

The release of CO₂ in MSF and ME distillers and its use for the recarbonation of the distillate: a comparison

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Abstract

The knowledge of the CO₂ release from the evaporating brine in multi-stage flash (MSF) and multiple-effect (ME) distillers is very important for the design, the operation, and the cost of desalination distillers as well as for the recarbonation of the distillate. The CO₂ released in MSF plants can be used for the recarbonation of the distillate. The recarbonation method using CO₂ from the vent gases of MSF plants has gained importance, since the post-treatment of potable water can be effected without additional CO₂ production by combustion processes. The paper discusses if this approach may also be applied to ME distillers. The amounts of CO₂ required for the recarbonation are compared with the amounts released. To this end, the CO₂ release rates were calculated on the basis of two previously developed models. One consequence of the differences between MSF and ME systems regarding evaporation mechanisms, fluid flow, operating temperatures, and concentrations is that in MSF distillers the specific phase interface areas are lower and the mass transfer coefficients are higher compared to ME distillers. Consequently, the specific CO₂ release in MSF plants is significantly higher than in ME plants. The total CO₂ release in MSF recycle distillers amounts to about 82 g CO₂ per ton of distillate at a top brine temperature of 100°C. The calculations show that, for the limestone/carbon dioxide method, MSF recycle distillers produce sufficient CO₂ to allow recarbonation of the distillate up to an alkalinity of 187 ppm as CaCO₃ at a top brine temperature of 100°C. The total CO₂ release in ME distillers amounts to about 55 g CO₂ per ton of distillate at a top brine temperature of 64°C. Utilizing the CO₂ for the limestone recarbonation process would result in an alkalinity in the distillate of 124 ppm as CaCO₃. Provided that the alkalinity values aimed at in the distillate range from 50 to 80 ppm as CaCO₃, the calculations suggest that the CO₂ release in ME distillers may be just sufficient in terms of stoichiometric consumption to allow for recarbonation using the limestone method.

Keywords: CO₂ release; MSF; MED; Recarbonation; Post-treatment

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1. Introduction

The distillate produced by thermal desalination plants is usually characterized by very low salinity, in the range of few ppm, and low pH. Therefore, it is highly corrosive, unpalatable and unhealthy. Recarbonation is necessary to overcome these problems. A number of different recarbonation processes are available for the distillate of seawater desalination plants. The dissolution of limestone or lime by using carbon dioxide is widely used.

The CO₂ released in multi-stage flash (MSF) plants can be used for the recarbonation of the distillate [1–4]. The recarbonation method using CO₂ from the vent gases of MSF plants has gained importance, since the post-treatment of potable water can be effected without additional CO₂ production by combustion processes. In the following, it is discussed if this approach may also be applied to multiple-effect (ME) distillers. The CO₂ release rates in MSF and ME distillers as well as main differences in the release process are discussed. To this end, the CO₂ release rates were calculated on the basis of two previously developed models [5–9]. The amounts of CO₂ required for the recarbonation are compared with the amounts released.

2. Recarbonation processes for the distillate

The distillate produced by seawater evaporation plants contains salts and dissolved gases including CO₂, O₂ and N₂ only in very low concentrations. Its salt content amounts to about 25 ppm, but may also be considerably below this value, depending on the age and operation conditions of the plant. It normally has a pH of around 6.0. For practical purposes, the contents of alkalinity and carbon dioxide in the distillate are negligible [1]. The composition of the distillate is not

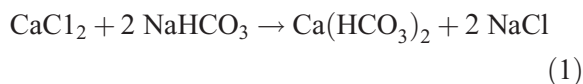
significantly influenced by the salinity of the seawater and the product yield.

The high purity renders the distillate to be highly reactive and aggressive towards nearly all components in the water distribution system, resulting in very severe corrosion problems. One of the by-products of this chemical attack is ferric hydroxide, a red-brown rust, which results in what is called 'red water' [3]. Moreover, water containing a reduced amount of minerals has poor taste and may create health problems. In order to overcome the problems with corrosiveness and poor taste of the distillate, a number of post-treatment processes have been proposed and practiced [1–4, 10–13].

Post-treatment processes include a recarbonation/alkalinization step for elevating carbonate and calcium hardness and pH correction in such a way that carbonate-CO₂ equilibrium is largely achieved. Recarbonation is then followed by further conditioning steps such as dosing, e.g., phosphates to promote the formation and build-up of corrosion protection layers, aeration and disinfection by chlorine dosing to ensure proper hygienic conditions for storage and distribution of the drinking water [1, 2].

Various types of processes for the recarbonation of desalinated water have been proposed [1, 2]. The main recarbonation processes are as follows [1, 2, 10, 11]:

1. Dosage of chemical solutions, i.e. direct injection of suitable chemicals such as calcium chloride and sodium bicarbonate as per reaction:



This remineralization method is a viable option only for small capacity plants because of its relatively high cost of chemicals and the

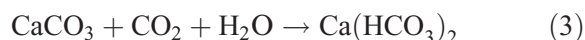
difficulty in handling large quantities of chemicals.

2. Lime dissolution by using carbon dioxide. This process involves treatment of CO₂ acidified distillate with milk of lime. The reaction involved is:



3. Limestone dissolution by using carbon dioxide.

Contacting limestone with CO₂ acidified distillate as per reaction:



The limestone dissolution is the most widely used method because limestone is cheaper than lime and it requires stoichiometrically half the CO₂ amount as compared to the lime method to produce the same alkalinity. The equipment for handling limestone is much cheaper compared to the system required for preparing and dosing lime slurries. The only advantage of the lime process is that the reaction proceeds almost to completion, whereas in the limestone process the reaction is much slower and does not reach completion so that residual excess CO₂ has to be neutralized by addition of NaOH or Na₂CO₃. For large capacity plants, a degasifier is used to strip out the excess CO₂ [10].

The carbon dioxide, required for the recarbonation process according to Eq. (2) or (3), can be made available in different ways depending on the local conditions [1, 2]:

- as liquid CO₂ from outside suppliers, using suitable dosing equipment
- from in-plant production by combustion processes. CO₂ generation by the combustion of oil is successfully practiced in a number of desalination plants (e.g. [12]). The CO₂ present in the flue gas is first scrubbed with seawater, being followed in some cases by post-scrubbing with soda solution. Then, the CO₂

is absorbed in MEA (monoethanolamine) solution. Heating of the MEA/CO₂ solution causes the carbon dioxide to be desorbed in higher purity, whereupon it is subjected to scrubbing with sodium or potassium permanganate and potable water. After further purification in an activated carbon filter stage it is compressed and split into two streams. The main flow is directed to the distillate recarbonation stage. Surplus CO₂ produced is liquefied and stored [2].

- if MSF plants are employed, CO₂ can be obtained from their vent gases. The O₂ and N₂ released in the deaerator as well as the CO₂ released from the brine during evaporation together with the air from leaks is drawn off from the MSF plant by the vacuum system. The water vapor present in the vent extractions is condensed in the surface condensers of the vacuum system. The CO₂-air mixture is then compressed by gas compressors and fed to gas purification filters filled with activated carbon before entering the recarbonation unit.

The aim of the recarbonation process is an increase in calcium hardness and alkalinity of the water to a desirable level and to adjust the pH so that the water becomes well buffered and non-aggressive with a slightly positive Saturation Index (preferably between +0.1 and +0.3) [12]. Thus, the treated water can form a thin, corrosion-inhibiting protective layer of calcium carbonate in the water distribution system. The minimum alkalinity level is 40–50 ppm as CaCO₃ [3, 12, 14]. In product waters obtained from seawater desalination plants the alkalinity is normally increased to a value between 50 and 80 ppm as CaCO₃ [1, 4, 11, 15].

3. Comparison of the CO₂ release in MSF and ME distillers

In MSF distillers and in ME distillers, CO₂ is released from the brine during the

evaporation process and has to be removed by adequate venting. When the brine enters the flash chambers in MSF distillers and the horizontal tube evaporators in ME distillers respectively, the solubility of CO_2 suddenly decreases due to the pressure drop. CO_2 is released into the vapor phase. The release of CO_2 is difficult to predict as it is not only molecularly dissolved in seawater, but mainly chemically bound in HCO_3^- and CO_3^{2-} ions. The liberation of CO_2 impairs the equilibria between HCO_3^- , CO_3^{2-} , CO_2 , H^+ and OH^- in the brine. New molecular CO_2 is produced by chemical reactions and released subsequently. The release of CO_2 is a complex process involving chemical reaction kinetics, mass transfer processes as well as phase equilibria at the brine/vapor interface.

In order to utilize the CO_2 present in the vent gas for the recarbonation of the distillate, it is essential to know how much CO_2 is released in the individual distiller stages at different operating conditions. To this end, the CO_2 release rates were calculated on the basis of two previously developed models [5–9]. The models describe the reactions and mass transfer processes in the brine as well as the CO_2 release in the individual stages of MSF and ME distillers. In MSF distillers, the flow path of brine through the condensers, the brine heater and the flash chambers is considered. In ME distillers, the flow path of brine through the final condenser and the evaporator stages is included. The desorption of CO_2 is treated as a problem of coupled mass transfer with chemical reaction. Depending on the desorption regime, the mass transfer may be enhanced by the chemical reaction.

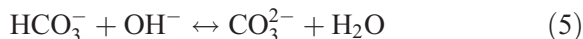
In the following, the main principles of the model for the simulation of the CO_2 release and the carbonate system in MSF and ME distillers are outlined. A detailed description of the models and the chemical reaction

kinetics involved in CO_2 release can be found in [5–9].

3.1. Modelling of the CO_2 release and the carbonate system

For describing the CO_2 release and the carbonate system, the flash chambers in MSF distillers and the liquid film flowing over the horizontal tubes in ME distillers are divided into volume elements. Precipitation of CaCO_3 and $\text{Mg}(\text{OH})_2$ is assumed to occur at negligible rates. Total alkalinity TA and total inorganic carbon TC at the outlet of a volume element are calculated by means of mole balances. Total alkalinity is only influenced by the evaporation of water. The loss of CO_2 has no effect. Total inorganic carbon TC is affected by both the evaporation of water and the release of CO_2 . After determining the total alkalinity and the total inorganic carbon, the remaining quantities CO_2 , HCO_3^- , CO_3^{2-} and pH at the outlet of a volume element can be calculated by applying the law of mass action.

It was found that in the temperature and pH range prevailing in the flash chambers of MSF distillers and in the evaporators of ME distillers the alkaline reaction mechanism



with the rate-determining step (4) predominates [5, 8, 9]. For modelling the mass transfer with chemical reaction, it can be assumed that the OH^- ions are present in excess so that their concentrations in the boundary layer at the gas/liquid interface are not considerably changed by the reactions and can be considered to remain constant [5, 8, 9]. Reaction (4) can be assumed to be pseudo first order with the rate constant

$$k_1 = k_2 C_{OH^-} \quad (6)$$

where k_2 is the second order rate constant of the forward reaction (4) and C_{OH^-} is the concentration of the OH^- ions.

It was found that in MSF distillers the main portion of the CO_2 is released during bubble evaporation in the front part of the flash chambers [5, 6]. At the end of the flash chambers, where most evaporation occurs at the brine surface, the CO_2 release rates are low because the interfacial area and the concentration difference of CO_2 between bulk and phase interface, the driving force for mass transfer, are small.

In the front part of the flash chambers, where bubble evaporation occurs, the rate-determining reaction is slower than the mass transfer. The reactions are, however, fast enough to keep the bulk concentrations in chemical equilibrium. Thus, the desorption takes place in the diffusional regime, a sub-regime of the slow reaction regime. The reactions occur only in the bulk flow, not in the boundary layer at the phase interface. The mass transfer in the boundary layer is not enhanced by the reactions. The desorption rate is given by

$$\dot{N}_{CO_2} = k_L^\circ A_{Ph} \left(C_{CO_2,B}^{eq} - C_{CO_2,Ph}^* \right) \quad (7)$$

where k_L° is the mass transfer coefficient in the liquid phase, A_{Ph} is the phase interface area, $C_{CO_2,B}^{eq}$ is the chemical equilibrium concentration of CO_2 in the liquid bulk, and $C_{CO_2,Ph}^*$ is the CO_2 concentration at the phase interface.

In ME distillers, desorption takes place in the transition regime between slow and fast reactions [8, 9]. Mass transfer and chemical reactions simultaneously take place in the boundary layer. The bulk of the brine is in

chemical equilibrium. In the transition regime between slow and fast reactions the chemical rate constant k_1 becomes more important and the mass transfer coefficient becomes less important. The desorption rate is given by

$$\dot{N}_{CO_2} = k_L^\circ A_{Ph} \frac{(1 + K) \left(C_{CO_2,B}^{eq} - C_{CO_2,Ph}^* \right)}{1 + K \tanh \left(Ha \sqrt{1 + \frac{1}{K}} \right) / \left(Ha \sqrt{1 + \frac{1}{K}} \right)} \quad (8)$$

with the chemical equilibrium constant of reaction (4) considered to be pseudo first order

$$K = \frac{K_1}{K_w} [OH^-]^{eq} \quad (9)$$

and the Hatta number

$$Ha = \frac{\sqrt{D_{CO_2} k_1}}{k_L^\circ} \quad (10)$$

where D_{CO_2} is the diffusivity of CO_2 in seawater and k_1 is the chemical rate constant of the pseudo first order reaction due to (6).

Computer programs were written for simulating the CO_2 release and the carbonate system in MSF and ME distillers. Given the main dimensions of the MSF or the ME distiller, the operating data and the measurable pH and total alkalinity of seawater, the programs simulate the CO_2 release rates as well as the HCO_3^- , CO_3^{2-} , CO_2 , H^+ and OH^- concentrations in the brine on its flow path through the distillers. In order to verify the model for the CO_2 release in MSF distillers, the release of CO_2 was experimentally investigated at two MSF recycle distillers in the Arabian Gulf. The mean difference between model and experiment was 15%, the maximum difference was 30 % [5, 16].

3.2. Simulation results

The models were applied to

- a 20-stage MSF recycle distiller with a distillate production of 1290 t/h at a top brine temperature of 100°C,
- a 20-stage MSF once-through distiller with a distillate production of 1290 t/h at a top brine temperature of 100°C and
- a 5-stage multiple-effect distiller with a distillate production of 380 t/h at a top brine temperature of 64°C.

A variety of multiple-effect process configurations have been installed. For modeling the CO₂ release process, the configuration shown in Fig. 1 was chosen as a reference. After leaving the final condenser the feed water flow is distributed in equal shares to the stages and trickles down the horizontal tube bundles. The feed water is preheated to the evaporation temperature on the upper tube rows and then part of it is evaporated on the lower tube rows.

A more concise description of the reference distillers can be found in [5, 9, 17]. In the following, some selected simulation results are presented.

Fig. 2 shows the release of CO₂ in the individual stages of the reference MSF recycle and once-through distiller and the ME distiller. The desorption rates are related to the total distillate flow rates.

In MSF distillers, the CO₂ release notably decreases from the first to the last stage. The

brine successively flows from the first to the last stage and the proceeding CO₂ release reduces the total inorganic carbon content and increases the pH value. Thus, the concentration difference of CO₂ in the brine between bulk flow and phase interface drops. Additionally, the mass transfer coefficient k_L and the interfacial area A_{Ph} decrease.

The absolute CO₂ release rates are considerably higher in the once-through distiller than in the recycle distiller. Since part of the brine of the last stage is mixed with the make-up flow in recycle distillers, the total inorganic carbon content in the recirculation flow is lower, the pH is higher and the CO₂ concentrations in the brine bulk are lower than in once-through distillers. Thus, the driving concentration difference for mass transfer is lower.

As shown in Fig. 2, in ME distillers, the decrease in the CO₂ release from the first to the last stage is less pronounced, because the feed water flow is distributed in parallel to the stages and, thus, the carbonate system in the brine entering each stage is the same. The CO₂ release decreases, because the mass transfer coefficient and the concentration difference of CO₂ in the brine between bulk flow and phase interface drop with decreasing temperature.

The effect of the top brine temperature on CO₂ desorption in MSF distillers is shown in Fig. 3. In the high temperature stages, the specific CO₂ release significantly increases

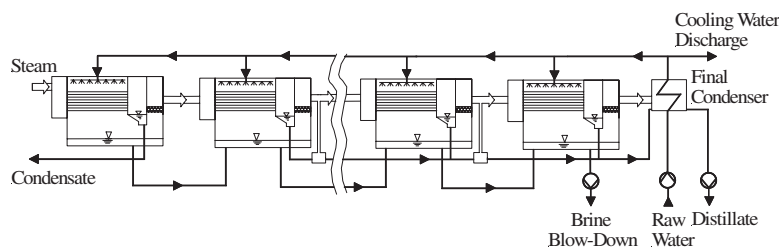


Fig. 1. Process configuration of the reference multiple-effect distiller.

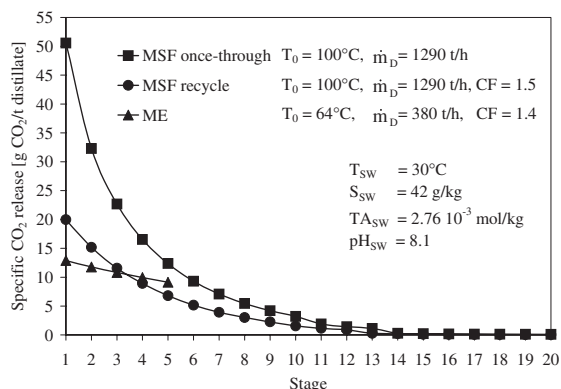


Fig. 2. Specific CO₂ release (related to the total distillate flow rate) in the individual stages of the reference ME distiller, MSF recycle distiller and once-through distiller.

with increasing top brine temperature, because the surface area of the vapor bubbles and the mass transfer coefficient rise. The difference between CO₂ concentration in the bulk and at the phase interface, the driving force for mass transfer, increases with the top brine temperature.

The more CO₂ is released in the first stages, the less total inorganic carbon remains in the brine. Accordingly, the CO₂ concentration in the bulk of the brine and the driving concentration difference for mass transfer

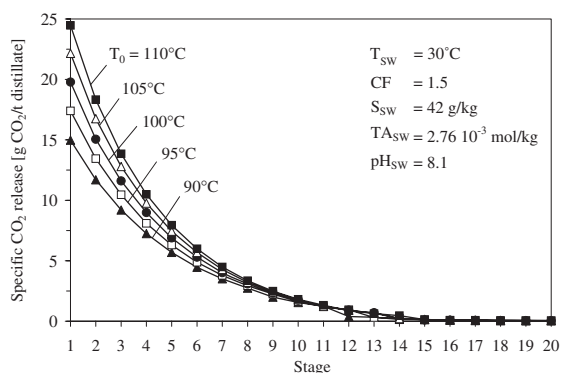


Fig. 3. The effect of top brine temperature T₀ on the specific CO₂ release (related to the total distillate flow rate) in the reference MSF recycle distiller.

decrease. Thus, the CO₂ release profile becomes steeper with increasing top brine temperature.

Fig. 4 shows the effect of the top brine temperature on the specific CO₂ release in the individual stages of the reference ME distiller. Similar to MSF distillers, the CO₂ release rises with increasing top brine temperature due to an increased mass transfer coefficient and a higher driving concentration difference.

3.3. Discussion of the underlying mechanisms

There are decisive differences between MSF and ME systems regarding evaporation mechanisms, fluid flow, concentrations, operating temperatures, and pressures which affect CO₂ release. In the following, the main differences in the release process in MSF and ME distillers are discussed. Therefore, the ME distiller is compared with the once-through distiller, because in both distillers the brine is not recirculated. As shown in Fig. 2, the brine recirculation has a great impact on the CO₂ release because the total inorganic carbon content in the recirculation flow is reduced. Thus, the CO₂ concentration in the bulk of the brine and the driving

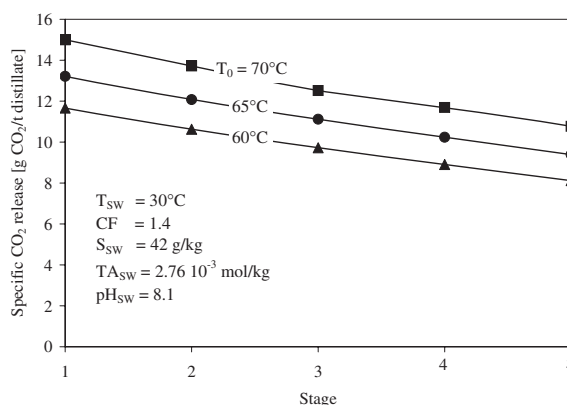


Fig. 4. The effect of top brine temperature T₀ on the specific CO₂ release (related to the total distillate flow rate) in the reference ME distiller.

concentration difference for mass transfer decrease.

In MSF distillers, the brine flows through the flash chambers and flash evaporation occurs. CO_2 is released into the vapor bubbles as well as into the vapor space above the brine surface, as shown in Fig. 5. The interfacial area is the sum of the surface area of the vapor bubbles plus the area of the free brine surface including waves and spray jets. Both the growth rates of the bubbles in the superheated brine and their rise velocity are determined in the model. The accumulation of CO_2 in the bubbles on their way to the brine surface is also considered.

In ME distillers, the brine is flowing over a bundle of horizontal tubes and surface evaporation occurs, as shown in Fig. 6. CO_2 is liberated into the vapor space above the brine film. The phase interface area is the sum of the surface area of the liquid film on the tubes plus the surface area of the liquid between the tubes. For the description of the phase interface area in the model, it is assumed that predominantly liquid jets occur. It is assumed that the CO_2 is continuously removed with the ascending vapour so

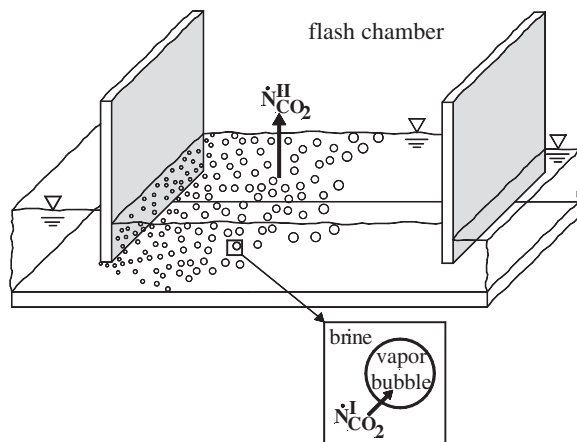


Fig. 5. Brine flow through flash chambers in MSF distillers.

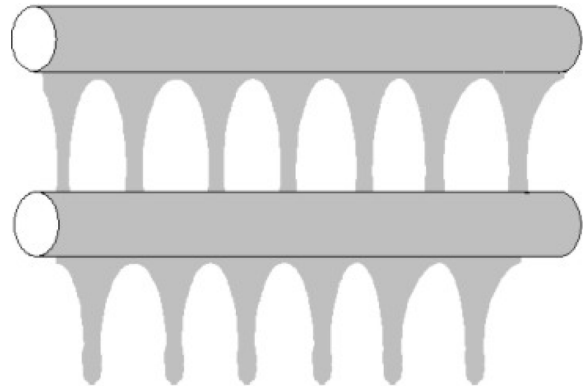


Fig. 6. Brine flow over horizontal tubes in ME distillers.

that no accumulation occurs in the gas phase over the brine.

In MSF distillers, the surface area of the vapor bubbles and, thus, the phase interface area significantly decrease from the first to the last stage. The effective specific phase interface area (related to the total distillate flow rate of the distiller) in the reference MSF once-through distiller at a top brine temperature of 100°C amounts to $1.3 \text{ m}^2/(\text{t/h})$ in the front part of the first stage, $0.2 \text{ m}^2/(\text{t/h})$ in the tenth stage, and $0.01 \text{ m}^2/(\text{t/h})$ in the front part of the last stage. Assuming that the heat transfer area is the same in all the stages of the ME distiller, the phase interface area changes only slightly from the first to the last stage. In the reference ME distiller at a top brine temperature of 64°C , the phase interface area is about $16.4 \text{ m}^2/(\text{t/h})$ in each stage. Thus, the specific phase interface area is significantly higher in the ME distiller than in the MSF distiller.

Mass transfer in liquid films flowing over horizontal tubes differs from mass transfer in open channel flow with flash evaporation. In the front part of the flash chambers of MSF distillers, where bubble evaporation occurs, mass transfer coefficients are very high. At the end of the flash chambers, where surface

evaporation predominates, the mass transfer coefficients drop to much lower values. In the reference MSF distiller at a top brine temperature of 100°C, the average mass transfer coefficient amounts to about $4.1 \cdot 10^{-2}$ m/s in the front part of the first flash chamber where the main portion of the CO₂ is released, about $2.6 \cdot 10^{-2}$ m/s in the front part of the tenth stage, and about $1.4 \cdot 10^{-2}$ m/s in the last stage. The average mass transfer coefficient in the first stage of the reference ME distiller at a top brine temperature of 64°C is about $4.6 \cdot 10^{-4}$ m/s, $4.0 \cdot 10^{-4}$ m/s in the third stage, and $3.5 \cdot 10^{-4}$ m/s in the last stage. Thus, the mass transfer coefficients in the ME distiller are considerably lower than the ones in the MSF distiller.

As described above, it was found that in ME distillers the desorption occurs in the transition regime between slow and fast reactions where the chemical rate constant becomes more important and the mass transfer coefficient becomes less important. Mass transfer is slightly enhanced by the chemical reaction which compensates for the decreasing influence of the mass transfer coefficient.

The CO₂ concentrations in the brine bulk at the inlet of the stages in the MSF distiller are different from the ones in the ME distiller due to the following reasons:

- It is assumed that there is no CO₂ release during preheating of the brine in the condensers of the MSF distiller and in the final condenser of the ME distiller respectively. The concentrations in the carbonate system change due to the changes in the dissociation constants of carbonic acid in seawater with increasing temperature.
- Since the stages in MSF distillers are fed in series, the total inorganic carbon content in the brine decreases and the pH value increases from the first to the last stage. The CO₂ concentration in the bulk of the brine decreases. Due to the parallel feed

water flow to the stages in the ME distiller, the concentrations in the carbonate system of the brine at the inlet of the stages are the same.

The CO₂ concentrations in the brine at the phase interface depend on the partial pressure of CO₂ in the vapor phase and, thus, on the total pressure in the stage and the absolute CO₂ release rates. They further depend on Henry's law coefficient which varies with the temperature and the salinity of the brine.

Furthermore, in MSF distillers, CO₂ accumulates in the vapor bubbles on their way to the brine surface and, thus, the CO₂ concentration in the vapor bubbles rises. The accumulation of CO₂ results in higher CO₂ concentrations in the brine at the phase interface. It follows that the driving concentration difference for mass transfer is reduced. To show the effect of the CO₂ accumulation in the vapor bubbles, the CO₂ release was calculated for the fictitious case that there is no CO₂ accumulation in the bubbles on their way to the brine surface, i.e. the surface area of the vapor bubbles was considered as free surface area. In this fictitious case the specific CO₂ release in the first stage of the reference MSF once-through distiller would be almost 12 times higher than the release in the first stage of the ME distiller.

The mean difference between CO₂ concentration in the brine bulk and at the phase interface, the driving force for mass transfer, amounts to $4.1 \cdot 10^{-3}$ mol/m³ in the front part of the first stage of the reference MSF once-through distiller, $3.4 \cdot 10^{-3}$ mol/m³ in the front part of the tenth stage and $2.3 \cdot 10^{-3}$ mol/m³ in the last stage. In the reference ME distiller, the mean driving concentration difference is about $9.4 \cdot 10^{-3}$ mol/m³ in the first stage, $9.2 \cdot 10^{-3}$ mol/m³ in the third stage and $9.0 \cdot 10^{-3}$ mol/m³ in the fifth stage.

The overall effect of the differences described above is that the specific total

CO₂ release in the reference MSF once-through distiller at a top brine temperature of 100°C is 3 times higher than the CO₂ release in the ME distiller at 64°C.

4. Utilization of vent gas from MSF and ME plants for distillate recarbonation

In view of the differences in CO₂ release between ME and MSF distillers, it was studied whether the CO₂ released in ME distillers could be used for the recarbonation of the distillate as it is successfully practiced in a number of MSF plants. It was determined which alkalinity would be reached in the distillate when using the CO₂ released for the recarbonation with the limestone method. The amounts of CO₂ released were based on the simulation results described above. A difficulty in the design of limestone beds is the lack of reliable data on the kinetics of dissolution of limestone by CO₂ acidified water [10]. In the frame of this study only the stoichiometric consumption of CO₂ as per reaction (3) was considered and taken as minimum demand.

Fig. 7 shows the specific CO₂ release in ME and MSF distillers and the alkalinity of the distillate

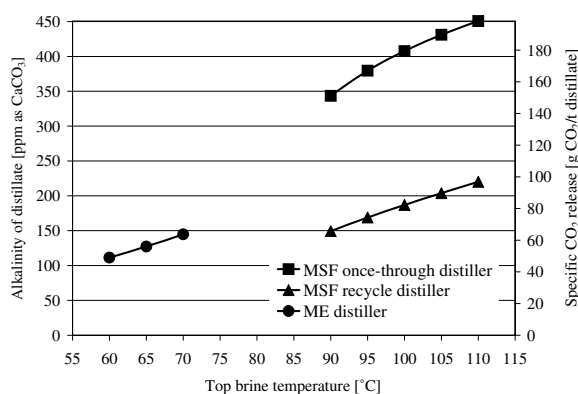


Fig. 7. Specific total CO₂ release in MSF and ME distillers and the resulting alkalinity of the distillate after recarbonation using the limestone method.

the distillate that would be obtained when utilizing the released CO₂ for the recarbonation.

The total CO₂ release in MSF recycle distillers increases from about 66 g CO₂ per ton of distillate at a top brine temperature of 90°C to about 97 g/t at 110°C. The calculations show that, for the limestone/carbon dioxide method, MSF recycle distillers produce sufficient CO₂ to allow recarbonation as per reaction (3) up to an alkalinity of 149 ppm as CaCO₃ at a top brine temperature of 90°C and up to an alkalinity of 220 ppm as CaCO₃ at 110°C. Provided that the alkalinity values aimed at in the distillate range from 50 to 80 ppm as CaCO₃, this method provides an adequate reserve in terms of CO₂ production. The calculations further suggest that, for the lime method requiring twice as much CO₂ as compared to the limestone process, the CO₂ release is sufficient only at higher top brine temperatures.

The total CO₂ release in ME distillers increases from about 49 g CO₂ per ton of distillate at a top brine temperature of 60°C to about 64 g/t at 70°C. Provided that the alkalinity values aimed at in the distillate range from 50 to 80 ppm as CaCO₃, the calculations suggest that the CO₂ release in ME distillers may be just sufficient in terms of stoichiometric consumption to allow for recarbonation using the limestone method. For the lime dissolution method, however, the CO₂ produced in ME distillers operating at low temperatures is not sufficient.

5. Conclusion

In order to utilize the CO₂ present in the vent gas of thermal seawater desalination plants for the recarbonation of the distillate based on the limestone or lime dissolution methods, it is essential to know how much

CO₂ is released in the individual distiller stages at different operating conditions. To this end, previously developed models that describe the reactions and mass transfer processes in the brine as well as the CO₂ release in the individual stages of MSF and ME distillers proved as valuable tools. The CO₂ release calculated for typical operating conditions of industrial distillers was compared with the amounts stoichiometrically required for the recarbonation of the distillate.

The calculations confirmed that the CO₂ released in MSF distillers at different operating conditions can be used for the recarbonation of the distillate. The calculations further suggest that the utilization of the CO₂ released in ME distillers for recarbonation using the limestone method seems feasible in principle. For a practical realization it must be examined if the CO₂ release covers the actual CO₂ demand especially also under consideration of kinetic aspects.

6. Symbols

| | |
|----------|---|
| A_{Ph} | Phase interface area, m ² |
| C_i | Concentration of the component i, mol/m ³ |
| CF | Concentration factor |
| D_i | Diffusion coefficient of the component i, m ² /s |
| Ha | Hatta number |
| [i] | Concentration of the component i, mol/kg solution |
| K | Equilibrium constant of the pseudo first order reaction |
| K_1 | First dissociation constant of carbonic acid in seawater on the basis mol/kg solution |
| K_w | Dissociation constant of water in seawater on the basis mol/kg solution |
| k_1 | First order rate constant of the forward reaction, 1/s |

| | |
|------------------|---|
| k_2 | Second order rate constant, m ³ /(mol s) |
| k_L | Mass transfer coefficient in liquid phase, m/s |
| \dot{m} | Mass flow rate, kg/s |
| \dot{N}_{CO_2} | Molar desorption rate of carbon dioxide, mol/s |
| pH | pH value |
| S | Salinity, g/kg |
| T_0 | Top brine temperature, °C |
| T_{SW} | Seawater temperature, °C |
| TA | Total alkalinity, mol/kg |
| TC | Total inorganic carbon, mol/kg |
| Indices | |
| B | Bulk |
| CO ₂ | Carbon dioxide |
| D | Distillate |
| eq | Chemical equilibrium |
| L | Liquid side |
| OH ⁻ | Hydroxide ion |
| Ph | Phase interface |
| SW | Seawater |
| * | Physical equilibrium |

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References

- [1] H. Ludwig, Post-treatment and potabilisation – A quality determining factor in the generation of drinking water by means of sea water desalination. Proc. IDA Congress on Desalination and Water Reuse, Madrid, Spain, Vol. 4 (1997) 15–36.
- [2] H. Ludwig and M. Hetschel, Treatment of distillates and permeates from seawater desalination plants. Desalination, 58 (1986) 135–154.
- [3] H.E. Al-Rqobah and A. Al-Munayyis, A recarbonation process for treatment of distilled water

- produced by MSF plants in Kuwait. *Desalination*, 73 (1989) 295–312.
- [4] Y. Yamauchi, K. Tanaka, K. Hattori, M. Kondo and N. Ukawa, Remineralization of desalinated water by limestone dissolution filter, *Desalination*, 66 (1987) 365–383.
- [5] H. Glade, Transport und Reaktion von Kohlendioxid in Entspannungsverdampfern zur Meerwasserentsalzung. VDI Fortschritt-Berichte, Series 3, No. 699, VDI Verlag, Düsseldorf, 2001.
- [6] H. Glade and K. Genthner, The carbonate system in MSF distillers. Proc. IDA World Congress on Desalination and Water Reuse, Bahrain, 2002.
- [7] H. Glade, Chemical reaction kinetics and mass transfer phenomena controlling the release of CO₂ in MSF distillers, Proc. IDA World Congress on Desalination and Water Reuse, San Diego, USA, 1999, Vol. 1, 375–388.
- [8] A.E. Al-Rawajfeh, H. Glade and J. Ulrich, CO₂ release in multiple-effect distillers controlled by mass transfer with chemical reaction. *Desalination*, 156 (2003) 109–123.
- [9] A.E. Al-Rawajfeh, Modelling and simulation of CO₂ release in multiple-effect distillers for seawater desalination. PhD Thesis, Martin-Luther-University of Halle-Wittenberg, Shaker-Verlag, Aachen, 2004.
- [10] D. Hasson, Modeling remineralization of desalinated water by limestone dissolution. Proc. IDA World Congress on Desalination and Water Reuse, Bahrain, 2002.
- [11] G. Odone, Recarbonation systems for desalination plants. EDS Newsletter, Issue 16 (2002) 6–7.
- [12] A.D. Khawaji and J.M. Wie, Potabilization of desalinated water at Madinat Yanbu Al-Sinaiyah. *Desalination*, 98 (1994) 135–146.
- [13] E. Gabbrielli, A tailored process for remineralization and potabilization of desalinated water. *Desalination*, 39 (1981) 503–520.
- [14] N.A. Nada, A. Zahrani and B. Ericsson, Experience on pre- and post-treatment from sea water desalination plants in Saudi Arabia. *Desalination*, 66 (1987) 303–318.
- [15] P. Leroy, Corrosion control of feeders carrying potable water from desalination of sea water. *Desalination*, 44 (1983) 317–324.
- [16] H. Glade and K. Genthner, Measurements of the CO₂ release rates in the individual stages of MSF distillers. Proc. IDA World Congress on Desalination and Water Reuse, San Diego, USA, Vol. 1 (1999) 361–373.
- [17] A.E. Al-Rawajfeh, H. Glade, H.M. Qiblawey and J. Ulrich, Simulation of CO₂ release in multiple-effect distillers. *Desalination*, 166 (2004) 41–52.