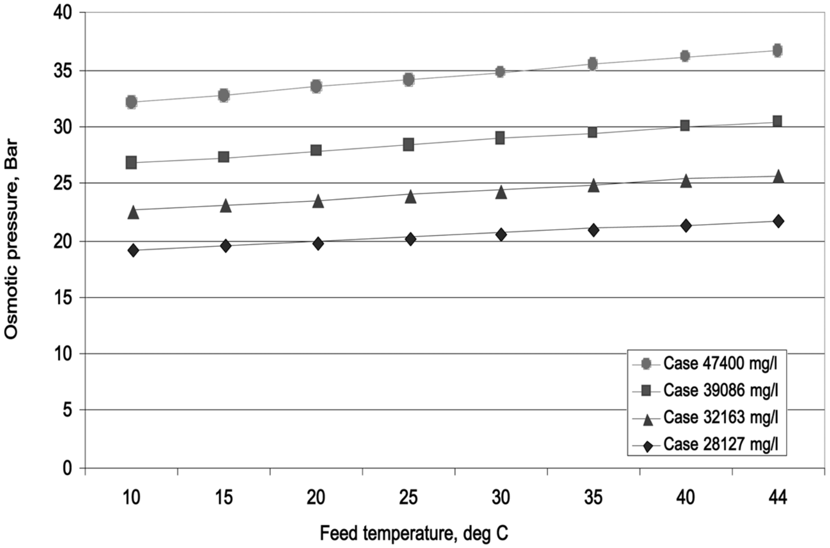


FIG. 20.12 Osmotic pressure as function of feed temperature and TDS.

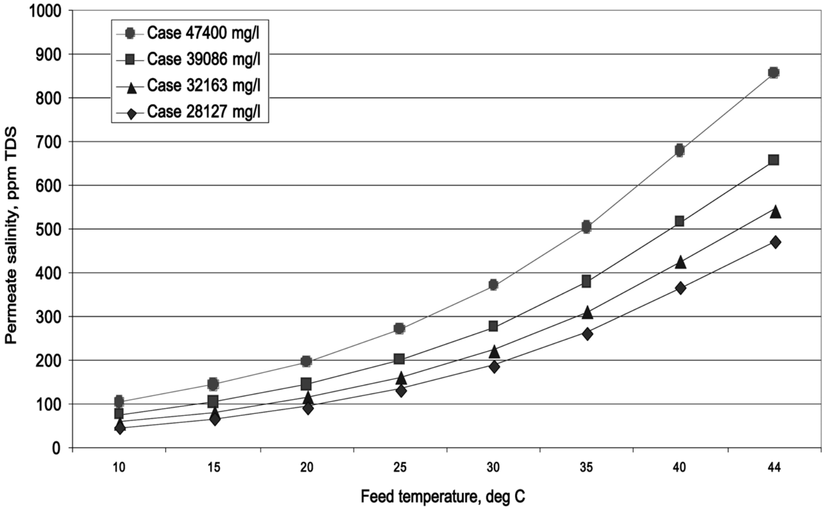


FIG. 20.13 Permeate salinity as a function of feed temperature and TDS.

The permeate TDS systematically increases as the feed temperature and recovery rate are increased (Fig. 20.13). Fortunately, this salinity increase can be easily compensated in hybrid systems (RO + thermal desalination unit) where the ratio of distilled water to membrane permeate can be controlled to achieved required product TDS.

The increase of recovery rate at constant feed pressure at increased temperature in a RO hybrid system leads to reduction of specific power consumption (Fig. 20.14).

A direct consequence is the reduction of the desalination costs with increased feed temperature as shown in the Fig. 20.15. Naturally, for given temperature the desalination cost increases with higher feed TDS. The economics of RO operation was calculated using DOW EVA Elements Value Analysis program. All computer runs included as input values the same feed flow rate, number of elements and feed pressure, which gave good approximation of impact of temperature and feed salinity on a life cycle cost.

The above calculations illustrate the potential for improved economics of operation of RO at elevated feed water temperature in hybrid systems (RO + thermal desalination unit). The full economic benefits of increased membrane permeability can be realized if it would be possible to operate RO membranes at much higher permeate flux rate then it is custom today. Operation at high flux rate will require feed water of high quality. It is very likely that it will require incorporation of membrane pretreatment seawater RO process to achieve sufficiently improved feed water quality.

Some of the critics of higher temperature of operation of RO and NF membranes suggest higher rate of fouling due to increased biological activities, If this is the case an effective method of biological control would have to be de-

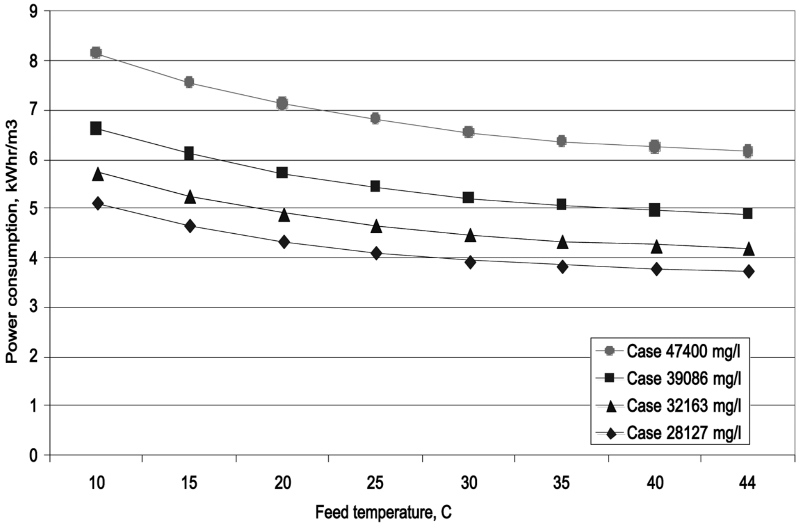


FIG. 20.14 Power consumption as a function of feed temperature and TDS.

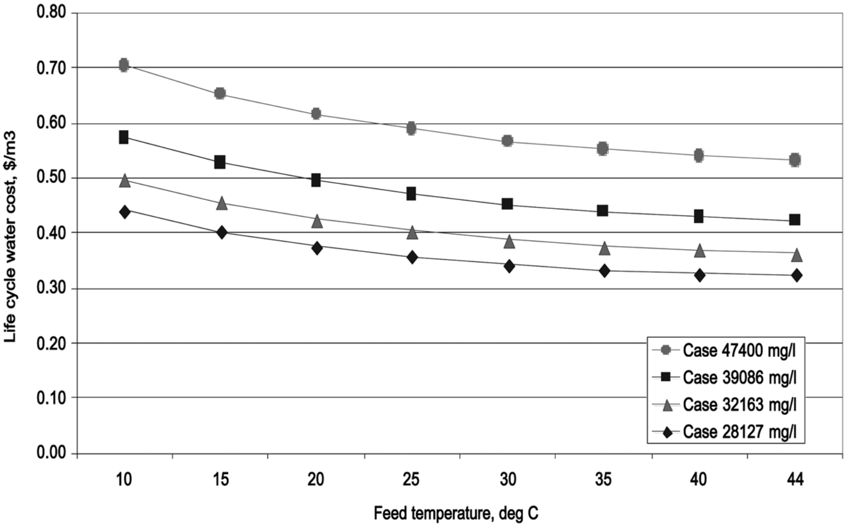


FIG. 20.15 Life cycle water cost as a function of feed temperature and TDS.

veloped for high temperature operation. The increase of seawater temperature which is happening inside the condenser or reject section of the distillation plant is being achieved in a matter of seconds. The assumption is that this rapid rate of temperature increase my act as a thermal shock, possibly reducing biological activity in seawater feed to the membrane unit. Another issue of concern is the compaction of membrane material (permeability decline) during long term operation at high feed pressure and elevated temperature. Both of these issues will have to be tested in field conditions and their effect evaluated against economic benefits of operation of RO unit at elevated temperature in a hybrid system configuration.

20.6.4. Perfomance of nanofiltration membranes as a function of temperature

In nanofiltration systems the increase of temperature of seawater feed could result in higher rate of water permeability increase than it is expected in RO unit. This was one of conclusions of theoretical evaluation work by Agashichev published recently (213). According to author concentration polarization is a significant factor in reduction of available net driving pressure (NDP). In nanofiltration membranes concentration polarization increase with temperature is lower then in RO membranes due to significantly higher salt transport through NF membranes.

In hybrid systems use of Nanofiltration membranes operating also at higher temperatures, due to available heat from power plant condenser or reject section of distillation plants in combination with RO and MSF/MED, has some additional opportunities to reduce desalination costs.

This is shown in the data from the joint research on the new LET NF process conducted by DOW FilmTec and Toray under the direction of LET. As shown in Fig. 20.16 the improvement in productivity is from 2.5 to 3 times at 55°C vs. 25°C (131°F vs 77°F) for specific Nanofiltration membrane SR 90. For other Toray NF membrane the dependence on temperature of operation is shown in Figs. 20.17 and 20.18.

Specifically by using feed comprising variable proportions of softened seawater and water containing a higher concentration of hardness ions than the softened stream, concentration of hardness is sufficiently reduced, thereby allowing a beneficial increase in the TBT of the distillation desalination process. Higher operating temperatures provide an increase in productivity, recovery and performance at lower energy and chemical consumption. As a result, the cost of desalinated water production, including operation and maintenance could be significantly reduced.

20.6.5. Savings due to control of SWRO plant feed temperature

Examining the performance of a commonly used SWRO membrane through its manufacturer's software program (Hydranautics' SWRO System Design

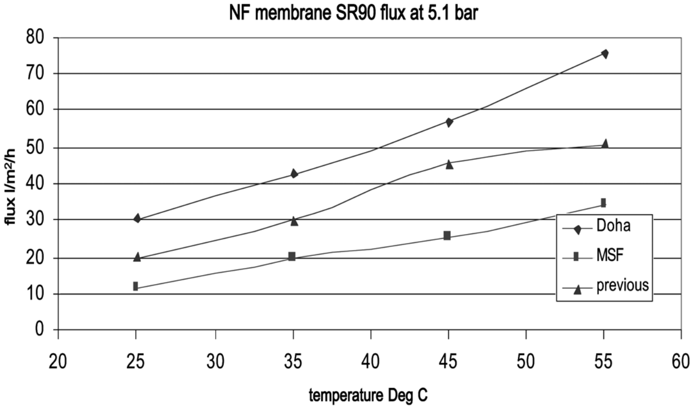


FIG. 20.16 Membrane flux vs temperature at constant feed pressure.

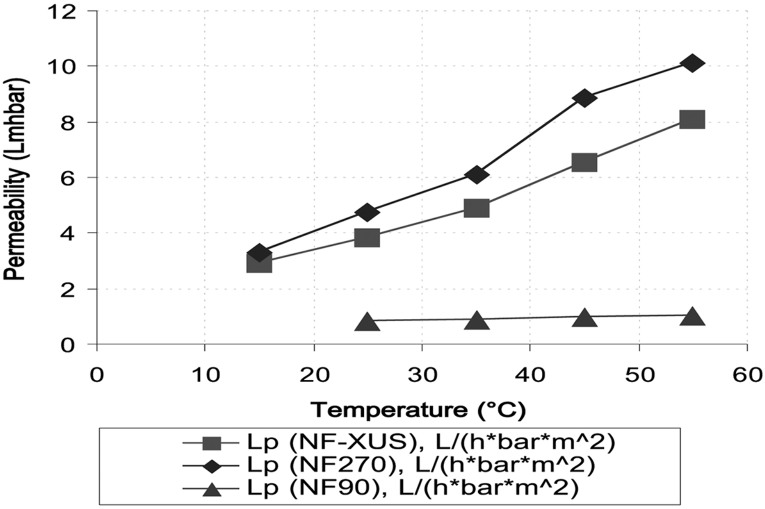


FIG. 20.17 Passage of ions and flux as a function of temperature.

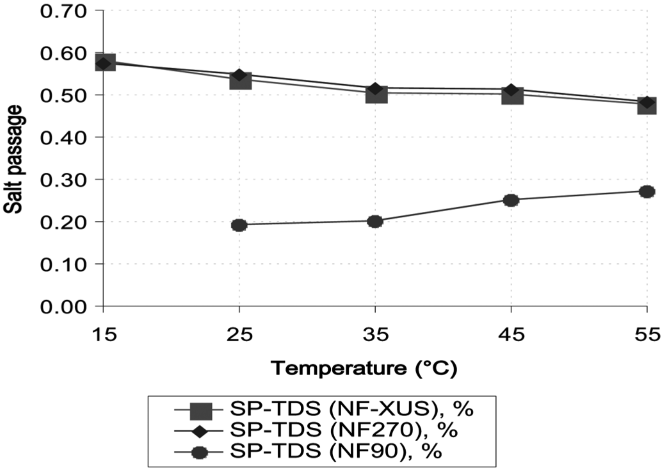


FIG. 20.18 Passage of ions and flux as a function of temperature.

Software, version 7.0, for its SWC-3 seawater membranes), Hoffman (155) found that the following equations have been used to define the relation between temperature and membrane flux and salt rejection: the flux increase factor = 0.015; the temperature increase (in °C) (155); the salinity increase factor = 0.0125; the temperature increase (in °C).

Note that Hoffman elected to base calculations on the above membrane flux rate of increase, 1.5% per 1°C, even though is only about one half the 2.5–3% per 1°C figure specified in some membrane manufacturers' product catalogs and quoted and used repeatedly in many papers promoting hybrid plants.

The feed water temperature elevation in any hybrid plant will be a function of the mix ratio of seawater and reject cooling water forming the feed. This mix, in turn, depends on the amount of cooling water available (i.e., the GOR and design temperature rise in the heat rejection condenser of the distillation plant) and the ratio of the outputs of the SWRO and the thermal plants.

The main results are:

1. Hybrid plants have the potential to increase the average annual membrane permeate flow through increased flux rate and reduce the required membrane surface in the SWRO plants from 10.5%, when only thermal plant cooling water is used as SWRO plant feed, to 4.6%, when the ratio of the outputs of the SWRO and thermal plants is 6:1.
2. The corresponding increases in salt passage and SWRO plant product salinity will range from 4% to 9%.

The US ¢0.6/m^3 (¢2.3/kgal) membrane cost saving figure will be compounded by the savings due to the reduced investment in a range of other items of equipment related to the number of membranes in the plant. These include the membrane pressure vessels, the stainless steel high-pressure connection pipes and fittings, membrane racks, etc. Hoffman estimated the investments in these items as US \$90–100/m^3/d, or about 10% of total plant investment. Reducing the number of membranes by 10.5% will, therefore, reduce total water costs by about another US ¢0.6/m^3 (US ¢2.3/kgal), for a total of US ¢1.2/m^3, (US ¢13.1/kgal), or about 2%.

20.6.6. Savings due to blending SWRO and distillation plants' products

The blending of SWRO and thermal plants' products makes it possible to use the low-salinity (less than 20 ppm TDS) distillation plant product to compensate for higher salinity SWRO plant product. Based on past operating parameters of low recovery rate with current SWRO membranes performance (initial salt rejections of 99.6–99.8%), it is possible to obtain a lower than 500 ppm TDS product water in only one pass operation, even with high-salinity Gulf and Red Sea seawater (rather than with two passes, as required ten years ago, when membrane salt rejections were only 99.2%).However, if the plants are designed to operate at the high conversion ratios used today in most modern SWRO plants, 45–50%, it is projected that product salinity will exceed 500 ppm TDS after about four years of operation, as a result of membrane performance degradation. Membrane replacement costs will then be 25–50% higher than those experienced in plants operating on Atlantic Ocean or Mediterranean Sea seawater, where membrane lifetimes normally exceed the five-year guarantee given by most membrane manufacturers. Hoffman calculated that at a cost of US \$650 per membrane element with a 10 m^3/day (2650 gpd) design output (after three years of operation), the added replacement costs, including labor and idling of equipment costs, are equivalent to about US ¢1.5–2.0/m^3 (¢5.7–7.6/kgal). Note that, at the costs and performance applicable 5–10 years ago these added replacement costs would have been about double.

Alternatively, the plants can be designed for lower recovery ratios, 30–35%, increasing seawater supply and brine disposal, pretreatment, and energy requirements and costs. These will add US \$0.06–0.07/m^3 (\$0.23–0.26/kgal), or about 11%, to SWRO water costs at OPEC fuel costs and US \$0.09–0.10/m^3 (\$0.34–0.38/kgal), or about 14%, at non-OPEC fuel costs. Hoffman elected to examine the savings due to the blending of distillation and SWRO plant products by assuming that the SWRO plant was designed for a conventional 45% conversion ratio, and analyzing the effect on membrane useful life span of increasing the allowed SWRO product salinity through blending with distillation plant product (at different blending ratios).The calculations were made with the assistance of the same Hydranautics SWRO System Design Software, version 7.0, computer program, and the Hydranautics type SWC-3 membrane They also took into account the effect of feed water temperature elevation on membrane performance (i.e., rise in flux and rate of salt passage), and the effect of reduction in flux increase in the plant's operating pressure with time.

In fact, the maximum operating pressure allowed for the selected membrane turned out to be the critical factor limiting membrane lifetime. This limit was 12 years, an extension of seven years to the guaranteed five-year lifetime and eight years above the four-year limit, corresponding to operation without any blending (i.e., the expected lifetime in non-hybrid SWRO plant). Thermal desalination plant product salinity was assumed to be constant, at 20 ppm TDS. The initial salinities of the blended products, for the range of hybrid-plant output ratios examined, 6:1 to 0.25:1, were 351 to 101 ppm TDS. Membrane life was considered terminated when the blended product salinity reached 500 ppm TDS or 12 years (the limiting factor with respect to operating pressure). SWRO plant product salinity at the time of membrane replacements would have ranged from 580 ppm TDS to 980 ppm, depending on the blending ratio, and, had there been no operating pressure limit, could have reached 2,420 ppm TDS for the minimal SWRO to thermal plant ratios.

The membrane replacement cost savings due to the blending of products within a 150,000 m^3/d hybrid plant, within the above range of SWRO and thermal plants' output ratios, are shown at the optimal output ratio of 2:1. The savings in membrane replacement costs in the corresponding 100,000 m^3/d (26.4 MGD) hybrid SWRO plant, compared with its equivalent 100,000 m^3/d non-hybrid plant, are about US $1,172,000 per year, or about US ¢3.6/ m^3 (¢13.6/ kgal). Fig. 20.19 shows the effect of SWRO to thermal plant output ratios and the blending of products on initial product salinity and SWRO plant feed temperature (with new membranes).

20.6.7. Increased recovery ratio

Recovery ratio (conversion) is one of the key RO design parameters. It determines the size of the feedwater handling system (e.g., intake, pretreatment, high pressure pumping) for a given plant size. Higher recoveries decrease the cost of the feedwater handling system and the required electrical and chemical consumption while increasing the initial and replacement costs of the membrane system.

Some of the reasons why higher recovery ratios have not been used in the past are related to the performance characteristics of the membranes and the product water quality specifications. Higher recovery ratio increases required feed pressure due to increase of the average osmotic pressure in the RO system. Also, due to the salt rejection property of available membranes, product water specifications (typically 500 ppm TDS and/or 250 ppm chloride) could not be easily

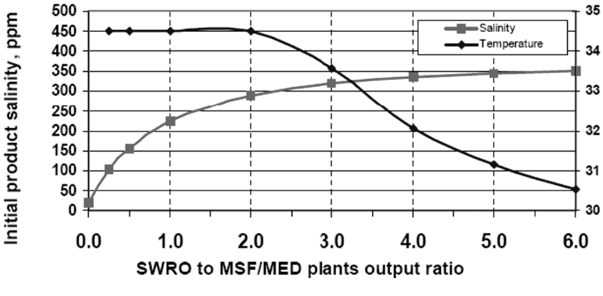


FIG. 20.19 The effect on initial product salinity and feed temperature of the ratio of the outputs of the SWRO and thermal plants and the blending of products.

met at higher recovery ratios. In a hybrid system, higher recovery ratios of RO unit can be incorporated into the plant design. Operation at increased feed water temperature requires lower NDP therefore provides some compensation for increased osmotic pressure. Blending of RO permeate with very low salinity distillate enables attaining the overall product water quality specifications.

20.6.8. Feedwater deaeration

Most aromatic composite membranes require dechlorination of the feedwater as they are very sensitive to even very small concentrations of residual chlorine and/or bromine. If feed water to an RO system is being chlorinated then addition of large quantities of sodium bisulphite is required to reduce free chlorine in the feed water. As an alternative, free chlorine removal can also be accomplished by use of a deareator, followed by significantly reduced quantities of sodium bisulphite.

Deaeration of the feed water also reduces corrosion significantly. In the case of hybrid systems, low pressure steam suitable to operate the deareator is readily available from the MSF plant at low cost. Deaeration can reduce the specification for high pressure piping from SMO-254, SS-317L to lower grades and more economical SS 316L.

20.6.9. Hybrid plant cost savings—summary

A summary of the estimated savings, at OPEC fuel costs and intermediate intake costs, due to the combining of SWRO plants with thermal plants within a 150,000 m^3/d hybrid plant, as a function of the ratio of the outputs of the SWRO and thermal plants is shown in Fig. 20.20. These savings presented graphically on annual and present value (over 20 years) bases. At this point the savings do not take into account the effect of the SWRO plants' power consumption on the economics of water and power cogeneration.

Fig. 20.21 provides a breakdown of the annual savings into capital recovery, energy and treatments, and membrane cost savings components. We see that the optimal ratios of the outputs of the SWRO and thermal plants are in the range of 1.5–3.0, the peak being at a ratio of about 2:1 (100,000 m^3/d SWRO and 50,000 m^3/d thermal).

Hoffman et al. (155) determined that the total savings (for the outputs of both SWRO and thermal plants) at this ratio will be about US ¢4.8/m^3 (¢18.2/kgal) or about US $2.4 million annually. The present value of this annual saving, at

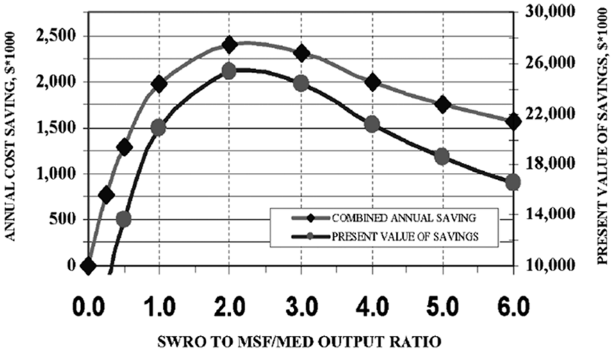


FIG. 20.20 Annual and present-value of savings due to combining SWRO and thermal plants within a 150,000 m^3/d hybrid plant at different ratios of the outputs of the plants.

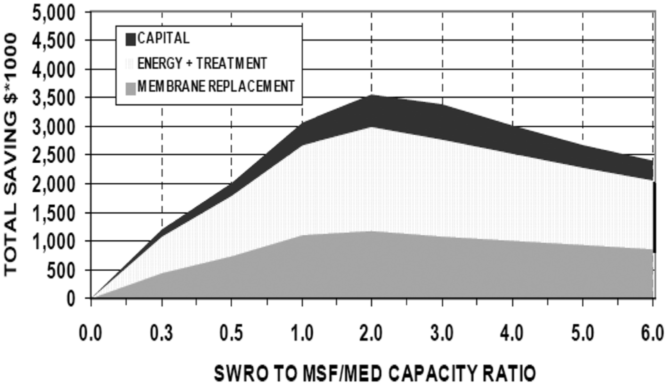


FIG. 20.21 Breakdown of annual savings due to combining SWRO and thermal plants within a 150,000 m^3/d hybrid plant at different ratios of the outputs of the plants.

7% interest and a 20-year depreciation period, is about US $25.3 million. To put these figures into perspective, Hoffman estimated, the costs of desalinated water from an equal-sized 100,000 m^3/d (26.4 MGD) non-hybrid SWRO plant, using the range of conditions and costs applicable to the MENA countries The absolute and relative magnitudes of the savings derived from the hybrid plant scheme, if they are all assigned to the 100,000 m^3/day hybrid SWRO plant's product costs only (i.e., none to the MSF plant's product costs), compared to the costs of our estimated non-hybrid SWRO plant desalinated water shown that the hybrid plant scheme will reduce non-hybrid SWRO plant water costs by as much as US ¢6–9/m^3 ($0.23–0.34/kgal), or 9–16%.

20.7 Examples of existing hybrids

20.7.1. Jeddah hybrid

The results of conceptual and design work (157, 159) led to construction of the simple hybrid project at Jeddah 1, phase I and II plants. The Jeddah 1RO plant is 30 mgd (113,600 m^3/d) combining Phase I which has been operated since 1989 and Phase II has been operated since March 1994. The plant is owned by the Saline Water Conversion Corporation (SWCC), design by Bechtel, constructed by Mitsubishi Heavy Industries, Ltd., under the supervision of SWCC technical committee. Al-Badawi et al. (165) reports the operation and analysis of the plant which utilized Toyobo Hollosep double element type hollow fiber RO modules. The Jeddah complex in addition to 30 mgd RO permeate, produces 80 mgd distillate from Jeddah II, III and IV and 924 MW electricity. Jeddah I RO plant adopted successfully an Intermittent Chlorine Injection method (ICI) in order to prevent membrane degradation by oxidation reaction and biofouling.

20.7.2. Yanbu –Medina hybrid

Objective to minimize power to water ratio lead to construction of Madina and Yanbu Phase II. Nada et al. (168) describes the design features of the largest SWRO plant in the Saudi Kingdom of 130,000 m^3/d (33.8 mgd) in Madina and Yanbu. The plant is able to produce power 164 MW electricity and 288,000 m^3/day (76 mgd) of desalinated water. Two 82 MW back pressure steam turbine (BTG) provides steam to four 36,000 m^3/d (9.5 mgd) MSF distillation units and the electricity to fifteen RO units of 8,500 m^3/day (2.25 mgd), each. Although the plant was not design as an integrated hybrid it provided very good example of significant reduction of the power to water ratio (PWR).

20.7.3. Fujairah hybrid

This seawater desalination and power plant is the largest in the world hybrid configuration of thermal processes and reverse osmosis to be implemented up to now. The paper presented by Ludwig (158) describes in the design considerations for this hybrid plant. The latest excellent description of the Fujairah Hybrid is contained in a paper presented by Doosan (203) describing the design and two years of operation.

The Fujairah plant due to hybridization generates only 500 MW net electricity for export to the grid, and 662 MW gross for water production capacity amounts to 455,000 m^3/d (100 mIgd). Otherwise similar MSF only plant in Shuiwaihat required 1500 MW for the same 455,000 m^3/d (100 mIgd) capacity. The Fujairah desalination plant is split into 284,000 m^3/d (62.5 mIgd) from the thermal part and 170,000 m^3/d (37.5 mIgd) from the membrane process.

The power plant is configured as a combined cycle with supplementary firing. It comprises four gas turbines each rated 109 MW (oil- or gas-fired) and four heat recovery steam generators each of 380 t/hr at steam parameters of 68 bars/537°C that supply the two steam turbine generators. The expanded steam from the turbines serves as process steam for the MSF units.

The Fujairah Project uses gas that is currently imported from the Sultanate of Oman, and will soon be imported from Qatar, when the Dolphin project is completed at a fuel cost of $1.6 per million Btu. At a rather low power-to-water ratio of 500 MW-to-100 mIgd, a hybrid MSF/RO technical solution was extremely attractive for the Fujairah project.

Doosan Heavy Industry and Construction Company were selected as the EPC contractor through an open and competitive bidding process. The main contract was awarded in June 2001. Doosan selected Degremont as a subcontractor to receive the basic design and major equipment supply of the SWRO Plant. During the design stage an extensive pilot plant testing of the RO process was conducted to confirm the performance of the technical solution selected for pretreatment and to determine the impact of dosing of various pretreatment chemicals.

The 100 mIgd (455,000 m^3/d) water productions started on June 31 2003, with a total construction, commissioning, and startup time of less than two years.

The Total Dissolved Solids (TDS) of the product water from MSF units was specified as 25 mg/liter, whereas, that from RO plant was not specified in Request For Proposal (RFP) documents from the client. However, the TDS of potable water after remineralization was specified as less than 200 mg/l. In order to meet the potable water quality of the MSF/RO hybrid process, RO plant should produce desalinated water having less than 180 mg/l of total dissolved solids at the end of fifth year to make blended product water having 60 mg/liter of TDS.

The RO plant is designed as a two-pass system, specifically to obtain the low chloride and TDS contents of the drinking water required for corrosion suppression.

TABLE 20.2
Design parameters of Fujairah hybrid plant (from reference 203)

Description	Data
Design seawater temperature	33.0°C
Design seawater salinity	40,000 ppm
Design ambient temperature	55.0°C
Design relative humidity	100%
Capacity	
Gross power output	660 MW
Net power output	500 MW
Net water output	454,600 m^3/d (100 mIgd)
Net water output (MSF)	284,000 m^3/d (62.5 mIgd)
Net water output (RO)	170,000 m^3/d(37.5 mIgd)
MSF plant	
No. of MSF units	5
Performance ratio	8.0 kg/2326 kJ
Steam supply pressure	3.3 bara
Top brine temperature	109°C
Product water TDS	25 ppm
RO plant	
Pre-treatment type	Dual media filter (gravity), cartridge filter
Number of RO trains	1st pass = 18, 2nd pass = 8
Recovery ratio (first pass/second pass)	43%/90%
Membrane type	Spiral wound, polyamide
Energy recovery type	Energy recovery turbine (pelton wheel)
Brine	Returned to sea through discharge culvert
Desalinated water TDS	180 ppm
Potable water plant	
Post-treatment type (combined)	Lime milk dosing
Potable water TDS	200 ppm
Potable water pH	8.0
Alkalinity increase	Maximum 1.2 mol/ m^3

The seawater desalination processes are designed for seawater TDS of 40,000 ppm and a seawater temperature design range of 22–35°C. Specified for the drinking water product is a maximum TDS of 200 mg/l and its chloride content should not exceed ~85 mg/l.

The blended product from MSF and SWRO is treated in a joint potabilization facility, supplied by CO_2 from the MSF vent gases. To compensate for the conditions that one of the MSF units being taken out of service or for enhanced hardening of the water, the CO_2 demand can be met by an additional CO_2 generation plant.

Seawater intake. The seawater intake is located at 320 meters from the seashore at 6 m above the seabed and 6 m below the surface of the mean sea level. The seawater intake system consists of three submerged pipes 1200 mm diameter, and 500 meters long. Minimum depth at intake point is 9–10 meters. The seawater intake serves RO Plant, MSF Plant and as well as Power Plant. Two of the pipes are dedicated to the MSF Plant for which seawater is chlorinated continuously. The third pipe is allocated to the SWRO Plant only. This to allow intermittent shock chlorination of the seawater used for RO to be carried out rather then continuous chlorine dosing applied to MSF feed. Two of the ten raw seawater pumps are assigned to the RO Plant. Onshore intake facility includes trash racks and traveling bend screens.

For this plant a design decision was made to separate intake for the RO plant, through which the specific chlorination requirements for SWRO can be maintained. It was chosen over the use of a common seawater extraction system. Feeding of preheated cooling water from the MSF reject section to the RO plant was also rejected because, here too, only water that had been chlorinated continuously, and in part shock dosed, was available.

In my opinion this decisions are controversial and in the future more considerations could be given to take clear advantage of common intake and feed temperature control. A study of shock chlorination on top of residual chlorine or de-aeration/dechlorination of RO feed could allow the benefits of hybrid integration.

SWRO plant. The RO Plant consists of two independent identical lines, called Line A and Line B. Each Line includes nine First Pass RO trains and four Second Pass RO trains. The First Pass RO train is designed to produce desalinated permeate water with a TDS of maximum 590 ppm at design condition at the end of fifth year. However, if the permeate water with a TDS of 590 ppm, set as design salinity limit from a single pass of RO plant, is blended with 25 ppm of desalinated water from MSF plant, the specified potable water quality target of 200 ppm could not be accomplished. Therefore, in order to meet the potable water quality requirement, a second pass RO trains has been incorporated ito system configuration. Partial second pass RO system brings the quality of permeate (TDS) to less than 180 ppm at the end of fifth year. Should the required quality (TDS) of potable water were above 300 ppm, which is still far better than WHO recommendation, only single pass of RO plant could have been enough for the hybrid plant. Then, this would have resulted in a more attractive economics of the MSF/RO hybrid water plant. The RO plant has a net permeate capacity rating of 170,475 m^3/d, corresponding to 37.5 mIgd The first pass is design for a recovery rate of 43% and consists of 18 trains, with 17 normally being in operation and one on standby. The second pass that consists of eight trains has a capacity of 74% of the maximum total output of the SWRO, and is designed for a recovery rate of 90%. The blended product from thermal desalination and SWRO is treated in a common potabilization plant. The process employs dosing the permeate with lime slurry and injection of carbon dioxide.

The essential components of the Fujairah RO plant are:

- Seawater intake and pumps
- Pretreatment chemical system
- Flocculators
- Dual media filters
- Filtered water tanks and pumps
- Backwash and dirty wash water system
- Cartridge filters
- First pass H.P pumps with RO rack modules
- Energy recovery turbines
- Second pass pumps and RO rack modules
- Cleaning and flushing facilities
- Product water tanks and pumps

MSF plant. The MSF plant consists of five MSF units, each producing 57,000 m^3/d (12.5 mIgd). The evaporators containing sixteen heat recovery stages and three heat rejection stages. It have been manufactured as a single module in South Korea and transported to the site on a barge. Major components of the MSF plant are:

- Seawater supply pumps
- Evaporators
- Brine recirculation pumps
- Brine blow down pumps
- Distillate pumps
- Vacuum ejectors and condensers
- Chemical dosing equipment
- Ball cleaning equipment
- Acid cleaning system for demisters

When raw seawater temperature is less than 22°C (72°F), hot seawater from heat rejection section of the evaporators can be supplied to the SWRO plant to increase feed water temperature and maintain the guaranteed water production of the RO plant.

The thermal desalination segment of the facility comprises five MSF units each of 57,000 m^3/d (12.5 mIgd) capacity, with a performance ratio of 8 and a top brine temperature (TBT) range of 107–109°C (224.6–228.2°F).

20.7.4. Performance of the Fujairah hybrid plant

Fujairah hybrid performance as reported by Sung W. Woo et al. (203) of Doosan deserves a more detailed review but is briefly summarized.

MSF plant performance. The performance of each MSF unit in terms of distillate flow rate and distillate conductivity is much better than the design and guaranteed values. The Performance Ratios (PR) of MSF units during reliability test periods was decreased to 7.5 for only one day for MSF unit 2 during partial load operation. Except fort this single event the average performance ratio of the MSF units was in the range of 9.1~9.5, which was higher than guarantee value of 8.0 at design condition.

During the reliability and performance test, specific power consumption of MSF plant including potable water plant was about 4.4 kWh/m^3 (16.7 kWh/kgal) of product water which is less than guaranteed value 5.1 kWh/m^3 (19.3 kWh/kgal).

Plant performance (SWRO). The SWRO plant commenced operation on June 31, 2003. The plant has performed satisfactorily, complying with all contract obligations as regards to water quantity and quality in accordance with performance specification defined in tender document.

Performance of pretreatment section

The pretreatment section has performed satisfactorily. During the last year, the Silt Density Index (SDI) remained between 3 and 4, which is much below the SDI limit value of 5.0, as specified by the membrane manufacturers. Backwash frequency of media filters also remained at design frequency, one backwash per 24 hours. The consumption of chemicals was almost similar to the value guaranteed in the contract.

RO membrane performance

Normalized permeate flow rate and salt passage. The normalized permeate product flow rates are higher than the projected initial permeate flow rates and the initial normalized salt passages are less than that of the projected salt passages until the beginning of October 2004. Therefore, since their loading on April 2003, the membranes need not be cleaned nor replaced.

Actual performance data versus projected values. Using data taken on April 29, 2004, a feed water TDS of 37,800 ppm (derived from a raw conductivity of 54,400 µS/cm, feed water temperature: 27°C, 80.6°F), projections were carried out. The actual permeate conductivities are within 3% of the projected values, for all but two of the nine trains in Line A.

For the majority of the trains in Line A, the actual permeate conductivity is lower than the projected values. Projected feed pressure is 69 bars (1000 psi) while actual pressure ranges between 66.7 and 67 bars (967–971 psi).

An additional comparison is also presented on the basis of data taken from on May 24, 2004. This data point was located toward the end of 8-day continuous run. Temperature had increased to 30°C (86°F) and raw seawater conductivity had increased to 56,700 µS/cm, which corresponds to feed TDS of 39,400 ppm, as was derived for the projection. Once again, the comparison with projections indicates good performance for all but two of the trains.

TABLE 20.3
Pretreatment performance during commissioning and as obtained on February 10, 2005 (from reference 202)

Parameter	Projected	Commissioning	February 10, 2005
Seawater SDI	<15	10–20	14–15
Seawater salinity (mg/L)	40,000	39,800–40,050	38,880
Temperature	22–35	28–34.5	23.5
Pretreated water pH	6.5–7.2	6.8–7.0	6.9
Pretreated water SDI	<3.3	2.4–3.0	3.0

All trains showed a trend of improving conductivity with time when operated continuously. Projected pressure is 67.6 bar (972 psi) while the actual pressure ranges between 67 and 67.5 bar (971–972 psi). Based on this comparison of projections with trains in Line B, the membranes are performing as expected, even though the operation of the trains was intermittent and for short periods of time. Performance trend indicates that continuous operation of the trains will produce permeate conductivities equivalent to or below the projected values.

The performance of SWRO membranes is good enough up to now even without chemical cleaning or membranes replacement. Although the conductivity from the two First Pass membranes (trains 1, 2 of Line A) is a little higher than the projection value and other First Pass trains, it is still acceptable for delivery to the distribution system.

Boron concentration was not of RO permeate quality specifications. Therefore, no particular equipment such as pH control or ion exchange bed, etc. has been installed. However, RO plant provided eighty percent (80%) of boron rejection, resulting in 0.7 ppm content in desalinated permeate water. When the permeate water from the SWRO plant was blended with the product water from MSF plant, the boron content in the mixed water was 0.3 ppm, which is less than WHO recommendation (0.5 ppm). In conclusion, the overall membranes performance is good till today.

TABLE 20.4
Permeate conductivity (actual and projected) of all first pass racks, line A (from reference 203)

Train	Conductivity (µS/cm) data taken on April 29, 2004 (27°C)	Conductivity (µS/cm) data taken on May 24, 2004 (30°C)
Projected value	725 (at 27°C)	845 (at 30°C)
1	948	1018
2	912	974
3	769	857
4	698	822
5	706	831
6	686	820
7	736	897
8	616	710
9	677	770

Specific energy consumption. During commissioning, reliability and performance test (RT/PT), the Specific Energy Consumption (SEC) of the RO system including seawater supply pump and product water pump was approximately 4.8 kWh/m^3 (18.2 kWh/kgal) while overall consumption (including pretreatment, air conditioning) was close to 5.0 kWh/m^3 (18.9 kWh/kgal) (158). Comparing to the electric power consumption of the MSF plant, the power consumption of the SWRO plant is only a little higher, by 0.4~0.5 kWh/m^3 (1.5~1.9 kWh/kgal) than that of the MSF plant. The MSF plant uses. In addition to electricity, low pressure steam from the power plant, to heat seawater feed and start the distillation process.

20.7.5. Overall Fujairah conclusions

After one and half year of operation, the Fujairah Hybrid Project has exceeded all performance requirements for Power, MSF and SWRO plant. The designed and guaranteed capacity, 100 mIgd (455,000 m^3/d) of potable water, can be produced with a product water TDS of less than 200 ppm. The electrical power consumption of the overall plant is about 5% less than guaranteed value; allowing net power export of 530 MW rather than 500 MW. SWRO pretreatment has performed beyond expectations. RO membrane performance to date exceeds the projections.

The combined power consumption of the Fujairah hybrid (SWRO + MSF) plant is lower than would be required by an MSF plant of the same capacity.

The possibilty of blending of RO permeate with MSF distillate enables reliable production of potable water of very low salinity in respect of every constituent, including boron.

The SWRO Plant has proved that it is well designed. It is an automatic and easy system to operate, as long as seawater feed quality remains within the observed characteristics. Preliminary full-scale plant results have demonstrated that the foreseen performance and water quality are achievable with the help of experience gained during six month pilot testing during period.

A proper combination of MSF/RO hybrid desalination plant to reduce capital and water cost depends on various parameters such as power-to water ratio, potable water quality, system configuration, etc.

Up to now, the potable water quality (TDS) from MSF plants in Middle East has been specified as less than 150 ppm. However, if the potable water TDS of an MSF/RO hybrid desalination plant is specified to be around 250–300 ppm, which is still quite less than WHO recommendations, then MSF/RO hybrid plant

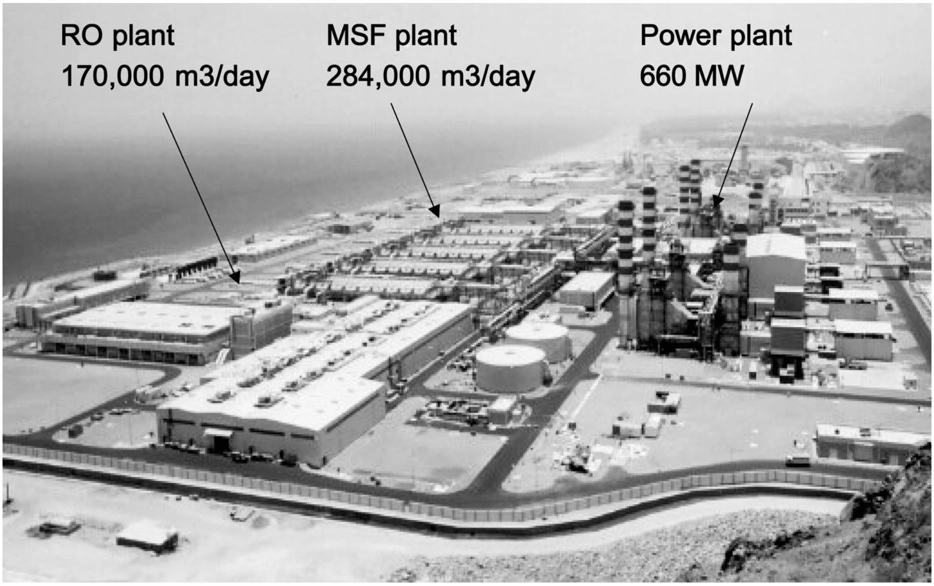


FIG. 20.22 Fujairah hybrid desalination complex, net output: power 500 MW, water 454,000 m^3/d (Courtesy Doosan Engineering). Fujairah plant-MSF area overview.

FIG. 20.23 Arial view of the Fujairah MSF desalination plant.

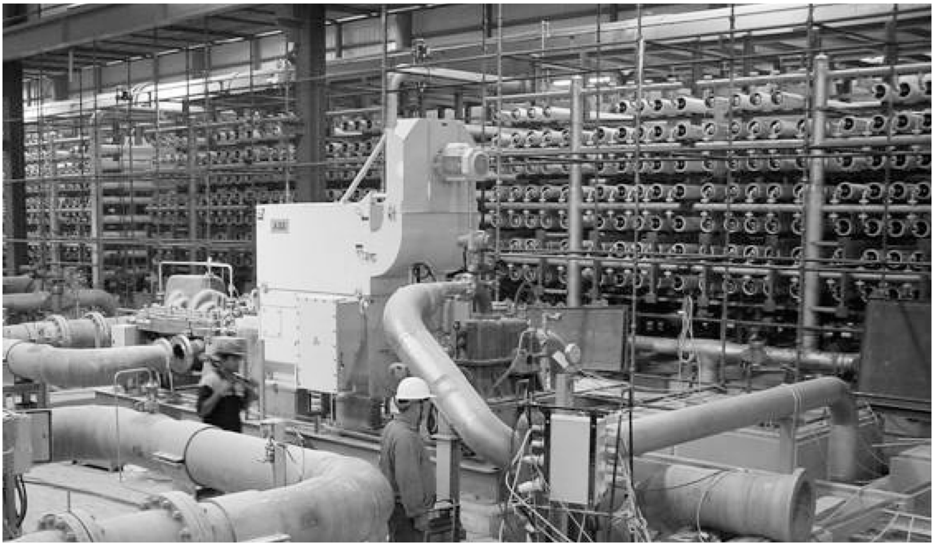


FIG. 20.24 Fujairah plant SWRO racks and feed pump/ER turbine arrangement.

will become much more competitive against MSF plant only, resulting in lower water cost.

20.8 Hybrid variations

As the concepts and applications of hybridization are accepted between distillation processes and RO, we believe that membrane manufactures will develop a new generation of membranes. This new generation of membrane (160, 175, 176) is characterized by a very high specific flux about double the flux of the current generation with small reduction in salt rejection. The current high flux membranes, developed for brackish water desalting demonstrated the ability to significantly reduce the cost of desalting and will be ideal for hybrid plants that includes distillation units.

20.8.1. Hybrid system using multi-effect distillation

Multi-effect distillation (MED) is in our opinion the most important large-scale evaporative process offering significant potential for water cost reduction.

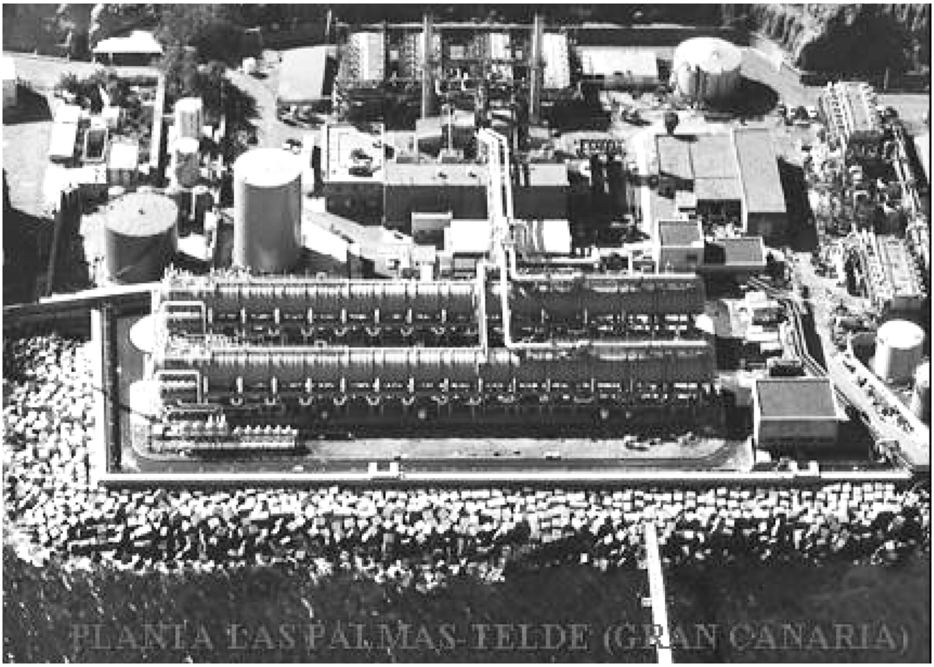


FIG. 20.25 One of more efficient MED plants with performance ratio of 12 in Las Palmas.

The major potential advantage of MED process is the ability to produce significantly higher Performance Ratio (PR) in excess of 15 pounds of the product per pound of steam where MSF practical limits PR to 10. The size of MED units is growing rapidly. In Sharjah SEWA operated for last two years the largest commercial MED units of 22,700–36,4000 m^3/d (5–8 mIgd). Simillar capacity unit is under construction in SEWA Layyah Station, and the design and demonstration module already exist for 45,500 m^3/d (10-migd) unit. MED recently received a lot of attention, as a result of numerous commercial successes of Thermocompression like MED for Al Taweelah A1 a 53 mIgd (240,000 m^3/d) capacity plant.

In general MED capital cost today varies from US$ 1000–1300/m^3-d (US$ 4.5–6.00/igpd) capacity. The future calls for increasing top operating temperature, finding new ways to improve heat transfer performance to reduce heat exchange area, search for an increase in heat transfer performance by tube enhancement, and use of very thin wall in tubular materials. The critical challenge is to adopt Nanofiltration as means to dramatically increase output and increase efficiency of MED plants.

20.9. Hybrid using nanofiltration-membrane softening

Membrane softening technology adapted to hybrid with distillation processes could lead to significant increase in productivity of existing and future distillation plants as well as resulting in better process economics.

Similar to reverse osmosis, nanofiltration (NF) is based on solution-diffusion as major transport mechanism; however, nanofiltration membranes contained fixed (negatively) charged functional groups on the membrane surface. As a result, the selectivity of NF membranes for monovalent and bivalent anions is significantly different as compared to regular RO membranes (172). Specially designed NF membranes have capability of high rejection for divalent ions (Ca, Mg and SO_4), while allowing relatively high passage of monovalent ions (Cl, Na and K). More details discussion about properties of nanofiltration membranes is included in Chapter 16. Nanofiltration technology and applications.

20.9.1 Nanofiltration hybrid background

The basic idea of use of ion selective membranes as a presoftening process for seawater distillation goes back to early publication in 1980 by Wensley et.al. (209) and Furukawa (172). Today pioneering work on Nanofiltration membrane NF softening technology as applied to desalination processes and specifically to seawater desalination is under active development by two groups the Leading Edge Technologies Ltd (LET) based on granted patents Awerbuch (211) and the Saline Water Conversion Corporation (SWCC) of Saudi Arabia based on Hassan patent (210). Numerous publications described the concept Awerbuch (167, 194, 199 and 211) and SWCC published extensively the results on tests of NF at the Research Desalination Center at Jubail and the plant at Umm Lujj, Hassan, Sofi et.al. (188, 190, 191), The latest status of both NF Technologies are described in the proceedings of IDA World Conference in Singapore 2005 Awerbuch (206) and Hamad et al. (207). The LET and SWCC have two different solutions but both are based on effective use of Nanofiltration softening membranes to increase efficiency of desalination process.

In case of LET the basic claim is that:

1. An improved desalination process to produce potable water which comprises:
    (a) passing a first stream of water containing a high concentration of hardness ions through an ion selective membrane to form a softened water having a reduced content of hardness ions;

(b) blending the softened water with a second stream of water containing a higher concentration of hardness ions than the softened water to form a feed to a desalination system;

(c) introducing the feed to the desalination system to form a water product of potable quality, wherein the improvement comprises the introduction of a feed of variable proportions of the softened and second stream of water to the desalination system to increase the top operating temperature of the system and increase recovery of potable water.

The LET invention of partial softening of the stream feeding desalination processes sufficient to achieve reduction in scaling potential can be directed to both to thermal processes like MSF, MED and VC and membrane processes like RO and as well as is an improvement on hybrid system. The inventions comprises the operation of ion selective membrane at variable pressure as a function of the cost of electricity, use of waste or reject heat to improve fluxes and soften only variable portion of the stream to be able to increase the operating temperature and recovery.

The scaling of seawater concentrate or recycle brine occurs due to inverse solubility of calcium sulphate at higher temperatures. At higher operating temperatures and high recovery or concentration factor the stable crystal form is Anhydrite and Hemihydrate (Fig. 20.26). In order to take advantage of higher productivity of distillation plants, through operation at higher temperature, we need to reduce calcium hardness and/or sulfate ions concentration in the feed water.

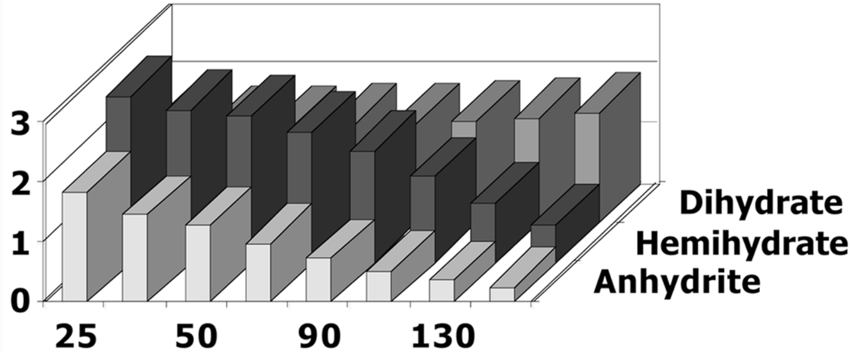


FIG. 20.26 Calcium sulphate solubility limits vs. temperature.

20.9.2. Design Experience with Nanofiltration Hybrid for MSF.

The great potential of nanofiltration membrane softening technology was brought to focus by recent award by Sharjah Electricity and Water Authority (SEWA) to Besix Leading Edge Water Technologies for the first commercial LET Nanofiltration System to increase capacity of existing MSF plant from nominal 22,7000 m^3/d to 32,800 m^3/d (5 mIgd to 7.2 mIgd). This over 40% increase in capacity of MSF unit was a result of a two year demonstration and simulation program developed jointly with SEWA.

The data analysis and modeling of the Test Data provided extremely valuable information allowing improvements in operations as well the development of an integrated program for the optimization of the power-desalination plant.

The results demonstrated that the output of the existing plant described by the Sommariva et al. (187, 188) was increased from the designed capacity of 1,010.5 t/h at 105°C (221°F), or the designed capacity of 1,044.4 t/h at 110°C (230°F) to an output of 1253 t/h. This is equivalent of raising output from 5.33 mIgd to 6.61 mIgd, a 24% increase in plant output without any major modifications having been made to the plant. The maximum production of 1,260 t/h, equal to 6.65 mIgd, was achieved when the TBT was increased to 117°C (242.6°F) with conductivity of product at 454 S/cm^2 (Fig. 20.27).

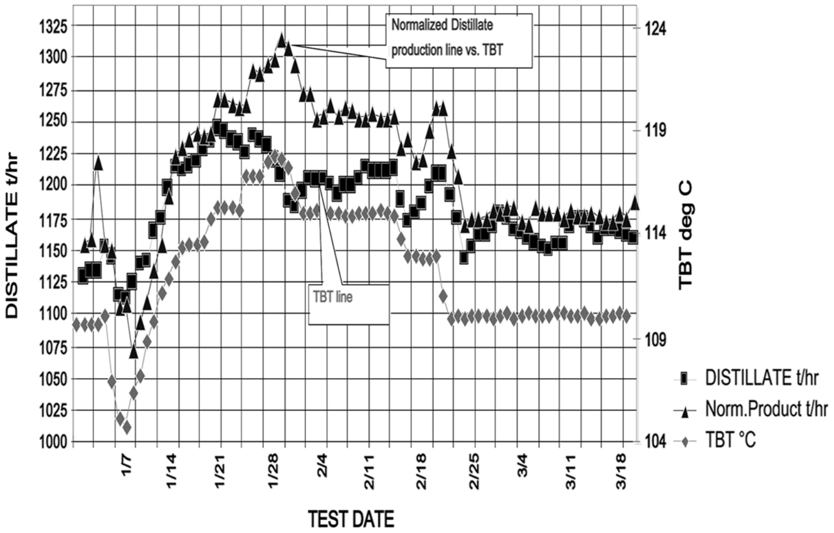


FIG. 20.27 Distillate production as a function of TBT at the Sharjah plant.

This was the first time anywhere that a commercial MSF plant using chemical additives was operated at these TBT temperatures. At these elevated temperatures the major concern is scale formation of calcium sulfate. Due to the simulated conditions of the LET NF System, no fouling of hard scale or soft scale was encountered at these elevated temperatures, and the plant operated in a reliable fashion throughout the test period. In fact the subsequent analysis of the critical Fouling Factor (FF) in the brine heater indicated a decline. This reduction in the FF was possible due to an increased dosing of chemical anti-scalant and was stabilized applying a Tapproge procedure (on-line continuous mechanical cleaning system utilising specially engineered sponge rubber balls which are cycled with seawater through the condenser tubes.

These very good FF results were achieved not withstanding the fact that during the runs the recycle brine concentration was higher then specified by LET. The on-line acid cleaning which removed soft carbonate scale and brought the MSF to higher production than before the test demonstrated that there was no build up of hard scale and the lower FF implied that also there was no build-up of soft scale during high temperature runs. Not withstanding these good FF results the team developed additional means to protect the MSF plants from scale.

Any MSF plant will produce more output with an increased flashing range (defined as TBT minus blowdown temperature), or with an increased recycle flow or both. The analysis clearly demonstrated that with achievable increased flashing range and brine recycle flow (normalized) it is possible to produce 1,309 t/h of distillate at 118°C (244°F).

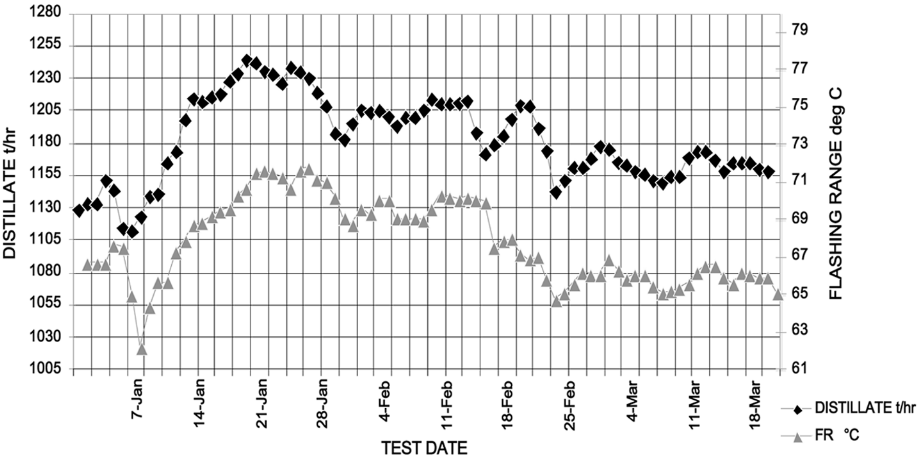


FIG. 20.28 Distillate production as a function of flash range at the Sharjah plant.

One of the main constraints was the increased conductivity in the last stage during the periods of highest temperatures, which forced a reduction in the flashing range, and therefore, reduced the maximum output. The data analysis identified many reasons why the last stage vapor velocity exceeded significantly the normal design. The data analysis identified excessive flashing of the make-up in the deareator, stripping steam flow from stage 13, and heat transfer from upper, high temperature, stages to the last stage, due to the "double-deck" construction of the plant.

While certain features of the plant need to be adjusted to further increase and maximize the plant output, in response to higher operating temperatures and increased product volumes, no major technical issues were encountered that could prevent the application of the LET technology. Once the plant has been suitably modified and upgraded in accordance with mutually agreed recommendations the MSF plant will operate reliably at maximum output (182).

The additional capacity (Fig. 20.29) is achieved without building new intake structure or new power plant in a very limited space which would not allow construction of new desalination plant. The system involves construction of NF plant to provide partial membrane softening of feed to MSF as well as modifications to existing MSF plant to be capable to achieve the increased capacity.

The concentration of sulfate and calcium ions determines in the distillation process the top temperature and concentration factor. Even partial elimination of calcium and sulfate from the feed will dramatically improve the performance of distillation plants. By increasing top temperature from current 95–110°C to 120–125°C would increase water production from existing MSF plants by 25% to 45%. The partial removal of sulfate and calcium ions from the feed has a multiplying positive effect on reduction of scale potential. With the current high quality materials of construction the negative corrosion effects of higher temperature would be minimal.

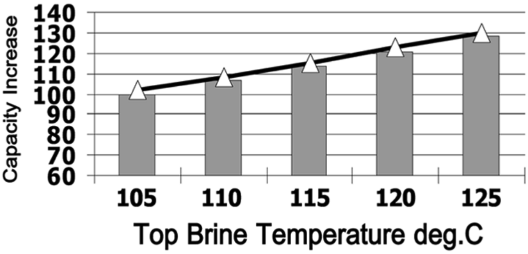


FIG. 20.29 MSF capacity increase vs. TBT. The NF system will substantially increase water production from MSF plants.

20.9.3 Design and construction of the commercial let NF plant.

The construction of first commercial nanofiltartion Hybrid plant with existing MSF awarded to Besix Leading Edge Water Technologies JV is now completed and the final performance testing is conducted in Sharjah Emirate showed boosting of the MSF output to over 7.5 MIPG of desalted water (183).

- Projected benefits of NF—MSF plant at Sharjah:
- Increase the capacity of the existing MSF plant by 44%, from 22,700 m^3/d to 32,800 m^3/d (5 mIgd to 7.2 mIgd)
- Minimum footprint (site has no room new additional plants)
- Reduction of operating cost
- No change to existing intake structure
- No increase of power facilities
- Reduction of capital cost for additional capacity by 40%

The additional capacity is achieved without building new intake structure or new power plant in a very limited space which would not allow construction of new desalination plant.

The main features. The softening process is based on nanofiltration technology, apart from an optimized hydraulic operation; the implementation of the technology allows the thermal units to be safely operated at an increased Top Brine Temperature (TBT) thus allowing to substantially increasing the potable water production.

The plant incorporates the following features:

1. A blending system for hot and cold seawater to keep the feed water temperature in the right range. The blending facilities are located at MSF plant. After blending, the water is pumped to the nanofiltration plant.
2. The raw seawater needs to be pre-treated to avoid fouling and clogging of spiral wound NF membrane elements. Therefore, the water is first pre-treated by means of sand filtration. In order to enhance the efficiency of this pre-treatment, before it enters the sand-filters, the water is pre-treated by means of pH control and coagulation/flocculation.
3. After sand filtration, the water passes a cartridge filter system which acts a final barrier to retain water contaminants. This filter system is to

be considered as a guard cartridge filter system which "boosts" the feed water quality after the sand filters and which protects the membranes in case the efficiency of the sand-filtration units is reduced.

4. Prior to the injection of the pre-treated seawater in the membrane, the water is conditioned in order to maximize membrane life time and in order to reduce the risk of bio fouling and scaling. The treatment incorporates the injection of SBS to remove free-chlorine, the shock dosing of biocide to control bacteriological growth and the continuous, on-line dosing of anti-scalant.
5. Water then passes a two stage nanofiltration membrane system. In order to pass the water through the membranes, medium pressure pumps are used. After the membranes, seawater is partially discharged as permeate—which is the softened water—and as concentrate—this is rejected and pumped back to the sea.
6. Each membrane system is subject to clogging. This clogging can be caused by biological fouling as well as through scaling phenomena. When this "clogging" reaches a certain level, the system pressures will reach their maximum operational values. Membranes need than to be cleaned by means of different chemicals. This is executed "in situ" (after which the system can be taken back into service.
7. Due to the specific site conditions at the power plant, the gravity discharge of the rejected concentrate and drains is not possible. Therefore, all waters are collected in a pump-pit below grade level after which the water is pumped to the existing outfall culvert.
8. After leaving the membranes, the softened water is discharged to an intermediate storage tank. From this tank, the water is pumped at a controlled flow to MSF where is injected in the deareator and/or in the hot-well. The storage tank offers a spare capacity of approx. 1–1/2 hour which allows TBT of MSF to be reduced when softened water feed is interrupted due to failure of the nanofiltration plant.

The feed water source for the NF system is tied in to the seawater piping of the existing MSF unit from where it is pumped to the location of the NF facilities. In order to obtain the most optimum feed water temperature for the NF membranes, "cold"—and "hot" seawater can be "mixed" before the NF supply pump to achieved constant temperature to NF around the year. Maximum temperature of the water actually entering the membrane should be in range and not exceed 38–40°C with today available membranes.

In order to optimize the performance of the pre-treatment and the NF-membranes, the feed water flow is pre-conditioned by chemicals. This main purpose is to obtain better SDI values after the pre-treatment and to control the pH range of the feed water to optimize the water chemistry with respect to membrane scaling and softened water production.

pH control is obtained through the in-line addition of HCl. Better SDI values are obtained by enhancing the filterability of the feed water flow through the addition of chemicals which favor floc-formation.

All chemicals are dosed in line and mixed with the feed water by an in-line static mixer prior to the sand filters. There are 8 sand filters which can be operated as a single stage unit or as a dual stage unit. The sand filters are pressurized and are of the dual media type which means that two filter media are used (filtration sand 0.45 mm and hydro-anthracite) in the same vessel.

The cartridge filter serves as safety filter prior to the main booster pump and the membrane system. The system incorporates 12 separate pressure vessels which contain each a cartridge filter with high filtration efficiency. Water can only be directed to this cartridge filter if the sand filtration units work properly and the water after the sand filters meets the quality requirements (SDI <4).

NF-membrane system. The system is of the two stage design and incorporates the following features:

1. *One main-booster pumps* to pressurize the feed water prior to injection in the membrane system. System pressure: 12.5 to 17.5 bar (181–254 psi).
2. *First stage NF membrane treatment*—80 pressure vessels with 480 membranes. The first stage is split-up in two identical skids (arrays), which 40 pressure vessels each and 240 membranes each. During the membrane filtration process, the feed water is split-up into two flows: the permeate (the softened water) which passes the membrane and the concentrate which did not pass the membrane and is "rejected." The permeate from the first stage is collected and flows to the product water tank, the concentrate serves as feed water for the second stage
3. *One intermediate booster pump* which re-pressurizes the concentrate of the first stage (= feed water the second stage) prior to injection in the second stage. System pressure: 15–20 bar (218–290 psi).

4. *Second stage NF membrane treatment*—40 pressure vessels with 240 membranes. The second stage incorporates one skid. During the membrane filtration process, the feed water (which is actually the concentrate of the first stage) is split-up into two flows: the permeate (the softened water) which passes the membrane and the concentrate which did not pass the membrane and is "rejected." The permeate from the second stage is collected and collected in a product water tank; the concentrate is discharged to the outfall pit from where it is returned to the sea.

The two stage design allows to obtain a high recovery rate (recovery = ratio between useful softened water output and total feed water flow to the membrane system). The recovery rate of this system is approx. 70%.

When the NF plant is into operation, the softened water storage tank is full and no alarms from the NF plant are communicated back to the MSF automation system, the output of MSF 9 can be gradually increased until an output a potable water of 7.2 mIgd is reached. Throughout this process, plant data should be monitored (including the plant fouling factor)

At first, total NF permeate flow (softened water) will be directed to MSF 9. The water will be injected by preference prior to the de-aerator. However, flooding of the de-aerator should be avoided. Flow can be directed to the de-aerator up to the point of flooding, the remaining softened water flow to be directed to the hot well. This will be tested prior to increasing the TBT.

TBT shall not be increased by more than 2°C (3.6°F) at a time and should never exceed 121°C (250°F). When a capacity of 32,800 m^3/d (7.2 mIgd) is

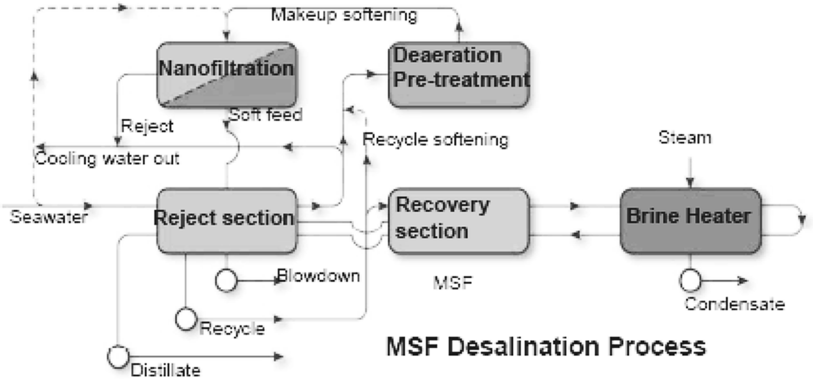


FIG. 20.30 Configuration of LET NF system for treatment of MSF makeup water.

FIG. 20.31 NF trains and media filtration units.

reached, TBT should not be further increased, even if TBT at that moment is lower than 121°C. The Control system is arrange in such a way that MSF plant can return safely to lower temperature of operation below 110°C (230°F).

20.9.4 Nanofiltration hybrid variations

There are many potential variants for NF hybridization with NF-MSF-RO as well as NF-MED-RO. Below are a few examples developed by Bechtel-LET and proposed for large scale implementation.

The case above (Fig. 20.32) is the basic case of NF system previously described as a SEWA project. The following schematics (Fig. 20.33) shows a combination of preheated feed being softened and fed to MSF and RO based on optimum split between to desalination processes to achieve the lowest product cost.

In the final scheme (Fig. 20.34) the seawater is preheated in MSF reject section, then is softened by nanofiltartion membrane, follow by SWRO. The reject brine of SWRO has significantly reduced level of scaling ions sulfate, calcium and magnesium and therefore the reject brine can be the feed for distillation plant.

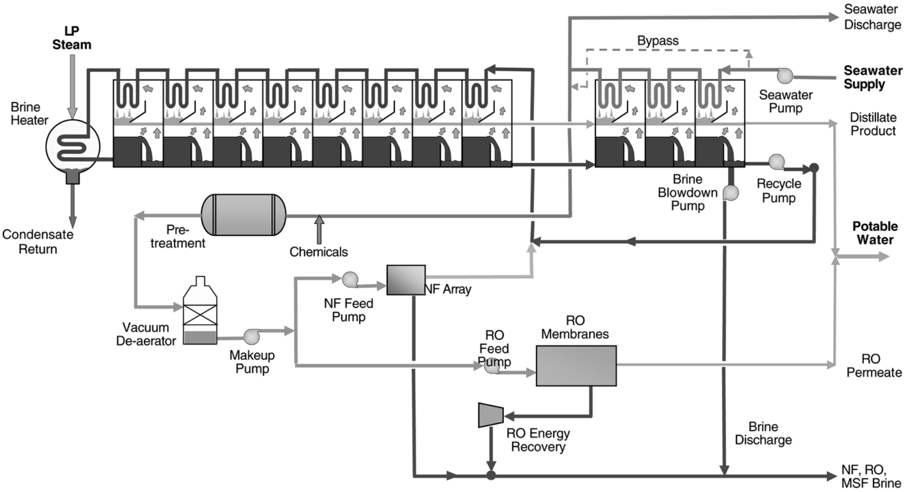


FIG. 20.32 Hybrid is with NF prior to MSF.

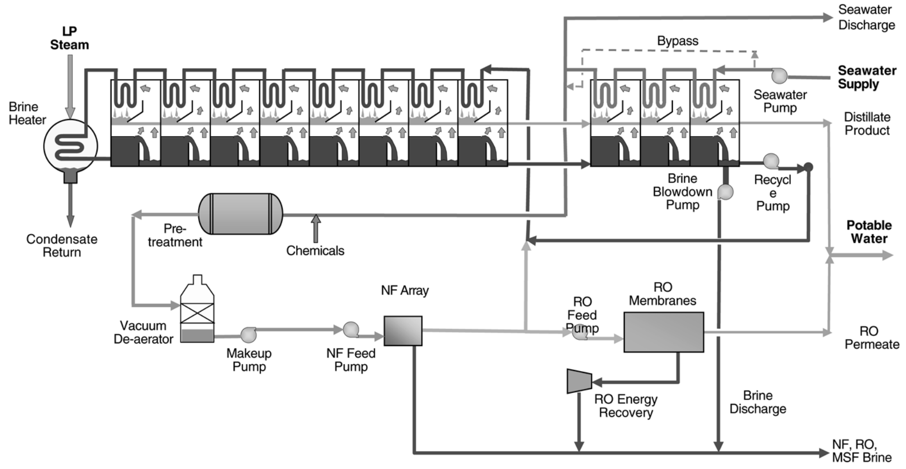


FIG. 20.33 The hybrid solution with NF prior to MSF/RO.

N F membrane softening technology could significantly improve operation and reduce the cost of the MED process, specifically when applied to MED processes using advanced heat transfer surfaces like double fluted tubes, by eliminating the risk of scaling and fouling. NF technology will permit increase in the top temperature resulting in significant increase in output and performance ratio.

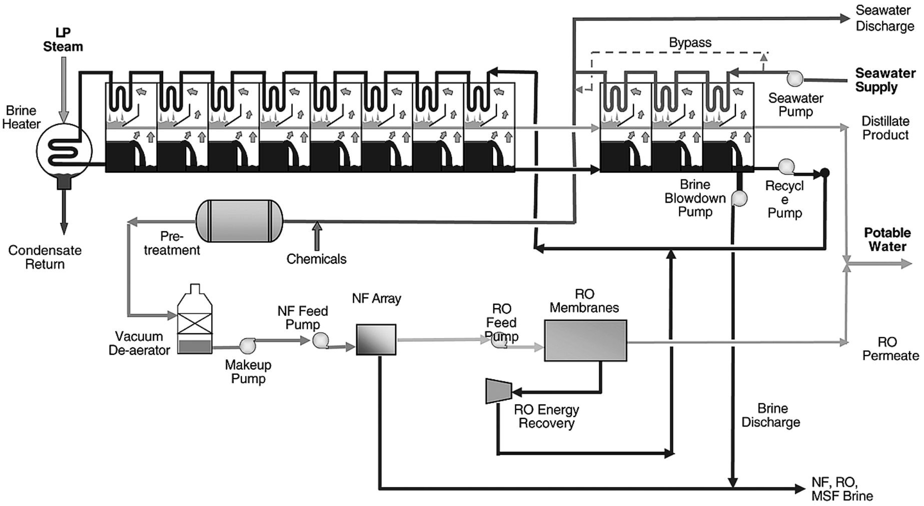


FIG. 20.34 The hybrid with NF and RO reject feeding MSF.

The latest references to use of nanofiltration to hybrid system can be found in papers presented by Awerbuch (206) and as a trihybrid NF/RO/MSF configuration by Hamad et al. (207).

20.10 Dual purpose facilities and potential for hybridization

The importance of power to water ratio and potential for hybridization becomes more relevant, after sorting out the comparative advantages of currently available power and water technologies. Many proposed hybrid ideas Al-Sofi et al. (184), Hornburg et al. (185) and Gluckstern et al. (186) In this presentation we will omit nuclear options for power generation.

Given that the water and power requirements of real utility varies seasonally by a significant amount, the optimal choice for power desalting plant is not obvious when operation extends beyond base load.

It is interesting to note, that the more efficient is the base load operation for generating electricity, the less effective is production of water and power in peaking and intermediate modes. The most advance combined cycle desalination plant has very high PWR, choice of which would provide a significant surplus of unused power capacity in winter time. An excellent analysis of the optimal power and water plants design and demand analysis was reported by Lenox et al. (179), Awerbuch (178), Sommariva at al. (198) and Kamal (181).

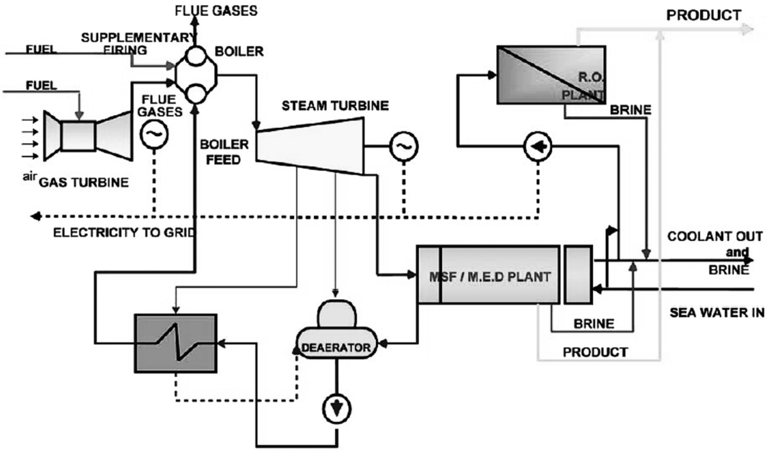


FIG. 20.35 Typical combined cycle hybrid distillation and RO system.

TABLE 20.5
Typical power to water ratios for different desalination technologies.

Technology	PWR (ratio of MW electricity required per IMGD water production)
Steam turbine	
BTG-MSF	5.0
EST-MED	7.0
EST-MSF	10.0
Gas turbine	
GT-HRSG-MED	6.0
GT-HRSF-MSF	8.0
Combined cycle	
BTG-MED	10.0
BTG-MSF	16.0
EST-MED	12.0
EST-MSF	19.0
Reverse osmosis	0.8–1.5

The above data show that electrically driven desalination processes like RO and VC, clearly require minimum investment in the power plant. At the same time where seasonal and daily variations occur, electrically driven technology can provide an excellent choice to be hybridized with more conventional dual purpose plants. The hybrid approach could achieve the lowest cost of total investment, flexibility in production and the lowest cost of power and water production.

20.11 Hybrid systems using vapor compression distillation

The Vapor Compression distillation (VC) technology offers unique potential. Today power/MSF/MED/RO plants can be hybridized with VC to take advantage of increase distillation output, using electrically driven technology.

Currently the largest scale unit of VC is 3000 m^3/d capacity or 0.8 mgd in a single unit, which consists of three evaporator-condenser effects couple to a single high volumetric compressor.

This large scale VC guarantee unit specific electricity consumption of 7.5–8.5 kWh/m^3 (28.4–32.2 kWh/kgal) of product. (excluding sea water supply). They produce high purity 10–20 ppm distillate at high plant availability of 94–96%.

In future vapor compression distillation units, will grow in capacity and number of effects. A single-effect 2.5 IMGD (11,000 t/d) vapor compression system with a conventional axial flow compressor and with an unconventional radial inflow compressor of a novel design was described by Yehia M. El-Sayed (170).

Design of VC over four and more effects, staging compressor in series or parallel will allow effective hybridization of power with MED, MSF and RO. This particularly will be important in cases where power to water ratio has to be minimized in favor of water production.

20.12. Hybrid systems using MSF-MED

In distillation processes there is no interaction between MSF and MED energy process streams. Substantial efficiency improvements could be obtained if process streams between MSF and MED are exchanged in order to take advantage of the different operating temperature conditions of each plant.

In particular due to the low MED operating temperature (61–67°C, 142–153°F) this process could be thermally driven by process streams properly sourced by an adjacent MSF plant.

A number of novel technology options (LET–Mott McDonald patent pending) that have been studied and their possible implementation in a real scale plant should be available soon.

The objective of the MSF-MED hybrid is to increase energy efficiency, distillate production and minimize operational costs. Results of such hybridization combined also with RO and NF is well described Sommariva et al. (208).

20.13. Hybrid systems and Desalination Aquifer Storage Recovery (DASR)

Cost-effective integration of three proven technologies, desalination, power and aquifer storage recovery (ASR) can secure a reliable, sustainable and high-quality fresh water supply for the Gulf States. LET pioneered in the Middle East the concept of strategic and economic storage and recovery of desalinated water (DASR) and waste water (WASR) to the security of its communities. The idea is covered in many papers (177, 197, 200–202).

The seasonal surplus of unused idle power could be used by electrically driven desalination technologies RO and Hybrid Systems including NF/RO/MSF process in combination with ASR creating a system of Desalination/Aquifer Storage and Recovery (DASR). The ability to store and recover large volumes of water can contribute to the average downsizing of power and water facilities with substantial operational cost savings.

DASR provides strategic reserves of potable water, to prevent damage or depletion to existing oasis or aquifers, for controlling salt-water intrusion, or improvement in water quality. DASR is of strategic importance to the Middle East Principle of DASR technology

- Electricity demand drops 30–40% of peak demand during the winter months
- During that period over 50% of power generation plants are idle
- The idle power can be utilized to produce low cost water using hybrid technologies
- Produced water is stored in underground aquifer for summer use

A desalination plant will operate continuously with modulating its output depending on power demand. Typical water storage volumes for desalinated water are limited to providing less than one day of water supply, a highly vulnerable situation.

20.14. Resources conservation and environmental impacts of various hybrid configurations

Resources conservation and environmental impact, too, are aspects that have to be considered when designing hybrid systems The reduced primary (fuel) energy consumption when coupling reverse osmosis with thermal processes in a hybrid configuration amounts to from 30–40% against reference plants (boiler for heat and condensing turbine power plant for electricity). CO_2 emissions from a gas-fired combined cycle plant with a corresponding SWRO share are likewise substantially lower than for a conventional cycle, that is a condensing turbine power plant alone with SWRO for water production. There is a rise in heat dissipated to the atmosphere, but considering environmental impacts this is offset by less pollution emitted in the flue gases from a gas-fired power plant and the consequently reduced need for flue gas cleaning. Showing a substantial reduction is dissipation of heat from a hybrid plant to the sea as compared with the conventional heat cycle/SWRO configuration. In recent years the consideration of carbon dioxide tax will have a significant impact in justifying hybrid plants in the Gulf.

20.15. Hybridization conclusions

Combining thermal and membrane desalination processes and technologies within a single plant or in hybrid plant schemes can *reduce desalinated water costs*, and, as part of dual-purpose stations; *add flexibility* to the combined water and power production and *reduce any existing water and power demand mismatch problems*

It can be seen that applying hybrid solutions will reduce desalinated water costs, compared with non-hybrid schemes, from as little as 2–3% to as much 15%.

In large desalination plants, there should also be little loss of economies of scale due to the use of two or more different processes, in two or more smaller units, in lieu of one large, single-technology plant. Many such plants, at the same site, are based on the same process (MSF), but utilize different designs and have different performance figures. All the solutions whether stand-alone high-GOR plants (LT-MED/TVC, HT-MED) or hybrid schemes (MSF/SWRO, MVC/MED, MVC/TVC, etc.) requires use of the largest size plants available.

The hybrid of power-desalination systems, from its early concept of power–MSF–RO to blend the products and minimize power generation, leads to many new ideas.

- Hybrid of MED-RO has many of the same advantages than the MSF-RO, but has the ability to cut significantly power water (PWR) ratio.

- Hybrid of MSF–MED with VC has the potential of boosting water output through simple or full integration and at the same time reduces power to water (PWR) ratio.
- Hybrid with Nanofiltration–Softening Membrane will provide the ability to increase desalination output of distillation plants MSF and MED, by reducing scaling potential of the feed, increase the top brine temperature and provide significant better concentration factors and recovery for all distillation processes.
- Hybrid with electrically driven desalination technologies RO and VC would allow use off peak power for water production, and minimize power capacity by shutting down RO or VC daily during the peak.
- The seasonal surplus of unused idle power could be used by electrically driven desalination technologies RO and VCR in combination with aquifer storage and recovery to create effective DASR solutions.

All of the above ideas have a goal to maximize and optimize benefits of power and water generation in order to provide lower cost water the "Essence of Life."

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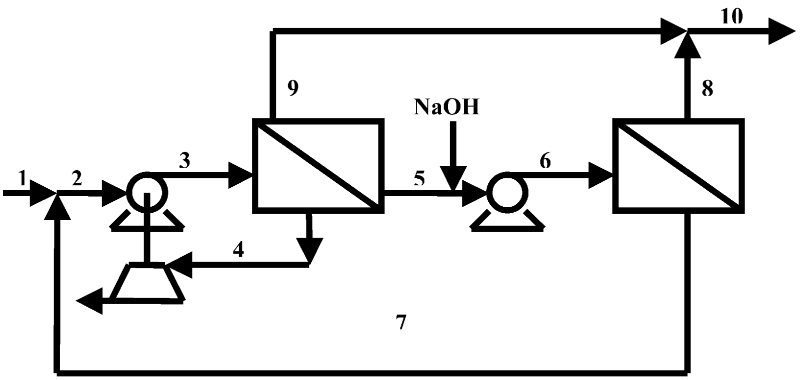
Appendices

Appendix A Example of Membrane unit configurations

Mediterranean feedwater temperature 14°C
Split partial two pass: SWRO + BWRO

Feed water source: Mediterranean seawater	Feed water type: surface intake
System configuration: split partial two pass	Output capacity: 9666 m^3/d (2.55 mgd)
Permeate recovery 1st pass: 50%	Permeate recovery 2nd pass: 90%
Total system recovery: 49.3%	Calculation for membrane age: 3 years (–20% nominal flow, +30% of nominal salt passage)
Feed flow 1st pass: 833.4 m^3/h (3670 gpm)	Feed flow 2nd pass: 138.9 m^3/h (612 gpm)
Feed pressure 1st pass: 69.1 bar (1002 psi)	Feed pressure 2nd pass: 14.7 bar (213 psi)
Concentrate pressure 1st pass: 67.8 bar (924 psi)	Concentrate pressure 2nd pass: 9.7 bar (141 psi)
Elements type 1st pass: SWC5	Elements type 2nd pass: ESPAB
Array 1st pass: 100 PV (8M)	Array 2nd pass: 9: 4 PV (8M)
Average flux 1st pass: 14 l/m^2/h (8.2 gfd)	Average flux 2nd pass: 32.3/m^2/h (19.0 gfd)
Feed pH 1st pass: 8.1	Feed pH 2nd pass: 10

Constituent	Feed, ppm	Permeate, ppm	Concentrate, ppm
Temperature	14°C (57°F)		
pH	8.1	8.1	8.4
Ca	483	0.3	950
Mg	1557	0.9	3061
Na	12200	34.9	23961
K	481	1.7	944
CO_3	5	0.0	11.2
HCO_3	162	0.7	318
SO_4	3186	1.8	6264
Cl	22599	54	44294
F	1.4	0.0	2.7
NO_3			
B	5	0.3	9.3
SiO_2	1.6	0.0	3.3
TDS	40686	95	79929



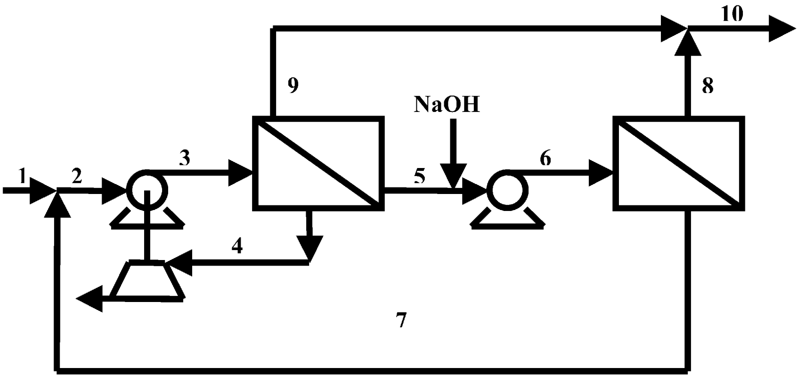
Location	1	2	3	4	5	6	7	8	9	10
Flow, m^3/h (gpm)	820.1 (3611)	833.4 (3670)	833.4 (3670)	416.7 (1835)	138.9 (612)	138.9 (612)	13.9 (61)	125.0 (550)	277.8 (612)	402.8 (1773)
Pressure, bar (psi)			69.1 (1002)	67.8 (983)		14.7 (213)	9.7 (141)			
TDS ppm	40668	40080	40080	79928	432	432	4249	13	131	95

	Pass 1 and total	Pass 2
Feed pressure, bar (psi)	69.1 (1002)	14.7 (213)
Concentrate pressure, bar (psi)	67.8 (983)	9.7 (141)
Pump flow, m^3/h (gpm)	833.4 (3550)	138.9 (612)
Turbine flow, m^3/h (gpm)	416.7 (1835)	
Pump efficiency, %	86	84
Motor efficiency, %	94	94
Energy recovery device efficiency, %	86	
Combined pumping power, kW (hp)	2016.8 (2705)	
Recovered power, kW (hp)	661.7 (887)	
Total power requirement, kW (hp)	1355.1 (1817)	
Pumping energy, kWh/m^3 (kWh/kgal)	3.36 (12.71)	

Mediterranean feedwater temperature 28°C
Split partial two pass: SWRO + BWRO

Feed water source: Mediterranean seawater	Feed water type: surface intake
System configuration: split partial two pass	Output capacity: 9333 m^3/d (2.46 mgd)
Permeate recovery 1st pass: 50%	Permeate recovery 2nd pass: 90%
Total system recovery: 48.3%	Calculation for membrane age: 3 years (–20% nominal flow, +30% of nominal salt passage)
Feed flow 1st pass: 833.4 m^3/h (3670 gpm)	Feed flow 2nd pass: 277.8 m^3/h (1223 gpm)
Feed pressure 1st pass: 65 bar (942 psi)	Feed pressure 2nd pass: 10.6 bar (154 psi)
Concentrate pressure 1st pass: 63.7 bar (924 psi)	Concentrate pressure 2nd pass: 6.2 bar (90 psi)
Elements type 1st pass: SWC5	Elements type 2nd pass: ESPAB
Array 1st pass: 100 PV (8M)	Array 2nd pass: 18: 9 PV (8M)
Average flux 1st pass: 14 l/m^2/h (8.2 gfd)	Average flux 2nd pass: 31.1/m^2/h (18.3 gfd)
Feed pH 1st pass: 8.1	Feed pH 2nd pass: 10

Constituent	Feed, ppm	Permeate, ppm	Concentrate, ppm
Temperature	28°C (82°F)		
pH	8.1	7.9	8.4
Ca	483	0.2	934
Mg	1557	0.5	3010
Na	12200	22.9	23572
K	481	1.2	929
CO_3	5	0.0	11.2
HCO_3	162	0.5	312
SO_4	3186	1.0	6159
Cl	22599	36	43664
F	1.4	0.0	2.7
NO_3			
B	5	0.3	9.3
SiO_2	1.6	0.0	3.3
TDS	40686	62	78615



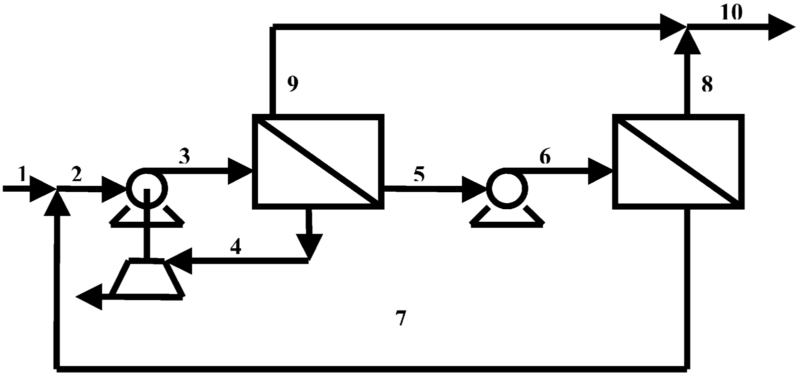
Location	1	2	3	4	5	6	7	8	9	10
Flow, m^3/h (gpm)	806.2 (3550)	833.4 (3670)	833.4 (3670)	416.7 (1835)	277.8 (1223)	277.8 (1223)	27.8 (122)	250.0 (1101)	138.9 (612)	388.9 (1712)
Pressure, bar (psi)			65.0 (947)	63.7 (924)		10.6 (154)	6.2 (90)			
TDS ppm	40668	39489	39489	78616	484	484	4647	28	124	62

	Pass 1 and total	Pass 2
Feed pressure, bar (psi)	65.0 (947)	10.6 (154)
Concentrate pressure, bar (psi)	63.7 (924)	6.2 (90)
Pump flow, m^3/h (gpm)	833.4 (3550)	277.8 (1223)
Turbine flow, m^3/h (gpm)	416.7 (1835)	
Pump efficiency, %	86	84
Motor efficiency, %	94	94
Energy recovery device efficiency, %	86	
Combined pumping power, kW (hp)	1973.7 (2647)	
Recovered power, kW (hp)	621.7 (834)	
Total power requirement, kW (hp)	1352.0 (1813)	
Pumping energy, kWh/m^3 (kWh/kgal)	3.47 (13.13)	

Persian Gulf feedwater temperature 16°C
Split partial two pass: SWRO + BWRO

Feed water source: Persian Gulf seawater	Feed water type: surface intake
System configuration: split partial two pass	Output capacity: 9666 m^3/d (2.55 mgd)
Permeate recovery 1st pass: 42%	Permeate recovery 2nd pass: 90%
Total system recovery: 41.2%	Calculation for membrane age: 3 years (–20% nominal flow, +30% of nominal salt passage)
Feed flow 1st pass: 992.2 m^3/h (4369 gpm)	Feed flow 2nd pass: 69.5 m^3/h (306 gpm)
Feed pressure 1st pass: 67.8 bar (983 psi)	Feed pressure 2nd pass: 12.4 bar (180 psi)
Concentrate pressure 1st pass: 66.1 bar (958 psi)	Concentrate pressure 2nd pass: 7.3 bar (106 psi)
Elements type 1st pass: SWC5	Elements type 2nd pass: ESPA2+
Array 1st pass: 110 PV (8M)	Array 2nd pass: 4: 2 PV (8M)
Average flux 1st pass: 12.7 l/m^2/h (7.5 gfd)	Average flux 2nd pass: 31.9/m^2/h (18.8 gfd)
Feed pH 1st pass: 7.0	Feed pH 2nd pass: 5.6

Constituent	Feed, ppm	Permeate, ppm	Concentrate, ppm
Temperature	16°C (61°F)		
pH	7.0	5.1	7.2
Ca	478	0.5	818
Mg	1672	1.7	2861
Na	14099	68.6	24090
K	530	3.2	905
CO_3	4.2	0.0	0.6
HCO_3	154	1.3	258
SO_4	3314	3.3	5707
Cl	24927	99	42606
F			
NO_3			
B	5	1.0	7.8
SiO_2			
TDS	45199	177	77256



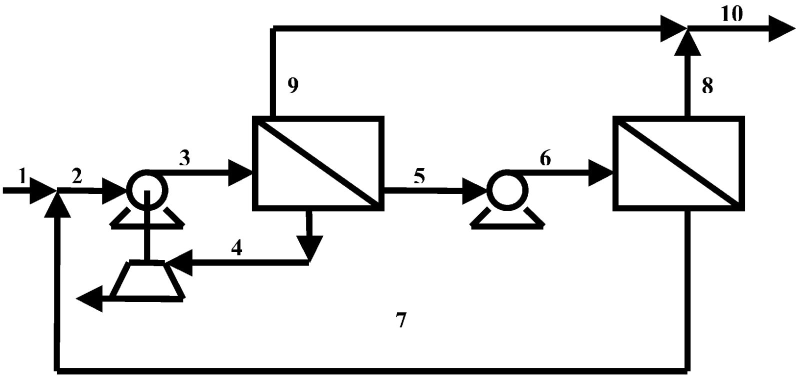
Location	1	2	3	4	5	6	7	8	9	10
Flow, m^3/h (gpm)	985.9 (3441)	992.2 (4369)	992.2 (4369)	575.5 (2534)	69.5 (306)	69.5 (306)	6.9 (30)	62.5 (275)	347.2 (1529)	409.7 (1795)
Pressure, bar (psi)			67.8 (983)	66.1 (958)		12.4 (180)	7.3 (106)			
TDS ppm	45199	44925	44925	77256	629	629	6158	15	206	177

	Pass 1 and total	Pass 2
Feed pressure, bar (psi)	67.8 (983)	12.4 (180)
Concentrate pressure, bar (psi)	66.1 (958)	7.3 (106)
Pump flow, m^3/h (gpm)	992.2 (4369)	69.5 (306)
Turbine flow, m^3/h (gpm)	575.5 (2534)	
Pump efficiency, %	86	84
Motor efficiency, %	94	94
Energy recovery device efficiency, %	86	
Combined pumping power, kW (hp)	2304.1 (3090)	
Recovered power, kW (hp)	890.8 (1195)	
Total power requirement, kW (hp)	1413.3 (1895)	
Pumping energy, kWh/m^3 (kWh/kgal)	3.45 (13.06)	

Persian Gulf feedwater temperature 34°C
Split partial two pass: SWRO + BWRO

Feed water source: Persian Gulf seawater	Feed water type: surface intake
System configuration: split partial two pass	Output capacity: 9666 m^3/d (2.55 mgd)
Permeate recovery 1st pass: 42%	Permeate recovery 2nd pass: 90%
Total system recovery: 41.2%	Calculation for membrane age: 3 years (–20% nominal flow, +30% of nominal salt passage)
Feed flow 1st pass: 992.2 m^3/h (4369 gpm)	Feed flow 2nd pass: 138.9 m^3/h (611 gpm)
Feed pressure 1st pass: 64.4 bar (934 psi)	Feed pressure 2nd pass: 10.7 bar (155 psi)
Concentrate pressure 1st pass: 62.8 bar (911 psi)	Concentrate pressure 2nd pass: 7.1 bar (103 psi)
Elements type 1st pass: SWC5	Elements type 2nd pass: ESPA2+
Array 1st pass: 110 PV (8M)	Array 2nd pass: 8: 4 PV (8M)
Average flux 1st pass: 12.7 l/m^2/h (7.5 gfd)	Average flux 2nd pass: 31.9/m^2/h (18.8 gfd)
Feed pH 1st pass: 7.0	Feed pH 2nd pass: 5.8

Constituent	Feed, ppm	Permeate, ppm	Concentrate, ppm
Temperature	34°C (93°F)		
pH	7.0	5.2	7.2
Ca	478	0.5	812
Mg	1672	1.6	2841
Na	14099	72	23919
K	530	3.5	899
CO_3	4.2	0.0	0.9
HCO_3	154	1.3	265
SO_4	3314	3.3	5660
Cl	24927	106	42303
F			
NO_3			
B	5	1.9	7.2
SiO_2			
TDS	45199	190	76707



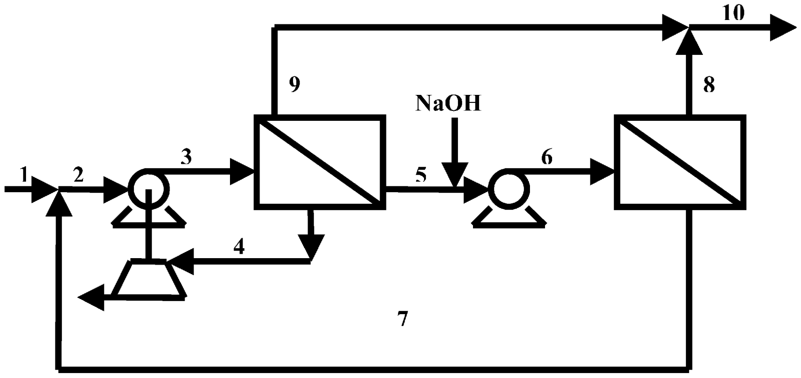
Location	1	2	3	4	5	6	7	8	9	10
Flow, m^3/h (gpm)	979.0 (4308)	992.2 (4369)	992.2 (4369)	575.5 (2534)	138.9 (611)	138.9 (611)	13.9 (61)	125 (550)	277.8 (1222)	402.8 (1772)
Pressure, bar (psi)			64.4 (934)	62.8 (911)		10.7 (155)	7.1 (103)			
TDS ppm	45199	44695	44695	76707	977	977	9153	68	245	190

	Pass 1 and total	Pass 2
Feed pressure, bar (psi)	64.4 (934)	10.7 (155)
Concentrate pressure, bar (psi)	62.8 (911)	7.1(103)
Pump flow, m^3/h (gpm)	992.2 (4369)	138.9 (611)
Turbine flow, m^3/h (gpm)	575.5 (2534)	
Pump efficiency, %	86	84
Motor efficiency, %	94	94
Energy recovery device efficiency, %	86	
Combined pumping power, kW (hp)	2195.3 (2941)	
Recovered power, kW (hp)	846.4 (1135)	
Total power requirement, kW (hp)	1348.9 (1809)	
Pumping energy, kWh/m^3 (kWh/kgal)	3.35 (12.68)	

Red Sea feedwater temperature 16°C
Split partial two pass SWRO + BWRO

Feed water source: Red sea seawater	Feed water type: surface intake
System configuration: split partial two pass	Output capacity: 9666 m^3/d (2.55 mgd)
Permeate recovery 1st pass: 47%	Permeate recovery 2nd pass: 90%
Total system recovery: 45.9%	Calculation for membrane age: 3 years (–20% nominal flow, +30% of nominal salt passage)
Feed flow 1st pass: 886.6 m^3/h (3904gpm)	Feed flow 2nd pass: 185.2 m^3/h (815 gpm)
Feed pressure 1st pass: 68.7 bar (996 psi)	Feed pressure 2nd pass: 13.6 bar (197 psi)
Concentrate pressure 1st pass: 67.2 bar (977 psi)	Concentrate pressure 2nd pass: 8.7 bar (126 psi)
Elements type 1st pass: SWC5	Elements type 2nd pass: ESPA2+
Array 1st pass: 100 PV (8M)	Array 2nd pass: 12: 6 PV (8M)
Average flux 1st pass: 14.0 l/m^2/h (8.2 gfd)	Average flux 2nd pass: 31.1/m^2/h (18.3 gfd)
Feed pH 1st pass: 8.2	Feed pH 2nd pass: 10

Constituent	Feed, ppm	Permeate, ppm	Concentrate, ppm
Temperature	16°C (61°F)		
pH	7.8	7.6	8.0
Ca	500	0.2	923
Mg	1540	0.7	2844
Na	13300	31	24546
K	490	1.4	904
CO_3	2.3	0.0	4.4
HCO_3	126.8	0.5	238
SO_4	3240	1.5	5995
Cl	23180	45	42786
F			
NO_3			
B	5.3	0.4	9.5
SiO_2			
TDS	42389	80	78252



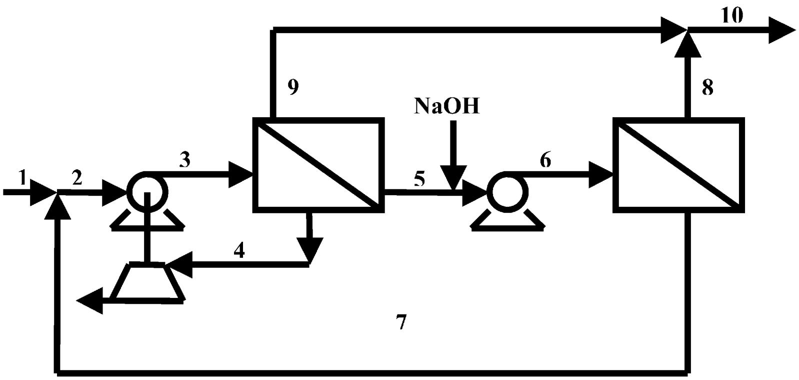
Location	1	2	3	4	5	6	7	8	9	10
Flow, m^3/h (gpm)	868.7 (3825)	886.6 (3904)	886.6 (3904)	469.9 (2069)	185.2 (815)	185.2 (815)	18.5 (81)	166.7 (734)	231.5 (1019)	398.2 (1753)
Pressure, bar (psi)			68.7 (996)	67.2 (977)		13.6 (197)	8.7 (126)			
TDS ppm	42389	41591	41591	78252	399	399	3941	13	128	80

	Pass 1 and total	Pass 2
Feed pressure, bar (psi)	68.7 (996)	13.6 (197)
Concentrate pressure, bar (psi)	67.2 (977)	8.7 (126)
Pump flow, m^3/h (gpm)	886.6 (3904)	185.2 (815)
Turbine flow, m^3/h (gpm)	469.9 (2069)	
Pump efficiency, %	86	84
Motor efficiency, %	94	94
Energy recovery device efficiency, %	86	
Combined pumping power, kW (hp)	2147.7 (2880)	
Recovered power, kW (hp)	739.5 (992)	
Total power requirement, kW (hp)	1408.2 (1888)	
Pumping energy, kWh/m^3 (kWh/kgal)	3.53 (13.36)	

Red Sea feedwater temperature 26°C
Split partial two pass SWRO + BWRO

Feed water source: Red Sea seawater	Feed water type: surface intake
System configuration: split partial two pass	Output capacity: 9666 m^3/d (2.55 mgd)
Permeate recovery 1st pass: 47%	Permeate recovery 2nd pass: 90%
Total system recovery: 45.3%	Calculation for membrane age: 3 years (–20% nominal flow, +30% of nominal salt passage)
Feed flow 1st pass: 886.6 m^3/h (3904gpm)	Feed flow 2nd pass: 277.8 m^3/h (1223 gpm)
Feed pressure 1st pass: 65.6 bar (951 psi)	Feed pressure 2nd pass: 11.0 bar (159 psi)
Concentrate pressure 1st pass: 64.1 bar (929 psi)	Concentrate pressure 2nd pass: 6.4 bar (93 psi)
Elements type 1st pass: SWC5	Elements type 2nd pass: ESPA2+
Array 1st pass: 100 PV (8M)	Array 2nd pass: 18.9 PV (8M)
Average flux 1st pass: 14.0 l/m^2/h (8.2 gfd)	Average flux 2nd pass: 31.1/m^2/h (18.3 gfd)
Feed pH 1st pass: 8.2	Feed pH 2nd pass: 10

Constituent	Feed, ppm	Permeate, ppm	Concentrate, ppm
Temperature	26°C (79°F)		
pH	7.8	7.5	8.0
Ca	500	0.2	914
Mg	1540	0.5	2814
Na	13300	23	24294
K	490	1.1	894
CO_3	2.3	0.0	5.8
HCO_3	126.8	0.5	237
SO_4	3240	1.0	5928
Cl	23180	35	42341
F			
NO_3			
B	5.3	0.4	9.3
SiO_2			
TDS	42389	61	77440



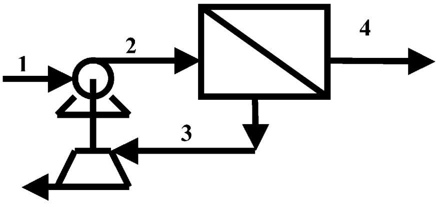
Location	1	2	3	4	5	6	7	8	9	10
Flow, m^3/h (gpm)	859.5 (3785)	886.6 (3904)	886.6 (3904)	469.9 (2069)	277.8 (1223)	277.8 (1223)	27.8 (122)	250 (1100)	138.9 (611)	388.9 (1711)
Pressure, bar (psi)			65.6 (951)	64.1 (929)		11.0 (159)	6.4 (93)			
TDS ppm	42389	41591	41591	78252	399	399	3941	13	128	80

	Pass 1 and total	Pass 2
Feed pressure, bar (psi)	65.6 (951)	11.0 (159)
Concentrate pressure, bar (psi)	64.1 (929)	6.4 (93)
Pump flow, m^3/h (gpm)	886.6 (3904)	277.8 (1223)
Turbine flow, m^3/h (gpm)	469.9 (2069)	
Pump efficiency, %	86	84
Motor efficiency, %	94	94
Energy recovery device efficiency, %	86	
Combined pumping power, kW (hp)	2117.7 (2840)	
Recovered power, kW (hp)	705.4 (946)	
Total power requirement, kW (hp)	1412.3 (1894)	
Pumping energy, kWh/m^3 (kWh/kgal)	3.63 (13.74)	

Caribbean feedwater temperature 26°C
Single pass SWRO

Feed water source: Caribbean seawater	Feed water type: surface intake
System configuration: single pass	Output capacity: 10000 m^3/d (2.64 mgd)
Permeate recovery: 50%	Feed flow 1st pass: 833.4 m^3/h (3667gpm)
Feed pressure pass: 60.5 bar (877 psi)	Concentrate pressure: 59.2 bar (858 psi)
Array: 100 PV (8M)	Elements type: SWC5
Average flux: 14.0 l/m^2/h (8.2 gfd)	Calculation for membrane age: 3 years (–20% nominal flow, +30% of nominal salt passage)
Raw water pH 1st pass: 8.2	Feed water temperature: 26°C (79°F)

Constituent	Feed, ppm	Permeate, ppm	Concentrate, ppm
pH	8.2	6.5	8.4
Ca	477	1.0	953
Mg	1160	2.5	2318
Na	11322	114.3	22530
K	386	4.9	767
CO_3	2.3	0.0	4.5
HCO_3	137	0.5	271
SO_4	2600	6.0	5194
Cl	20034	184	39883
F			
NO_3			
B	5.3	1.1	9.5
SiO_2			
TDS	36149	316	71983



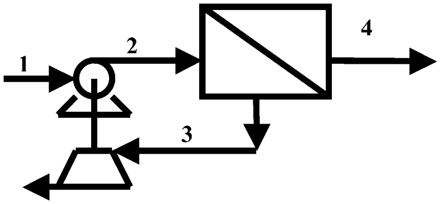
Location	1	2	3	4
Flow, m^3/h (gpm)	833.4 (3667)	833.4 (3667)	416.7 (1833)	416.7 (1833)
Pressure, bar (psi)		60.5 (877)	59.2 (858)	
TDS ppm	36149	36149	71983	316

	RO train
Feed pressure, bar (psi)	60.5 (877)
Concentrate pressure, bar (psi)	59.2 (858)
Pump flow, m^3/h (gpm)	833.4 (3667)
Turbine flow, m^3/h (gpm)	416.7 (1833)
Pump efficiency, %	86
Motor efficiency, %	94
Energy recovery device efficiency, %	86
Combined pumping power, kW (hp)	1738.6 (2331)
Recovered power, kW (hp)	577.7 (745)
Total power requirement, kW (hp)	1160.9 (1557)
Pumping energy, kWh/m^3 (kWh/kgal)	2.78 (10.52)

Pacific feedwater temperature 20°C
Single pass SWRO

Feed water source: Pacific seawater	Feed water type: surface intake
System configuration: single pass	Output capacity: 10000 m^3/d (2.64 mgd)
Permeate recovery: 50%	Feed flow 1st pass: 833.4 m^3/h (3667gpm)
Feed pressure pass: 57.6 bar (835 psi)	Concentrate pressure: 56.2 bar (815 psi)
Array: 100 PV (8M)	Elements type: SWC5
Average flux: 14.0 l/m^2/h (8.2 gfd)	Calculation for membrane age: 3 years (–20% nominal flow, +30% of nominal salt passage)
Raw water pH 1st pass: 8.0	Feed water temperature: 20°C (68°F)

Constituent	Feed, ppm	Permeate, ppm	Concentrate, ppm
pH	8.0	6.3	8.3
Ca	440	0.8	879
Mg	1300	2.3	2598
Na	10200	85	20315
K	380	4.0	756
CO_3	2.0	0.0	4.0
HCO_3	170	2.2	338
SO_4	3000	5.6	5994
Cl	18500	137	36863
F			
NO_3			
B	4.5	0.9	8.1
SiO_2			
TDS	34000	238	67763



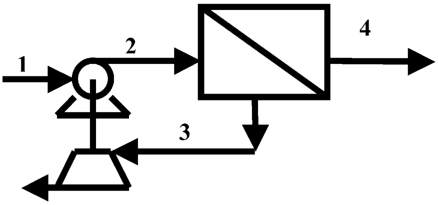
Location	1	2	3	4
Flow, m^3/h (gpm)	833.4 (3667)	833.4 (3667)	416.7 (1833)	416.7 (1833)
Pressure, bar (psi)		57.6 (835)	56.2 (815)	
TDS ppm	34000	34000	67763	237

	RO train
Feed pressure, bar (psi)	57.6 (835)
Concentrate pressure, bar (psi)	56.2 (815)
Pump flow, m^3/h (gpm)	833.4 (3667)
Turbine flow, m^3/h (gpm)	416.7 (1833)
Pump efficiency, %	86
Motor efficiency, %	94
Energy recovery device efficiency, %	86
Combined pumping power, kW (hp)	1652.9 (2217)
Recovered power, kW (hp)	548.5 (735)
Total power requirement, kW (hp)	1104.4 (1481)
Pumping energy, kWh/m^3 (kWh/kgal)	2.65 (10.03)

Atlantic feedwater temperature 20°C
Single pass SWRO

Feed water source: Atlantic seawater	Feed water type: surface intake
System configuration: single pass	Output capacity: 10000 m^3/d (2.64 mgd)
Permeate recovery: 50%	Feed flow 1st pass: 833.4 m^3/h (3667gpm)
Feed pressure pass: 60.0 bar (870 psi)	Concentrate pressure: 58.7 bar (851 psi)
Array: 100 PV (8M)	Elements type: SWC5
Average flux: 14.0 l/m^2/h (8.2 gfd)	Calculation for membrane age: 3 years (–20% nominal flow, +30% of nominal salt passage)
Raw water pH 1st pass: 8.0	Feed water temperature: 20°C (68°F)

Constituent	Feed, ppm	Permeate, ppm	Concentrate, ppm
pH	8.0	6.3	8.3
Ca	410	0.7	819
Mg	1302	2.3	2602
Na	10812	90	21534
K	389	4.0	774
CO_3	2.0	0.0	4.0
HCO_3	143	1.9	284
SO_4	2713	5.1	5421
Cl	19441	145	38737
F			
NO_3			
B	4.5	0.9	8.1
SiO_2			
TDS	35240	250	70230



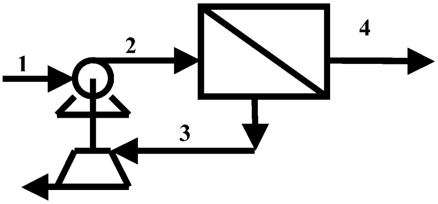
Location	1	2	3	4
Flow, m^3/h (gpm)	833.4 (3667)	833.4 (3667)	416.7 (1833)	416.7 (1833)
Pressure, bar (psi)		60.0 (870)	58.7 (851)	
TDS ppm	35240	35240	70230	250

RO train	
Feed pressure, bar (psi)	60.0 (870)
Concentrate pressure, bar (psi)	58.7 (851)
Pump flow, m^3/h (gpm)	833.4 (3667)
Turbine flow, m^3/h (gpm)	416.7 (1833)
Pump efficiency, %	86
Motor efficiency, %	94
Energy recovery device efficiency, %	86
Combined pumping power, kW (hp)	1723.2 (2311)
Recovered power, kW (hp)	572.9 (768)
Total power requirement, kW (hp)	1150.3 (1542)
Pumping energy, kWh/m^3 (kWh/kgal)	2.65 (10.03)

Canary Islands feedwater temperature 22°C
Single pass SWRO

Feed water source: Atlantic seawater	Feed water type: surface intake
System configuration: single pass	Output capacity: 10000 m^3/d (2.64 mgd)
Permeate recovery: 50%	Feed flow: 833.4 m^3/h (3667gpm)
Feed pressure pass: 64.8 bar (940 psi)	Concentrate pressure: 63.5 bar (921 psi)
Array: 100 PV (8M)	Elements type: SWC5
Average flux: 14.0 l/m^2/h (8.2 gfd)	Calculation for membrane age: 3 years (–20% nominal flow, +30% of nominal salt passage)
Raw water pH: 7.8	Feed water temperature: 22°C (72°F)

Constituent	Feed, ppm	Permeate, ppm	Concentrate, ppm
pH	7.8	6.1	8.1
Ca	464	0.9	927
Mg	1526	2.9	3050
Na	11700	105	23295
K	429	4.8	853
CO_3	3.2	0.0	6.4
HCO_3	204	2.9	405
SO_4	3059	6.1	6111
Cl	21344	170	42519
F			
NO_3			
B	4.5	1.0	7.96
SiO_2			
TDS	38739	294	77184



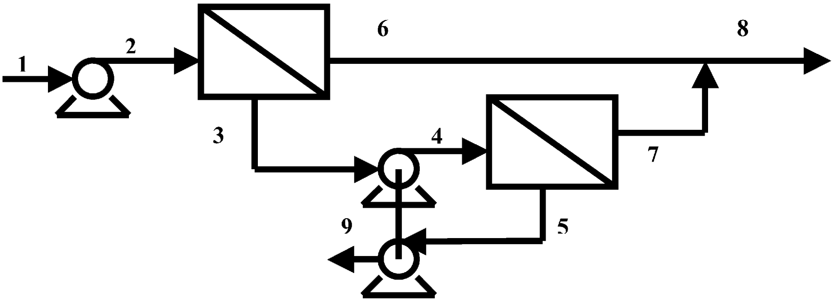
Location	1	2	3	4
Flow, m^3/h (gpm)	833.4 (3667)	833.4 (3667)	416.7 (1833)	416.7 (1833)
Pressure, bar (psi)		64.8 (940)	63.5 (921)	
TDS ppm	38739	38739	77184	294

	RO train
Feed pressure, bar (psi)	64.8 (940)
Concentrate pressure, bar (psi)	63.5 (921)
Pump flow, m^3/h (gpm)	833.4 (3667)
Turbine flow, m^3/h (gpm)	416.7 (1833)
Pump efficiency, %	86
Motor efficiency, %	94
Energy recovery device efficiency, %	86
Combined pumping power, kW (hp)	1865.5 (2502)
Recovered power, kW (hp)	619.7 (831)
Total power requirement, kW (hp)	1245.8 (15421671)
Pumping energy, kWh/m^3 (kWh/kgal)	2.99 (11.32)

High salinity brackish RO
Two stages with interstage booster pump at 28°C

Feed water source: brackish	Feed water type: well water
System configuration: two stages with interstage booster	Output capacity: 10000 m^3/d (2.64 mgd)
Permeate recovery: 80%	Feed flow: 520.9 m^3/h (2292gpm)
Feed pressure 1st stage: 17.1 bar (248 psi)	Concentrate pressure: 20.8 bar (280 psi)
Interstage booster pressure: 9 bar (131 psi)	Interstage flow m^3/h (gpm)
Array: 32: 16 PV (8M)	Elements type: ESPA4+
Average flux: 26.5 l/m^2/h (15.6 gfd)	Calculation for membrane age: 3 years (–20% nominal flow, +30% of nominal salt passage)
Raw water pH 1st pass: 7.0	Feed water temperature: 28°C (82°F)

Constituent	Feed, ppm	Permeate, ppm	Concentrate, ppm
pH	7.0	6.0	7.9
Ca	105	1.0	521
Mg	130	1.3	645
Na	1837	84	8854
K	85	4.8	406
CO_3	0.3	0.0	1.5
HCO_3	250	18.4	1176
SO_4	479	5.1	2374
Cl	2970	123	14356
F	1.4	0.1	6.5
NO_3	5.0	1.4	19.5
B			
SiO_2	17.0	0.5	83
TDS	5881	240	28444



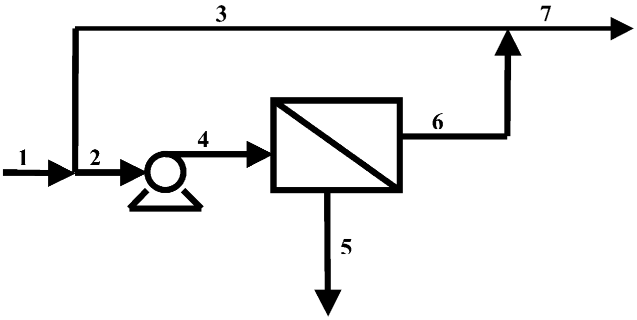
Location	1	2	3	4	5	6	7	8	9
Flow, m^3/h (gpm)	521.3 (2294)	521.3 (2294)	184.9 (814)	184.9 (814)	104.2 (458)	336.0 (1478)	80.7 (355)	416.7 (1833)	104.2 (458)
Pressure, bar (psi)		17.1 (248)	14.2 (206)	23.2 (336)	20.8 (302)				0.5 (7)
TDS ppm	5881	5881	16313	16313	28444	139	657	240	28444

	Main pump	Booster
Feed pressure, bar (psi)	17.1 (248)	9.0
Concentrate pressure, bar (psi)	14.2 (206)	20.8 (302)
Turbine discharge pressure		0.5 (7)
Pump flow, m^3/h (gpm)	521.3 (2294)	184.9 (814)
Turbine flow, m^3/h (gpm)		104.2 (458)
Pump efficiency, %	82	82
Motor efficiency, %	94	94
Energy recovery device efficiency, %		80
Combined pumping power, kW (hp)	374.4 (502)	
Recovered power, kW (hp)	46.1 (62)	
Total power requirement, kW (hp)	319.4 (428)	
Pumping energy, kWh/m^3 (kWh/kgal)	0.79 (3.0)	

Low salinity brackish RO
Two stages with permeate blending at 25°C

Feed water source: brackish	Feed water type: well water
System configuration: two stages with	RO permeate: 10000 m^3/d (2.64 mgd)
Blending ratio: 28.6%	Blended output flow: 14000 m^3/d (3.70 mgd)
Permeate recovery: 85%	Feed flow: 490.2 m^3/h (2157gpm)
Feed pressure pass: 8.0 bar (116 psi)	Concentrate pressure: 3.3 bar (48 psi)
Array: 32: 16 PV (8M)	Elements type: ESPA4+
Average flux: 29.2 l/m^2/h (17.2 gfd)	Calculation for membrane age: 3 years (–20% nominal flow, +30% of nominal salt passage)
Raw water pH 1st pass: 7.0	Feed water temperature: 25°C (77°F)

Constituent	Feed, ppm	Blended permeate, ppm	Concentrate, ppm
pH	7.0	6.6	7.7
Ca	96	29	628
Mg	11.7	3.5	77
Na	90	32.1	549
K	6.5	2.4	39
CO_3	0.0	0.0	0.0
HCO_3	72.6	30.4	408
SO_4	158.4	47.2	1041
Cl	190.7	61	1221
F	0.2	0.1	1.2
NO_3			
B			
SiO_2	24.3	9.3	143.3
TDS	647.3	215	4107



Location	1	2	3	4	5	6	7
Flow m^3/h (gpm)	657.4 (2893)	490.6 (2159)	166.8 (734)	490.6 (2159)	73.5 (323)	416.7 (1833)	583.5 (2567)
Pressure bar (psi)				8 (116)	3.3 (48)		
TDS, ppm	647	647	647	647	4107	41	215

	RO train
Feed pressure, bar (psi)	8 (116)
Concentrate pressure, bar (psi)	3.3 (48)
Pump flow, m^3/h (gpm)	490.6 (2159)
Pump efficiency, %	82
Motor efficiency, %	94
Pumping power, kW (hp)	138.4 (186)
Recovered power, kW (hp)	
Total power requirement, kW (hp)	138.4 (186)
Pumping energy, kWh/m^3 (kWh/kgal)	0.23 (0.87)

Appendix B

Example of desalination cost estimate

This appendix presents an example of a budgetary cost estimate for a 40,000 m^3/d (10.6 mgd) seawater desalination project. All costs included in this example are in year 2005 US$ and are based on actual data from similar size projects supplemented with cost information form budgetary vendor quotes and cost estimates for all key equipment, piping, materials and buildings.

Project description

Plant capacity and availability

The example project is a seawater desalination plant with an average annual plant production capacity of 40,000 m^3/d (10.6 mgd), and maximum installed production capacity of 48,000 m^3/d (13 mgd) when operated at 50% recovery. The plant is designed to have an availability factor of 100%, i.e., to produce 40,000 m^3/d (10.6 mgd) or more for 100% of the time (365 days per year) and operate in at recovery range of 45 to 50%. Plant's minimum daily production capacity is 32,000 m^3/d (8.5 mgd).

Plant location, intake and discharge

The plant is located on a 20,000 m^2 (4.9-acre) site in a commercially zoned area and is approximately 800 m (2,580 ft) from the shore. The plant site is an abandoned commercial property which has elevation of 10 m (33 ft) above the mean ocean tide level. The plant has open ocean intake that extends 200 m (660 ft) beyond the shore line. The plant discharge is a 950 m (3,120 ft) pipeline of which 150 meters (490 feet) extends into the ocean. The last 50 meters (160 feet) of the outfall are equipped with diffusers for concentrate dissipation.

Intake water quality

Key plant intake water quality parameters are summarized in Table B-1.

Product water quality specifications

The desalination plant will supply product water of water quality which is in compliance with the key parameters specified in Table B-2.

Key plant treatment facilities

The seawater desalination plant will have the following key treatment facilities:

- Intake pipeline—high density polyethylene pipe
- Bar racks—100 mm (4″) openings
- Intake screens—10 mm (0.4″) openings
- Intake pump station equipped with vertical turbine pumps
- Pretreatment facility combining coagulation and flocculation chambers dissolved air flotation system and dual-media (sand and anthracite) gravity filters
- Five duty and one standby RO trains of 8,000 m^3/d (2.1 MGD) production capacity each designed to operate at recovery range of 45 to 50%. Each RO train includes filter effluent transfer pump, cartridge filter, high-pressure pump coupled with Pelton wheel energy recovery turbine and an RO rack with membrane vessels and associated piping and equipment.
- Post-treatment system with limestone filters
- Chemical feed and storage systems
- Solids handling system, which consists of clarifiers for settling of the spent pretreatment filter backwash, and belt filter presses for dewatering of clarifier residuals
- Administration and RO system building
- Electrical substation
- Auxiliary facilities

Plant construction is planned to be completed in 22 months. The project will be implemented under a BOOT method of delivery. The debt financing for

TABLE B-1

Key intake seawater design characteristics

Parameter	Design minimum value	Design maximum value	Design average value
Intake flow, m^3/d (mgd)	68,000 (18.0)	114,000 (30.1)	84,000 (22.2)
Salinity (TDS), mg/L	32,500	34,500	33,500
Chloride, mg/L	16,900	20,800	18,000
Bromide, mg/L	52	79	73
Boron, mg/L	3.6	5.0	4.5
Temperature, °C	9	26	18
Turbidity, NTU	0.2	24	2
Total suspended solids, mg/L	0.5	30	4
pH	7.3	8.1	7.8

Note: All design characteristics are daily average values.

TABLE B-2

Key product water quality specifications

Quality parameter	Analytical method[1]	Sampling		Units	Concentration limits	
		Sample period[2]	Sample method		Central tendency[3]	Extreme[4]
Total dissolved solids	2540 °C	1 year	Weekly grab	mg/L	350	400
Chloride	4110 B	1 year	Weekly grab	mg/L	180	210
Bromide	4110 B	1 year	Weekly grab	mg/L	0.5	0.8
Boron	3120 B	1 year	Weekly grab	mg/L	No Limit	1[5]
Turbidity	2130 B	1 month	Continuous[6]	NTU	0.3	0.5

Notes:

1. All methods taken from Standard Methods On Line, published by APHA, AWWA, and WEF.
2. Sample period—concentration limits are calculated for this period.
3. Central tendency—can be exceeded in no more than 50% of samples taken.
4. Extreme—can be exceeded in no more than 10% of samples taken.
5. During first year boron may not exceed one (1) mg/L in more than ten (10) percent of the samples taken.
6. Continuous analysis, values at 15 minute intervals used in all calculations.

the project will be secured using commercial construction loan of 5% interest rate and 20-year term. The project will be financed with 10% equity and 90% debt. Project equity return on investment is 10%. The overall interest rate of the amortized investment is 5.5%.

This is a high-complexity project which will require two-year permitting process, a detailed hydrodynamic modeling of the plant discharge area and extensive source water sample collection and analysis. The project is likely to face legal challenges form local environmental groups.

Plant operations

The desalination plant will be highly automated and will be operated by staff comprising of 15 employees. The unit cost of power will be US$0.06/kWh. Dewatered residuals from the spent filter backwash will be disposed to a sanitary landfill in the vicinity of the plant. The cleaning solution from the reverse osmosis membrane cleanings will be discharged to the sanitary sewer.

Plant effluent discharge water quality will be measured at the point of exit from the desalination plant and the effect of this discharge on the marine environment will be monitored by collection and analysis of water quality samples at 10 monitoring stations located in the vicinity of the plant discharge.

Capital cost

The capital costs for construction, start-up and commissioning of the 40,000 m^3/d (10.6 mgd) seawater desalination plant are presented in Table B-3. The total capital costs are US$58 million (US$1,450/m^3/d, US$5.5/gpd).

Operation and maintenance cost

A breakdown of the annual O&M costs for the 40,000 m^3/d (10.6 mgd) project is presented in Table B-4. These costs total US$5.84 millions per year ($0.40/m^3, $1.51/kgal).

Cost of water

A summary of the fixed and variable components of the water cost for the 40,000 m^3/d (10.6 mgd) project is presented in Table B-5. The cost of water for this project is US$0.732/$m^3$ (US$2.771/kgal).

TABLE B-3
Project capital cost breakdown

Cost item	Capital cost US$	Capital cost % of Total
Direct capital (construction) costs		
1. Site preparation, roads and parking	580,000	1.0
2. Intake	2,760,000	4.8
3. Pretreatment	4,640,000	8.0
4. RO system equipment	18,560,000	32.0
5. Post-rreatment	1,160,000	2.0
6. Concentrate disposal	1,450,000	2.5
7. Waste and solids handling	870,000	1.5
8. Electrical and instrumentation systems	1,300,000	2.2
9. Auxiliary and service equipment and utilities	1,160,000	2.0
10. Buildings	1,740,000	3.0
11. Start up, commissioning and acceptance testing	1,160,000	2.0
Subtotal—direct (construction) costs (% of total capital costs)	US$35,380,000	61.0%
Project engineering services		
1. Preliminary engineering	580,000	1.0
2. Pilot testing	580,000	1.0
3. Detailed design	2,900,000	5.0
4. Construction management and oversight	1,740,000	3.0
Subtotal—engineering services	5,800,000	10.0
Project development		
1. Administration, contracting and management	1,160,000	2.0
2. Environmental permitting	2,610,000	4.5
3. Legal services	870,000	1.5
Subtotal—project development	4,640,000	8.0
Project financing costs		
1. Interest during construction	1,160,000	2.0
2. Debt service reserve fund	2,900,000	5.0
3. Other financing costs	2,320,000	4.0
Subtotal—project financing	6,380,000	11.0
Contingency	5,800,000	10.0
Subtotal indirect capital costs (% of total capital costs)	US$22,620,000	39.0%
Total capital costs	US$58,000,000	100%

TABLE B-4
Project annual operation and maintenance cost breakdown

	Annual O&M costs		
Cost item	Million US$	US$/m^3 ($/kgal)	% of Total
Variable O&M costs			
1. Energy	3.24	0.222 (0.840)	55.5
2. Chemicals	0.35	0.024 (0.091)	6.0
3. Replacement of membranes and cartridge filters	0.62	0.042 (0.159)	10.6
4. Waste stream disposal	0.26	0.018 (0.068)	4.4
Subtotal—variable O&M costs	4.47	0.306 (1.158)	76.5
Fixed O&M costs			
1. Labor cost	0.33 MM	0.023 (0.087)	5.7
2. Maintenance	0.38 MM	0.026 (0.098)	6.5
3. Environmental and performance monitoring	0.09 MM	0.006 (0.023)	1.5
4. Indirect O&M costs	0.57 MM	0.039 (0.148)	9.8
Subtotal—fixed O&M costs	1.37 MM	0.094 (0.356)	23.5
Total O&M costs	US$ 5.84 MM	0.400 (1.514)	100%

TABLE B-5
Project water cost breakdown

	Cost of water	
Cost of water item	US$/m^3 ($/kgal)	% of Total
Fixed cost component		
1. Capital cost recovery	0.332 (1.257)	45.4
2. Labor costs	0.023 (0.087)	3.1
3. Maintenance	0.026 (0.098)	3.6
4. Environmental and performance monitoring	0.006 (0.023)	0.8
5. Indirect O&M costs	0.039 (0.148)	5.3
Subtotal—fixed water cost component	0.426 (1.613)	58.2
Variable cost component		
1. Energy	0.222 (0.840)	30.3
2. Chemicals	0.024 (0.091)	3.3
3. Replacement of membranes and cartridge filters	0.042 (0.159)	5.7
4. Waste stream disposal	0.018 (0.068)	2.5
Subtotal—variable water cost component	0.306 (1.158)	41.8
Total cost of water	0.732 (2.771)	100%

Cost of water estimate for availability factor lower than 100%

The total cost of water in the example above is estimated assuming that the desalination plant will produce water 100% of the time. The actual desalination plants however, usually have less than 100% availability. Therefore, the cost of water has to be adjusted to account for the fact that for a portion of the time the plant will not deliver desalinated water to the final users (i.e., will not generate revenue from water sales) while incurring expenses associated with fixed plant costs. Because the variable cost component is proportional to flow, the plant availability factor would not have an effect on the variable water cost component expressed as unit cost (i.e., \$/m^3 or \$/kgal). However, the plant fixed unit cost will increase, since the same amount of fixed expenses would need to be recovered at reduced water sales.

For the example above, let's assume that the plant availability factor is 95% rather than 100%. In this case the actual annual average volume of desalinated water produced by the 40,000 m^3/d (10.6 mgd) desalination plant would be: 40,000 m^3/d × 95% = 38,000 m^3/d (10.0 mgd). As a result, the fixed component of the desalination cost would increase from \$0.426/m^3 (1.613/kgal) to \$0.426/m^3 / (95%) = \$0.448/m^3 (\$1.696/kgal). Therefore, the total cost of water will increase from \$0.732/m^3 (\$2.771/kgal) to: \$0.448/m^3 + \$0.306/m^3 = \$0.754/m^3 (\$2.854/kgal). This 3% increase in the desalination plant's total cost of water would account for the design plant availability factor and would allow to recover plant fixed expenses during times the plant is not delivering desalinated water to the final user.

Appendix C

Example of feasibility evaluation of RO concentrate disposal alternatives

The purpose of this appendix is to provide insight into how the feasibility of traditional disposal options is determined. The general nature of the regulatory process for the traditional disposal options is discussed followed by recommendations for conducting an initial feasibility screening of these options. As in Chapter 19, the comments and examples provided are based on regulations and experiences in the United States. The general principals and approaches taken, however, have broader applicability and are instructional. The example used is from consideration of concentrate disposal in the Denver, Colorado area. Feasibility issues are discussed. Detailed costs are not provided due to the extremely site-specific nature of costs.

A particular concentrate is used to illustrate the feasibility considerations. This concentrate has a flow of 5 MGD and a TDS of 4,000 mg/L. Table C-1 shows the water quality analysis for the concentrate and the levels of various receiving water standards that are discussed in the text.

1. Surface water disposal (inland)

1.1. General permitting process approach

From a regulatory standpoint, the feasibility of surface discharge depends on the impact the discharge might have on the receiving water. The degree of impact is regulated by water quality standards and toxicity test standards dependent on the specific receiving water.

Receiving waters are classified according to their beneficial uses. While the classifications may vary from state to state, they typically include such

TABLE C-1
Concentrate water quality and selected receiving water standards

	Concentrate level (mg/L)	South Plant River water quality standards	Irrigation ditch water quality standards	Groundwater standards for direct irrigation
K	37			
Na	523			
Mg	103			
Ca	471			
Sr	11.62			
Ba	0.21			2.0
CO_3	8			
HCO_3	1059			
NO_3*	206	50	500	
Cl	436	250		250
F	6.25			
SO_4	1074	250		250
SiO_2	65			
CO_2	26			
TDS	4000			
pH	7.6	6.5–9.0		

*The level of nitrate in units of nitrate nitrogen is 41 mg/L; the standard in these units is 10 mg/L.

categories as water supply, recreation, agriculture, aquatic life cold, aquatic life warm, etc. and these may have sub-classifications. Water quality standards (receiving water standards) are assigned to the different classifications representing maximum allowable levels of various parameters. The standards may be numeric (exact values) or formulas where actual values are dependent on pH, temperature, hardness, etc. When multiple uses (classifications) apply to a water, the most stringent standard for a given constituent applies. A parameter may have both acute and chronic standards. Acute standards are tied to lethal effects, and chronic standards to chronic toxic effects such as growth and reproduction.

Comparison is made between the water quality standards and the concentrate levels to gauge the potential impact of the concentrate on the receiving water. The comparison can be made in a variety of ways. Two extremes are discussed below.

Some states directly compare the parameters, commonly referred to as end-of-pipe regulation. If the concentrate parameter is less than the standard, then that parameter will not limit discharge. If the parameter is greater than the standard, a mixing zone may be granted. Thus, if C is the concentrate parameter:

• $C <$ water quality standard	discharge is generally allowed
• $C >$ water quality standard	discharge may or may not be allowed depending on mixing zone considerations

States are required to have mixing zone policies, with the mixing zone representing a form of regulatory relief where conditions can be met at the edge of a mixing zone as opposed to end-of-pipe. Some states, such as Colorado, do not use end-of-pipe regulation but include the mixing zone considerations in the comparison of parameters.

In this more complex and complicated regulatory approach, a mass load calculation is made which takes into account concentration and flow of discharge, and the flow, concentration and standard of the receiving body. With concentrations in mg/L and flows in m^3/d (MGD), in Colorado, quantities used in the calculations for a given constituent are listed in Table C-2.

TABLE C-2
Quantity and value parameters used in mass load calculations.

Quantity	Value
Discharge flow, Q_1	Design flow
Discharge concentration, Y	Design concentration
Upstream concentration, X	85th percentile for most constituents; 50th percentile for others
Upstream flow, Q_0	Statistically derived low flow that occurs once in three years; for acute calculations the low flow value is determined from analysis of daily values from the most recent 10 year period; for chronic calculations the low flow value is determined from analysis of all 30 consecutive day average values from the most recent 10 year period
Downstream flow, $Q_1 + Q_0$	Net flow from upstream and discharge flows
Downstream concentration, Z	This is the calculated quantity to be compared with the water quality standards (acute and chronic).

The calculation is: $Z = [Y \cdot Q_1 + X \cdot Q_0] / (Q_0 + Q_1)$

The possible outcomes and their implications are listed in Table C-3.

TABLE C-3

Possible outcome of mass load calculations

Result of calculations	Possible regulatory outcome
$Z <$ water quality standard	Discharge is generally allowed
$Z >$ water quality standard	Discharge may or may not be allowed depending on mixing zone considerations

In Colorado a physical mixing zone must be defined based on in-stream measurement (such as with dyes) or on modeling. The physical mixing zone is defined as the area between the discharge location and the downstream point at which the conductivity variation across the width of the stream is reduced to 10%. Separate from this, a regulatory chronic mixing zone (area) is defined as six times the full flow width squared. The acute mixing zone is defined as 10% of this area. If the physical mixing zone is less than the regulatory mixing zones, the full upstream dilution may be allowed. If the physical mixing zone is greater than the regulatory mixing zones, the discharge flow and/or concentration must be reduced.

While this approach is more complex than the direct end-of-pipe approach, it is not necessarily less forgiving. States using the end-of-pipe approach may have more restrictive mixing zone policies.

In addition to standards for specific parameters, permit conditions may include the requirement for whole effluent toxicity tests (WET tests) which determine the impact of the concentrate on organisms representative of the receiving water (136).

1.2. Special permitting process used in the Denver area for the South Platte River

In this example the proposed membrane plant was within 2 miles of the South Platte River and conveyance costs were not a major concern. The feasibility question rested on whether or not the discharge could be permitted. The river is unusual in that it is effluent dominated. That is, its flow is almost entirely made up from various discharges—the largest by far being that from the

Denver Metro wastewater treatment plant with a capacity of 757,000 m^3/d (200 MGD). Each new discharge may affect permit conditions of other dischargers, and river conditions can change over short periods of time. As a result, a more complex approach was taken to determine allowable discharges.

The approach used was to extensively model the entire river segment (approximately 32 km, 20 miles in length) to account for all input streams (tributary flows into the river, all dischargers) and all removal streams (irrigation drainage canals).

Historical flow and concentration data were updated with values used for the low flow analysis derived by statistical analysis of each stream's flow data. A series of models was developed to evaluate both conservative constituents (species that travel unchanged downstream) and non-conservative constituents (species such as ammonia, nitrate and dissolved oxygen).

The extensive modeling identified nitrate as the only constituent likely of concern. Because of this, discharge at the design flow conditions was determined not to be possible year round.

The state offered the municipality a permit for reduced flow for certain months. The municipality put construction of the plant on hold, where the situation still stands.

1.3. Discharge to irrigation canal

Another possibility of surface discharge of the concentrate in the Denver region example was to an irrigation ditch. Irrigation ditches are streams drawing water off of the South Platte River for agricultural purposes. The ditches are thus of lower flow volume than the South Platte River. They can be dry during drought conditions and are dry during 5 months of the non-growing season. The potential advantage of discharge to an irrigation ditch over discharge to the South Platte River is that the irrigation ditches have an agricultural classification and with that a nitrate standard 10 times higher than that for the South Platte. Thus nitrate levels would not be an issue. Water quality standards for other constituents are the same as or higher than for the South Platte River—so they were not a problem. Consequently, permitting of this option was evaluated to be feasible.

If membrane plant operation were year round, the concentrate produced for 5 months would have to be stored. At 19,000 m^3/d (5 MGD) this amounts to 2,877,000 m^3 (760 MG) or 2345 AF. This is a sizeable storage pond/area, depending on pond depth as shown below.

Depth, m (ft)	Area, m^2 (acres)
3 (10)	951,000 (235)
6 (20)	478,000 (118)

When the stored concentrate is also sent to the irrigation ditch during the irrigation season, this represents an additional TDS load of the ditch. Assuming 7 months irrigation season and uniform use of the stored concentrate, this adds13,500 m^3/d, 3.57 MGD (concentrate volume divided by the 213 days of the irrigation season) to the 19,000 m^3/d (5 MGD) being generated during the irrigation months, for a total of 3,244,000 m^3 (857 MG). The irrigation water TDS in this case is projected to be from 939 to 1145, or an increase from 239 to 445 ppm from the base level of 700. While this salinity level does not violate any irrigation ditch salinity standard [there are no salinity surface water standards in Colorado other than for the Colorado River Basin], it might be flagged by the regulators issuing a permit to discharge to the irrigation ditch. The high salinity raises questions from the perspectives of groundwater protection, cattle drinking, SAR ratios for crops, or other concerns.

Such a concentrate management scenario may be possible during the irrigation season. A different municipality in the Denver region is considering using this scheme for concentrate management. The plan is not to run the plant except to meet peaking needs; thus the membrane plant would be mothballed in the winter or run at a low rate. In this situation, concentrate storage during the non-irrigation season would be minimal.

A storage issue that would need to be addressed if this option were to be considered further is the accumulation of sparingly soluble solids that would be precipitated from the concentrate due to the limited time effectiveness of the antiscalant (from hours to days).

The distance from the proposed plant site to the irrigation ditch is a minimum of about 8 km (5 mi)—as a straight line. Issues of pipeline, pumping, right of way, etc. would need to be evaluated. Thus the benefit of not having to worry about nitrate levels is compensated for by the complexity and unknowns of the scenario. In summary these factors include:

- Conveyance of the concentrate to the ditch site
- Need for very large cold season storage areas for concentrate
- (Likely concern) TDS increase of irrigation ditch water due to generated and stored concentrate

This disposal option was not considered feasible.

1.4. Suggested approach to initial evaluation of surface disposal options

- Determine if there is a candidate receiving water within 32 km (20 mi) of the membrane plant
  - ○ if so, discharge may be feasible
  - ○ if not, conveyance costs may be prohibitively expensive
- Estimate concentrate parameters such as volume, salinity, and general water quality
- Interact with the local regulatory agency to:
  - ○ Get an initial indication of whether disposal to the receiving water is possible and worth pursuing further–based only on general information
  - ○ If so, document the specific regulations that would apply and how to obtain them
    - – Include classification system of waters
    - – Include corresponding water quality standards that apply
    - – Include mixing zone policy
- Determine which water quality standards apply; this may involve a few iterations with the regulatory agency
- Make a direct comparison of the concentrate water quality parameters with the water quality standards; highlight those parameters that are likely of concern.
- This then becomes the basis for further interaction with the regulatory agency
- Develop preliminary level estimates for capital and operating costs. Discussions of design and cost factors in reference (139) should be helpful in this regard

1.5. Comments

As the above examples illustrate, regulations can be complex and challenging to understand. It is important to interact with the regulatory agency to take advantage of their knowledge and understanding, and it is also important to develop an understanding of the regulations such that interactions with the regulatory groups can be productive.

Feasibility of general surface discharge is largely dependent on whether a discharge permit can be obtained. Other issues such as conveyance cost, outfall cost, monitoring costs, etc., are typically reasonable.

2. Disposal by land applications

The concentrate management options for land application included:

- Discharge to an irrigation ditch / canal for subsequent spray irrigation (previously discussed under surface water discharge)
- Direct discharge (after dilution) for landscape, golf course, etc. irrigation
- Direct discharge (after dilution) for agriculture spray irrigation

The second of these options is discussed below.

2.1. Direct discharge (after dilution) for landscape, golf course, etc. irrigation

Some of the same technical issues and factors just discussed apply–possible need for cold season storage and need for dilution of concentrate prior to use for irrigation.

The concentrate will need to be diluted to an extent determined by the discharge standards that apply. Comparison of the standards with the levels in the concentrate provide an estimate of the level of dilution required for each constituent, and from this, determination of the limiting constituent and required dilution.

Discharge limitations (groundwater standards) are dependent on classification of the groundwater. In Colorado, such classifications include: human health, secondary drinking water, and agricultural. Regulations specify classification and standards corresponding to these classifications. In the example case, for the purpose of establishing a basis for evaluating land application of concentrate, the following assumptions were made:

- The groundwater area is not classified, so other standards (interim narrative standards) apply
- The ambient quality of the ground water is not known.

For this situation the most stringent standards apply from among the human health, drinking water, and agricultural standards. A selected few are shown in Table C-4:

TABLE C-4
Colorado groundwater standards for some species where interim narrative standards apply

Barium	2.0 mg/L
Chloride	250 mg/L
Iron	0.3 mg/L
Nitrate (as N)	10 mg/L
Sulfate	250 mg/L

From groundwater maps of the area it appears that the TDS of the shallowest groundwater is less than 500 mg/L. Assuming this to be the case, the maximum allowable TDS concentration would be 400 mg/L or 1.25 times the background level—whichever is less restrictive. If the TDS is greater than 500 (but less than 10,000) the maximum allowable TDS concentration is 1.25 times the background level.

The minimum possible dilution level of concentrate required to meet the groundwater standards for each of the Table C-5 constituents may be calculated based on the concentrate levels, the standard levels, and the assumption of dilution water level of the species in question. Table C-5 shows the dilution ratios.

TABLE C-5
Dilution ratios required for dilution waters of different species levels to meet standard levels

Species in the concentrate	Concentration level in the concentrate (mg/L)	Standard concentration level (mg/L)	Volume ratio of dilution water to concentrate for different dilution waters		
			Level of species in dilution water as percent of standard		
	Concentrate level (mg/L)	Standard level (mg/L)	0% of standard	50% of standard	80% of standard
Chloride	436	250	0.74 : 1	1.49 : 1	3.72 : 1
Nitrate	206	50	3.1 : 1	6.2 : 1	16 : 1
Sulfate	1074	250	3.3 : 1	6.6 : 1	16.5 : 1
TDS if = 500	4000	500	8 : 1	8 : 1	8 : 1

Table C-5 may be interpreted as follows: for a concentrate of nitrate level 206 mg/L, if the dilution water has a nitrate level of 80% of the standard level

(80% of 50 or 40 mg/L), then for the mixture to be diluted to the standard level of 50 mg/L, 16 parts of the dilution water would need to be added to 1 part of the concentrate. From this it is obvious that the required level of dilution is strongly dependent on the water quality of the dilution water AND the makeup of the underlying groundwater. For the purpose of allowing a calculation, it is assumed that the dilution water is of TDS 500 mg/L and anion levels of chloride, nitrate, and sulfate are at 50% of the standard levels, and thus 125, 25, and 125 respectively. Levels of the other constituents are likely even less than the 50% level giving rise to TDS being the limiting parameter–as reflected in the next to last column of Table C-5. This would dictate a required dilution of 8: 1. It is unlikely that this volume of dilution water is available.

Another issue has to do with the amount of land required for irrigation.

Examination of local irrigation schedules and practices suggested average irrigation rates (over the entire irrigation season) to be from 4 to 12 gpm per acre. Based on these numbers, the acreage that can be serviced by the concentrate volume plus the dilution water volume (a total of 9 times the concentrate volume) is from 11 km^2 to 32 km^2 (2,610–7,830 acres)

In summary the problematic issues with direct irrigation include:

- Need for large amounts of dilution water
- Large amount of land required
- Need for distribution system to provide the water over the large area
- Need for cold season storage for concentrate

There may be other limiting concerns such as the ability of specific vegetation to uptake nitrate. As we have seen, however, it appears that nitrate is not the limiting parameter—it is salinity.

This option was evaluated to not be feasible.

2.2. Suggested approach to initial evaluation of land application disposal options

- Determine if land application possible year round
    - If possible, then option may be feasible
    - If not, option is likely not feasible due to need for storage or another disposal option
- Determine if the volume of concentrate less than 1 MGD
    - If so, then land application may be feasible
    - If not, the costs of land application may be prohibitive

- Estimate concentrate parameters such as volume, salinity, and general water quality
- Interact with the local regulatory agency to:
  - ○ Get an initial indication of whether disposal to the land is possible and worth pursuing further–based only on general information
  - ○ If so, document the specific regulations that would apply and how to obtain them
    - – Include classification system for groundwaters
    - – Include corresponding water quality standards that apply
- Compare concentrate water quality to groundwater standards
- Calculated dilution levels required and limiting constituent (maximum dilution required)
- Evaluated whether dilution water is available
- Estimate the land area required for using the diluted concentrate
- Develop preliminary level estimates for capital and operating costs. Discussions of design and cost factors in reference (139) should be helpful in this regard.
- If the option still looks promising, interact with regulatory agencies for further definition of requirements and more detailed evaluation

2.3. Comments

Land applications in the U.S. are not, in general, viable except where concentrate volumes are small, climate will support irrigation year round, and concentrate is of low salinity.

3. Discharge to sewer

In the United States a surface water discharge permit is not required for discharge to the sewer. Permission of the WWTP is required and the WWTP can dictate the conditions and costs of disposal to the sewer.

In response to a letter from the municipality considering the proposed membrane plant, the Denver Metro WWTP said that disposal to the front end of the POTW would require a one-time “buy-in” fee of over $30MM. This was more of a message that they were not encouraging further consideration of this possibility than a detailed cost estimate. In subsequent discussions with WWTP officials, various technical questions were raised including that of selenium levels.

They also suggested that legal issues would have to be addressed as their attorneys tell them that they would have to allow for the possibility of other WTPs wanting to do the same thing in the future.

Most of the South Platte River volume downstream of the Metro WWTP is the Metro effluent. Thus, in terms of dilution, disposal of the concentrate to the Denver Metro WWTP effluent is similar to discharging the concentrate to the South Platte River that was discussed under the surface water disposal sections.

This option was not considered feasible for reasons of the considerable one-time fee and the reasons discussed above in regards to discharge to the South Platte river.

3.1. Suggested approach to initial evaluation of disposal to sewer:

- Identify local WWTPs
- Obtain information about their capacities, effluent characteristics (including volume, water quality, and TDS)
- Estimate increase in WWTP effluent conditions due to concentrate
- Discuss the possibility of discharge to the sewer with WWTP

4. Deep well injection

In the United States, membrane concentrate is classified as an industrial waste and as such requires a Class I well for disposal. Class I wells utilize aquifers that are structurally isolated from overlying drinking water aquifers (defined as any aquifer of TDS less than 10,000). Monitoring requirements stipulate a tubing and packer arrangement that prevents direct leakage of concentrate from the well. The injection tubing is surrounded by an annular space filled with a monitoring fluid that is tested for changes in salinity to monitor leaks. Total cement casing is also required. These three requirements (wells below the drinking water aquifers, tubing/packer arrangement, total cement casing) add considerable cost to the injection wells. Class II wells for disposal of produced waters from oil and gas drilling do not have these restrictions. In Colorado Class I wells are overseen by the USEPA and Class II wells by the Colorado Oil and Gas Conservation Commission (COGCC).

The general needs for a disposal well include:

- Depth—for Class I the injection zone needs to be below the drinking water limit (known as the USDW depth)

- Layering—for Class I wells, the receiving aquifer needs to be structurally confined and isolated from overlying drinking water aquifers
- Salinity—the receiving water aquifer must be of salinity greater than 10,000 mg/L
- Porosity and permeability of the receiving formation—porosity is the void aspect and permeability is the communication or movement aspect; these two characteristics need to result in sufficient aquifer capacity (long-term aspect) and uptake rate (instantaneous aspect)

In theory, the injection well possibilities for disposing/reusing the concentrate include:

- Use of existing class I wells
- Drilling of class I wells
- Reworking of abandoned class II wells
- Using the concentrate for pressure maintenance (water flooding) of oil wells

4.1 Use of existing Class I wells

There is only one class I well in eastern Colorado. This well was drilled to a depth of over 2700 m (9000′). The well has 1400 mm (5.5″) casing and 70 mm (2.875″) tubing. It currently takes about 160 m^3/day (42,000 gpd). It would take 119 of these same sized wells to dispose the 19,000 m^3/d (5 MGD) concentrate. This option was determined as not feasible.

4.2. Drilling of Class I wells

Examination of well descriptions in a national deep well data base (151) shows the largest capacity Class I wells in several states to be:

Florida	83,000 m^3/d (22 MGD)	Wyoming	1,890 m^3/d (0.5 MGD)
Texas	11,300 m^3/d (3 MGD)	Louisiana	1,630 m^3/day (0.43 MGD)
N. Dakota	3,400 m^3/d (0.9 MGD)	Michigan	1,360 m^3/day (0.36 MGD)
Oklahoma	2,650 m^3/d (0.7 MGD)	Kansas	1,130 m^3/day (0.30 MGD)
Illinois	2,270 m^3/d (0.6 MGD)		

These data do not support the probability of finding an aquifer with high capacity in Colorado (and for that matter in many other locations).

The Denver area, in hydro-geologic terms, is referred to as the Denver Basin. It is roughly an oval shaped area extending north and south about 130 miles east and west about 80 miles. The geological formations increase in depth from the surface as one moves from the edge to the center of the basin–regardless of compass direction. The planned membrane plant is in the central area of the basin. Consequently, the depth to an adequate confining layer is likely represented by the single Class I well in Colorado. Depth to a confining layer might be substantially less 40 miles to the East, towards the edge of the basin.

In general, the depth to USDS and the capacity and uptake rate of a potential receiving aquifer are unknown. Discussions with local hydrologists suggest that it would take many wells to inject 19,000 (5.0 MGD) given the available, but limited, information about local aquifers. Further, due to the nature of the receiving formation, individual wells would likely be thousands of feet apart rather than hundreds as in Florida, where deep injection well conditions are much more ideal.

A more in-depth analysis of available data would be needed to determine if a candidate location/receiving formation could be identified with enough promise to warrant an exploratory bore hole. The depth to USDS and the receiving formation characteristics would likely still be unknowns before the drilling. Such a bore hole might cost $1MM or more. There is likely a tradeoff between the distance of the well from the plant location with depth to the confining aquifer. The closer the location is to Denver, the more difficult it will be to sell the idea to the public.

The unknowns, lack of encouraging information, and substantial cost of drilling a bore hole to obtain information suggest this disposal option is not feasible.

4.3 Reworking of an abandoned Class II well

Maps from on-line databases of oil and gas wells show many oil and gas wells in the general Denver area. Information about these class II wells is available from the COGCC online database. A major limitation for existing class II wells is the small casing diameter which is typically 100–140 mm (4–5.5″). If class II wells were reworked to make a class I, well the well depth would need to be increased along with the well diameter (the latter to allow for the injection tubing and surrounding annular space–both inside the casing).

Hydro-geological studies similar to that needed for a new well would be necessary to establish the confined nature of the aquifer and if suitable the capacity of the aquifer for uptake rate of concentrate.

As with the drilling of a Class I well, reworking of Class II wells still has the same unknowns of:

- Depth to get below the USDW level
- Existence of suitable underlying aquifer
- Depth to that aquifer
- Permeability/capacity of that aquifer—possible injection rate per well
- Well spacing for multiple wells

This option is judged not to be feasible.

4.4 Use of concentrate for pressure maintenance in existing Class II wells

The concentrate TDS of 5,000 mg/L would be less than that of some produced waters in Colorado. From this perspective, the use of concentrate for water flooding/pressure maintenance of Class II producing wells might be a possibility. Given the size of the Class II wells, however, it would take many wells to utilize the 19,000 m^3/d (5.0 MGD) of concentrate. This in turn would require a distribution system. Water compatibility is an issue that would need to be addressed, as well as chloride levels. Perhaps most limiting is the requirement that the application be available for the life of the desalination plant—an unlikely situation. This option is assumed to be not feasible.

4.5 Suggested approach to initial evaluation of deep well injection

- Estimate concentrate parameters such as volume, salinity, and general water quality
- Interact with the local regulatory agencies to:
    - Determine what groups oversee the municipal, industrial, and gas and oil wells.
    - Determine whether deep well injection has been done in the general area
    - Determine where and how to access more information about individual wells
- Characterize existing deep wells in terms of location, capacity, and depth.
- Discuss local aquifer characterization, general feasibility with hydrogeologists

- Develop preliminary level estimates for capital and operating costs. Discussions of design and cost factors in reference (139) should be helpful in this regard.

4.6 Comments

Deep well disposal is not feasible in many locations. This may be due to any of several reasons that include lack of suitable aquifer conditions, excessive distance from the membrane plant, regulatory policy not permitting deep wells, and cost.

Even in areas where deep well disposal may be possible based on an initial evaluation, the costs involved with a more extensive evaluation, with drilling a test hole, and with drilling and testing of the disposal wells can be significant.

While there are historical incidents of earthquakes associated with deep well disposal they are generally correlated with high injection pressures used to obtain higher injection rates—pressures typically greater than fracture pressure. Injection pressures for Class I wells are generally much less than fracture pressures and unless wells are located in a fault zone, such earthquake possibilities should not be of significance.

Deep disposal wells are required to undergo periodic mechanical integrity testing. For these times an alternative permitted disposal option is required.

If the volume of concentrate were dramatically reduced by a factor of 10, such as through volume reduction technologies, then disposal of 1,900 m^3/d (0.5 MGD) might be done in 2 to 3 wells—significantly reducing the overall costs. The unknowns (and associated risks) associated with deep well injection, however, would still be present.

5. Evaporation

Nearly all municipal membrane facilities in the United States using evaporation ponds for concentrate disposal are for plants of size less than 3,780 m^3/d, 1 MGD (product water) as shown below. This is because of the small economy of scale for evaporation ponds and the need for large, level, and relatively inexpensive areas of land.

The proposed example plant size is 56,780 m^3/d (15 MGD) with 19,000 m^3/d (5.0 MGD) of concentrate–10 times the size of the largest municipal desalting plant evaporation pond. The location is less favorable in terms of net evaporation rate than the locations of these other plants.

TABLE C-5

Listing of US membrane plants using evaporation ponds for concentrate disposal.

Municipal membrane plants using evaporation ponds (as of 2002)		
Location	Type	Size m^3/day (MGD)
Buckeye, AZ	EDR	5680 (1.5)
Terlingua, TX	BRO	190 (0.05)
Experanza, TX	BRO	220 (0.058)
El Paso, TX	BRO	300 (0.08)
Lucien, OK	MF	450 (0.12)
Sarasota, FL	BRO	760 (0.2)
Los Ybanex, TX	BRO	830 (0.022)
Austin, TX	BRO	550 (0.144)

At an estimated 1015 mm/y (40 in./y) net evaporation for the Denver Basin area and 19,000 m^3/d (5.0 MGD) of concentrate, the area required would be over 6.9 km^2, 1700 acres (evaporative surface plus land areas).

Use of evaporation ponds was determined not to be feasible based on the large land requirement and excessive cost.

5.1. Suggested approach to initial evaluation of evaporation ponds

- Determine if use of evaporation ponds is possible year round
  - If possible, then option may be feasible
  - If not, option is likely not feasible due to need for storage or another disposal option
- Estimate concentrate parameters such as volume, salinity, and general water quality
- Determine if the volume of concentrate is less than 4000 m^3/d (~1 MGD)
  - If so, then land application may be feasible
  - If not, the costs of evaporation ponds may be prohibitive
- Estimate the land required for evaporation ponds
- Interact with the local regulatory agencies to:
  - Determine what kind of permitting is required and what pond construction requirements are

  - Determine if there have been previous permitted evaporation ponds in the general area
  - Get an initial indication of whether disposal to the land is possible and worth pursuing further–based only on general information
- Develop preliminary level estimates for capital and operating costs. Discussions of design and cost factors in reference (139) should be helpful in this regard.

5.2. Comments

A maximum net evaporation rate in the United States–suitable for Las Vegas or Phoenix is 2000 mm/year (80 inches/year). This corresponds roughly to about 4l/1000 m^2 (4 gpm/acre). Generally, due to other factors including additional land area requirements beyond evaporative surface area, and factors that reduce net evaporation rates—including salinity—a rate of 2 gpm/acre is more appropriate for the maximum rate.

2005 costs range from \$7500/1000m^2, \$30,000/acre (low capital cost end and no disposal of solids) to \$50,000/1000m^2, \$200,000/acre (higher capital costs and disposal of solids). Thus, the high cost of large evaporation pond areas is evident as is general feasibility being limited to smaller areas.

Rapid buildup of solids from high salinity concentrate can result in significant additional cost. As a pond fills up with solids it may be covered over and retired from useful life and another pond constructed to handle additional solids. Another option is to remove solids from a pond and dispose of them in a landfill. Both options can be costly. In all cases, it needs to be determined if contaminants in the concentrate, when concentrated in the solids, will lead to the solids being hazardous. If so, mitigation steps, representing additional cost, need be considered.

6. Enhanced evaporation

There are several different enhanced evaporation systems. They take advantage of the increased evaporation possible through increasing the exposure of the water to the air. Some approaches used include spraying or misting the water into the air, letting the water fall through the air, or saturating a cloth material and exposing it to the air flow. These methods are relatively unproven but have the potential to reduce required land area and capital costs. Operational concerns include drift onto adjacent property, wind sensitivity of the enhanced evaporation effect, and worker concern with frozen mist in winter. In the best of conditions, the various enhanced evaporation systems increase the net evaporation rate (reduce the evaporative surface area required) by a factor of about 5–7. In the example case, this would reduce the area required from 6.9 km^2 to about 1.2 km^2 (1700 acres to about 300 acres)—still a sizable land area.

In a recent Ft. Bliss, Texas evaluation conducted on 12,000 m^3/d (3.2 MGD) of concentrate the use of enhanced evaporation decreased the capital costs of evaporation ponds by 46% and increased the operating costs by 330%. The area reduction was by a factor of 5.2. In this case the capital cost of the enhanced evaporation system is significantly more per acre, but this is more than compensated for by the reduced acreage required.

This approach for the Colorado example was judged not to be feasible on the basis of the large land requirement.

6.1 Suggested approach to initial evaluation of enhanced evaporation ponds

- The same procedures as used with conventional evaporation ponds should be followed.

Appendix D

Units conversion table

Category	Unit	Conversion factor multiplier	SI Unit
Length	inch	0.0254	meter
	foot	0.3408	meter
	yard	0.9144	meter
Area	inch^2	0.0006	meter^2
	foot^2	0.0929	meter^2
	yard^2	0.8361	meter^2
Volume	litter	0.0010	meter^3
	gallon	3.785E-3	meter^3
	acre-foot	1,233.5	meter^3
Density	lb/ft^3	16.02	kg/m^3
	lb/gallon	27,700	kg/m^3
Flow rate	cubic foot/sec	0.0283	meter^3/sec
	cubic foot/min	4.72E-4	meter^3/sec
	gallons/sec	3.785E-3	meter^3/sec
	gallons/min	6.308E-5	meter^3/sec
Pressure	bar	1.000E+5	newton/meter^2 (Pa)
	standard atm	1.013E+5	newton/meter^2 (Pa)
	lbf/in^2	6,894	newton/meter^2 (Pa)
	in H_2O	249.1	newton/meter^2 (Pa)
	in Hg	3,386	newton/meter^2 (Pa)
	mm Hg	133.3	newton/meter^2 (Pa)

Category	Unit	Conversion factor multiplier	SI Unit
Energy	kWh	3.600E+6	Joule
	hp/h	2.685E+6	Joule
	btu	1,055	Joule
Power	kW	1,000	Watt
	hp	745.8	Watt
	btu/s	1,055	Watt
Additional units			Additional units
Flux	gallon/ft^2-d	1.70	liter/m^2/h
Volume	acre-ft	325,851	gallon
Flow rate	acre-ft/yr	3.380	m^3/d
	gallon/min (gpm)	0.227	m^3/h
	gallon/day (gpd)	3.785E-3	m^3/d
Pressure	poun/in^2 (psi)	0.06895	bar
	bar	14.50	poun/in^2 (psi)
Specific flux	gfd/psi	24.85	l/m^2/h/bar
Temperature	Celsius (°C)	1.8 (°C +32)	Fahrenheit (°F)