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Domestic wastewater treatment with membrane filtration — two years experience

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Abstract

This study tested domestic wastewater treatment membrane filtration without external cleaning in sustained long term operation. Domestic wastewater treatment plant monitoring was performed at the municipal wastewater treatment plant Devínska Nová Ves, Bratislava between February 2005 and July 2007. Two membrane modules were tested by immersion in the domestic wastewater treatment plant. The flat sheet membrane module was operated without external cleaning at a flux of 20–60 L/m^2 h for 6 months. The hollow fiber membrane module was operated for 4 months without external cleaning with a flux of 20–45 L/m^2 h. Parallel operation of flat sheet and hollow fiber membrane modules showed similar results in effluent water quality. Both membrane modules were able to effectively remove organic matter (as much as 91%) and more than 97% of NH_4^+–N. Nitrogen removal via denitrification was observed during the short periods with low oxygen concentration. Treated water contained suspended solids under measurable limits.

Keywords: Domestic wastewater treatment plant; Immersed membrane modules; Sewage; Long term operation; Nitrogen removal

1. Introduction

European legislation on wastewater effluent discharge has led to a need for enhanced treatment processes capable of removing high percentages of BOD_5, suspended solids, nitrogen, phosphorus, bacteria, etc. One of the most promising technologies capable of fulfilling these requirements is the membrane filtration process. Combining membrane technology with biological reactors for wastewater treatment has led to the development of membrane bioreactors (MBRs). Ultrafiltration as a replacement for

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secondary sedimentation tanks in the activated sludge process was first described in 1969 by Ref. [1] and since then MBRs have been successfully used worldwide in industrial and municipal wastewater treatment in hundreds of applications.

One of the main limitations to extensive MBR usage in wastewater treatment is membrane fouling, because the membrane is challenged with very high total solids concentration arising from concentrated biomass. This high solids concentration coupled with varying levels of colloidal and dissolved extracellular polymeric substances (EPS) are the key challenges to the MBRs processes [2–4]. In the submerged processes generally there are three strategies to limiting fouling: (a) reducing flux, (b) increasing membrane aeration or (c) employing physical or chemical cleaning [2,5].

Installation and testing of MBRs in domestic wastewater treatment plants (less than 10 m^3/day capacity) has tended toward increasing membrane aeration with the goal of long term wastewater treatment plant (WWTP) operation without membrane cleaning. Increasing the aeration rate, and thus cross flow velocity, suppresses fouling and thereby increases flux. Recent studies carried out with submerged MBR [2,5,6] suggest that an increase in air flow rate at the membrane surface limits fouling. However, Ref. [7] observed an optimum aeration rate beyond which a further increase has no effect on fouling suppression. Details of the phenomena occurring during air sparging have been extensively reported in Refs. [8–10].

Slovak Republic entry into the European Union resulted in a marked expansion of sewage and municipal WWTP construction, albeit mostly in regions with more than 10 000 PE. However, due to inadequate experience with the technology, MBRs are not typically considered in designing treatment plant reconstructions. The first applications of MBR in small domestic WWTP are likely in the near future.

Czech and Slovak international cooperation coordinated by ASIO company (CZ) together with Slovak University of Technology Bratislava (SK) and the Institute of Chemical Technology Prague (CZ) has resulted in a long term pilot plant experiments with MBR. The results from the Prague experiments are reported in Ref. [11]. The purpose of this Slovak part of the study was to investigate the long term operation of a domestic WWTP with immersed membrane modules without backwashing or chemical cleaning.

2. Materials and methods

2.1. *Description of the domestic wastewater treatment plant with membrane filtration*

A commercially operated domestic WWTP (ASIO, Czech Republic) with immersed membrane was used for this study. The chosen site was a municipal domestic WWTP Bratislava – Devínska Nová Ves (ca. 35 000 PE) with a total working volume of 1.55 m^3. The wastewater was pumped after passing through fine screens (6 mm) into the pilot plant in amounts of 450–700 L/day (discontinuous in the eight time intervals per day). In the first plant treatment step sedimentation tank with a volume 0.7 m^3 — the particular suspended solids (SS) were settled and accumulated on the bottom of the sedimentation tank. The pretreated wastewater was passed into the biological activated sludge tank equipped with an immersed membrane module (MM) and with a fine-bubble aerator used activated sludge aeration as well as aeration cleaning of the MM. Treated water was pumped through the membrane using a small (12 W) pump with maximum flux capacity of 100 L/m^2 h. A schematic of the treatment system with MM is shown in Fig. 1. Table 1 shows the basic experimental conditions and Table 2 shows

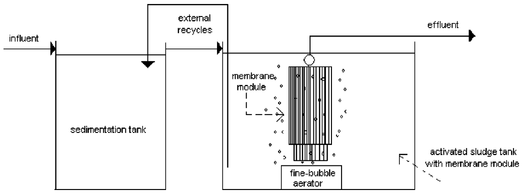

Fig. 1. Schematic of the treatment system with MM.

specifications of the MMs. The long term test of the MBR included five phases.

2.1. Analytical methods

The pilot-scale experiments with MBRs were running from February 2005 to July 2007. During the experiment, basic influent and effluent parameters — temperature, pH, COD, BOD_5, SS, ammonium nitrogen (NH_4^+–N), nitrite nitrogen (NO_2^-–N) and nitrate nitrogen (NO_3^-–N) as well as the activated sludge parameters — were monitored. All these parameters were analyzed using the standard methods [12]. The photometric semimicromethod [13] was used for COD analysis.

3. Results and discussion

3.1. The first phase (February 2005–June 2005)

The flat sheet membrane module (FSMM) was installed in domestic WWTP pilot plant with operation commencing on 14 February 2005. The pilot plant was not inoculated by activated sludge (by agreement with the operator). During the first phase after start-up, it was concluded that rapid clogging of the FSMM could be related to sludge inoculation. Most likely, microscopic, colloid or high-molecular weight particles in raw municipal wastewater caused preterm clogging of the membrane. The freely moving small particles in the activated tank rapidly entered the membrane pores. If the particles are in an environment with higher sludge concentration, it would initiate slower membrane plugging or it would be completely reduced because there would be physical or physical–chemical reactions among particles and sludge flocks. Despite the technical problems, samples were analyzed during the entire first phase. The COD values in the effluent were 12.4–124.3 mg/L (average value is 64.4 mg/L)

Table 1
Basic experimental conditions

Phase	Time period	Inflow rate (L/day)	Main phase goal	Used membrane type	HRT in aeration tank (day)
First	Feb 2005–Jul 2005	360	Start-up without seed sludge	Flat sheet module	1.87
Second	Jul 2005–Sep 2005	450	Start-up with seed sludge, long term operation	Flat sheet module	1.6
	Sep 2005–Jan 2006	700			1.1
Third	Mar 2006–May 2006	480	Parallel operation of two types of membranes	Flat sheet module parallel with hollow fiber module	1.6
	May 2006–Oct 2006	700			1.1
Fourth	Feb 2007–Jul 2007	800	Test of external aeration	Hollow fiber module	1.0
Fifth	Feb 2007–May 2007	7000×10^3	Test of external aeration in the real WWTP	Hollow fiber module	0.5

Table 2
Specifications of the MMs

Phase	First	Second	Third		Fourth	Fifth
Membrane producer	Martin Systems (D)	Martin Systems (D)	Martin Systems (D)	Anonymous (CZ)	Anonymous (CZ)	Anonymous (CZ)
Module configuration	Flat sheet	Flat sheet	Flat sheet	Hollow fiber	Hollow fiber	Hollow fiber
Material of membrane	Organic polymers	Organic polymers	Organic polymers	Organic polymers	Organic polymers	Organic polymers
Pore size (μm)	~0.04	~0.04	~0.04	<0.1	<0.1	<0.1
Surface area of membrane module (m^2)	6	6	6	4	8	48
Place of installed membrane	Activated sludge tank of domestic WWTP	Activated sludge tank of domestic WWTP	Activated sludge tank of domestic WWTP	Activated sludge tank of domestic WWTP	Activated sludge tank of domestic WWTP	Activated sludge tank of WWTP DNV

which represents 85% process efficiency. Nitrification was limited (the values of NH_4^+–N were approximately 33 mg/L during long term operation) and the increase of NO_3^-–N was not significant.

3.2. The second phase (July 2005–January 2006)

After the experiences from the first phase, the second phase was started with sludge inoculation. Fig. 2 shows sludge concentration and flux during the second phase.

The start-up sludge concentration was 0.6 g/L. During the second phase, solids accumulation and biomass growth led to an almost constant sludge concentration 3.0–3.5 g/L. After three months of relatively stable sludge concentration in the activated sludge tank, the pilot plant influent was increased from 450 L/day to 700 L/day (September 2005). This influent increase caused slow but continually increased sludge concentration in the activated sludge tank with values up to 12–13 g/L. One of the reasons for increasing sludge concentration in the system was overflow of primary sludge from the sedimentation tank. This sludge was accumulated and digested in the sedimentation tank during the summer season and was not removed from the system. The flux showed relatively stable values of 60 L/m^2 h and did not change during the first three months of operation. Similarly, as observed by sludge concentration measurements, the measured flux decreased significantly after the increase in influent wastewater. This decrease was probably caused by significant overflow of

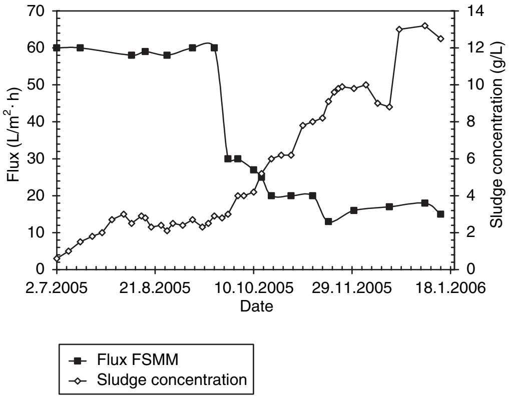

Fig. 2. Sludge concentration and flux FSMM during the second phase.

Table 3
Water qualities of the influent, outflow from sedimentation tank and effluent treated wastewater during the second phase

Concentration (mg/L)	Influent (raw municipal waste water)		Outflow (sedimentation tank)		Effluent (flat sheet membrane module)		
	Average	Scope	Average	Scope	Average	Scope	η (%)
COD	787.8	502.7–1419	157.6	145.6–663.7	45.6	12.4–86.9	94.2
BOD_5	327.4	227.2–420.1	188.5	80.5–333.7	5.8	2.8–12.2	98.2
SS	383.1	60–976.7	120.2	83.0–203.3	<3	–	≈ 100.0
NH_4^+–N	68.6	33.5–134.1	77.1	42.3–164.3	1.5	0.2–14.1	97.1
N_{total}	156.7	91.5–259.5	158.5	88.5–291.2	29.0	2.4–46.0	81.4

digested primary sludge into the activated sludge tank. This sludge had a consistency that was relatively too sticky for efficient membrane filtration. In the next experimental phases, digested sludge was regularly withdrawn from the sedimentation tank.

The main parameters measured to assess influent, outflow from the sedimentation tank and treated wastewater effluent quality are summarized in Table 3. Average influent COD concentration was 787.8 mg/L and average effluent COD was 45.6 mg/L. Parameter COD values in the effluent were relatively high compared with values reported elsewhere [14]. Analysis of COD by the photometric semimicromethod at low COD concentrations is believed to be highly accurate with low susceptibility to error (high precision). During operation of the second phase, there was no excess sludge withdrawal from the system and soluble components of the decayed sludge were slowly released. The elevated effluent measurements could have resulted due to this change in the pilot plant operation.

Nitrification occurred without any problems. High total nitrogen (N_{total}) concentrations in the influent were caused by grab sampling. In addition, wastewater sampling occurred in the morning when urea concentrations are highest; the efficiency of NH_4^+–N removal was 97.1%. Nitrogen removal as N_{total} was relatively efficient. This high efficiency was caused by denitrification in those parts of activated sludge tank which were not sufficiently aerated. Considering the high sludge concentrations and long retention times, denitrification was relatively successful.

3.3. The third phase (March 2006–September 2006)

In the third phase, the pilot plant experiment started with sludge inoculation. Two different membrane modules were inserted in the activated sludge tank. As shown in Fig. 4, during the third phase the sludge concentration continually increased from 0.6 g/L to 12 g/L. The flux values were measured in both MMs. For the FSMM, the starting flux value was 60 L/m^2 h and for the hollow fiber membrane module (HFMM), the starting flux value was 45 L/m^2 h. After the start-up of operation, moderate flux decrease was observed. However, in the entire 6 months of operation a relatively constant flux was observed with a value of more then 45 L/m^2 h for the FSMM and more than 35 L/m^2 h for the HFMM.

During the entire third phase, the FSMM was neither mechanically nor chemically cleaned. On 3 August 2006, the HFMM was cleaned by air back-flush because a significant flux decrease was observed (see Fig. 3). At the end of the third phase, a greater volume of digested primary sludge again entered into the activated sludge

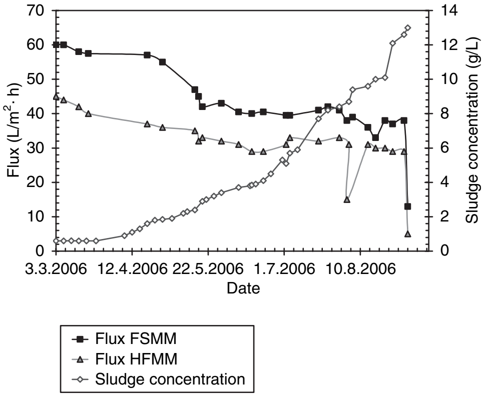

Fig. 3. Sludge concentration and flux FSMM and HFMM during the third phase.

tank, which caused a significant flux decrease in both membranes. At this occurrence, third phase experiments ceased.

Tables 4 and 5 show water quality of the influent, outflow from sedimentation tank and effluent treated wastewater. Average influent COD was 644.4 mg/L. The effluent COD concentration from the FSMM and HFMM was 56.3 mg/L and 51.6 mg/L, respectively. The effluent COD concentration and the organic matter removal efficiency were approximately identical in both MMs. The effluent NH_4^+–N concentration in both MMs was 0.9–1.1 mg/L. During this phase nitrification was stabilized. Parallel operation of FSMM and HFMM produced quite similar results in effluent quality. We confirmed that the FSMM, under long term operation, kept relatively higher flux values, thereby showing good operational efficiency. During the 6-month test period, different cleaning requirements were observed for both membrane systems. For the HFMM, membrane cleaning was necessary after 4 months of operation, while the FSMM performed well for the entire 6 months test period without needing cleaning.

During the entire third phase, relatively high NH_4^+–N concentrations (average 68 mg/L) in influent (raw municipal wastewater) were measured (see Table 4). As discussed previously, these high values were caused by grab sampling and by high morning urea concentrations in the wastewater. Despite the high NH_4^+–N concentrations in the influent, NH_4^+–N concentrations in the effluent were relatively low. After the start-up phase, during the stabilized period of operation, NH_4^+–N concentrations were measured below 1 mg/L (Fig. 4). Until the first membrane clogging (on 3 August 2006), the NH_4^+–N removal efficiency was 98%. In the period when the primary sludge started to overflow (after 18 July 2006), NH_4^+–N concentrations of about 2.5 mg/L were measured. Pilot plant nitrification occurred without problems. Effluent NO_3^-–N concentrations were relatively high because the pilot plant was not well adapted to denitrification. During operation, the NO_3^-–N concentration gradually increased, and after

Table 4
Water quality of the influent and outflow from sedimentation tank during the third phase

Concentration (mg/L)	Influent (raw municipal waste water)		Outflow (sedimentation tank)	
	Average	Scope	Average	Scope
COD	644.4	247.9–1163.9	347.8	156.5–617.3
BOD_5	239.4	110.0–450.0	88.3	35.0–180.0
SS	247.5	100.0–480.0	104.4	10.0–321.4
NH_4^+–N	67.5	28.8–147.6	58.6	28.1–102.9
N_{total}	149.1	88.8–268.2	119.6	75.5–177.3

Table 5
Water quality of the effluent treated wastewater FSMM and HFMM during the third phase

Concentration (mg/L)	Effluent (flat sheet membrane module)			Effluent (hollow fiber membrane module)		
	Average	Scope	η (%)	Average	Scope	η (%)
COD	56.3	14.9-109.9	91.3	51.6	17.8-118.6	91.9
BOD5	3.1	0.8-8.4	98.7	4.1	0.8-8.8	98.3
SS	<3	-	≈ 100.0	<3	-	≈ 100.0
NH_4^+–N	0.9	0.1-4.5	98.7	1.1	0.3-3.6	98.4
N_{total}	30.8	10.4-66.7	79.3	31.9	12.9-71.1	78.6

some time, stable concentrations at about 30 mg/L were observed (Fig. 4).

From the nitrogen balance (Table 5), we conclude that N_{total} removal efficiency was relatively high and obtained 70% efficiency. For a short period after high COD wastewater was pumped into the pilot plant, the oxygen concentration sharply decreased to below 1 mg/L (Fig. 5) and this caused reduction of NO_3^-–N in the activated sludge tank.

3.4. *The fourth phase (February 2007–July 2007)*

On 28 February 2007, the fourth test phase with immersed HFMM was started. The surface area of the membrane increased to 8 m^2 in comparison with the third test phase which operated with a surface area of 4 m^2. The module consisted of 10 organic polymer hollow fiber bunches fixed on supporting constructions. The module was situated adjacent to an aerated element. Membrane cleaning occurred with additional compressed air injection and aeration. The compressed air cleaning of the membrane was first used in the second half of the fourth phase (the first membrane clogging was on 26 March 2007). The pilot plant was inoculated by activated sludge obtained from the previously discussed third test phase. The starting sludge concentration in the activated sludge tank was nearly 3.2 g/L. Table 6 shows the water quality of the influent and effluent during the fourth phase.

At the start of the fourth phase a compressor with compressed air was not used and the HFMM was not subjected to additional (supplemental)

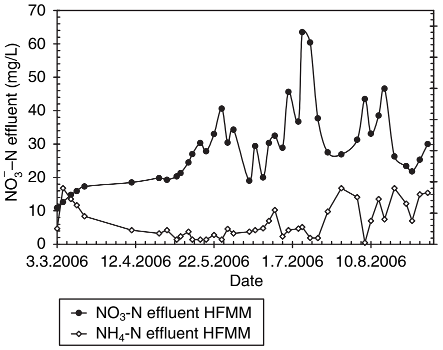

Fig. 4. NO_3^-–N and NH_4^+–N concentrations in the effluent during the third period.

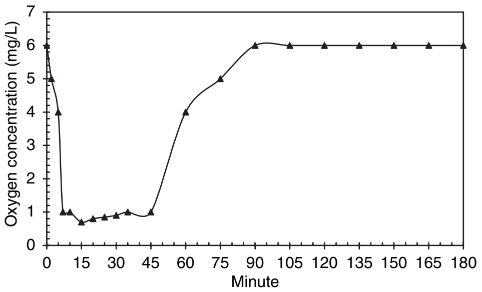

Fig. 5. Oxygen concentration during the three-hourly cycle pumping of the influent (raw municipal wastewater) during the third phase.

Table 6
Water quality of the influent and the effluent of treated wastewater HFMM during the fourth phase

Concentration (mg/L)	Influent (raw municipal wastewater)		Effluent (hollow fiber membrane module)		
	Average	Scope	Average	Scope	η (%)
COD	595.9	248.7–1037.9	46.9	14.9–78.2	92
BOD_5	291.6	255–330	3.6	2.9–8.6	99
pH	8.1	7.5–8.5	7.4	6.9–7.7	1
SS	500	160–530	<3	–	≈ 100.0
NH_4^+–N	54.1	23.2–76.6	1.5	0.4–2.6	97
N_{total}	139.5	114.3–200	27.4	3.5–45.3	80

aeration (only to aeration by the standard diffuser placed besides the HFMM). After start-up of the fourth phase, the flux was 20 L/m^2 h and gradually the flux decreased to the minimal value of 0.4 L/m^2 h at the end of March (see Fig. 6).

On 26 March 2007 the MM was completely clogged and the activated sludge overflowed the pilot plant. The membrane was subsequently cleaned and regenerated by backwashing with dilute NaOCl and by blowing. Additional injected air aeration of the membrane occurred for a duration of 10 s at 10-min intervals. The module was repeatedly started-up and the flux was held at stabilized values of about 20 L/m^2 h until the end of the experiment.

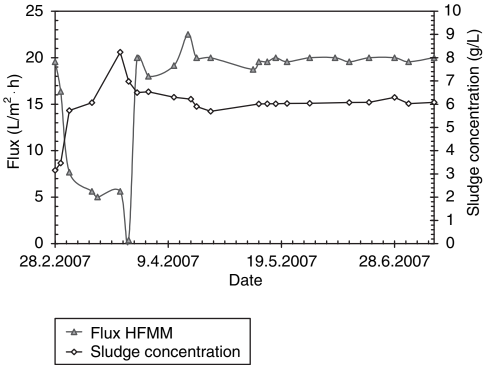

Fig. 6. Sludge concentration and flux HFMM during the fourth phase.

The decreased flux gradually increased the sludge concentration to a maximum value of 8.4 g/L (Fig. 6) at the time of total membrane clogging. After membrane regeneration, sludge concentrations were stabilized. By the end of the fourth phase (last month), sludge concentration did not change significantly. The average sludge concentration value was about 6 g/L. Despite the high age of activated sludge, its organic portion generated an average of 70% of suspended solids. The sludge index values during the entire fourth phase were about 70 mL/g.

On 12 July, after 3 months of stable measured values for the key parameters, the fourth phase of module measurement ended together with the fourth phase pilot plant test operation of the domestic WWTP.

3.5. *The fifth phase (February 2007–May 2007)*

On 23 February 2007 the fifth phase started using the HFMM with the installation of the membrane into the activated sludge tank of the real WWTP in Bratislava — DNV. The MM consisted of 60 organic polymer fiber bunches which were fixed on supporting construction. The surface area of the MM was 48 m^2. The HFMM was placed into the activated sludge tank with fine bubble aeration. Because it was anticipated that the HFMM would be insufficiently cleaned by fine bubble aeration, supplemental aeration by

Table 7
Water quality of the influent and the effluent HFMM during the fifth phase

Concentration (mg/L)	Influent (raw municipal wastewater)		Effluent (hollow fiber membrane module)		
	Average	Scope	Average	Scope	η (%)
COD	595.9	248.7–1037.9	44.9	29.3–66.7	92
BOD_5	291.6	255–330	4.2	3.1–8.9	99
pH	8.1	7.5–8.5	7.3	6.9–7.8	1
SS	500	160–530	<3	–	≈ 100.0
NH_4^+–N	54.1	23.2–76.6	1.2	0.5–1.9	98
N_{total}	135.6	92.3–210.4	29.5	11.2–62.1	78

injection of compressed air (analogous to the fourth phase) was prepared for use as needed.

In the start-up of the fifth phase, the HFMM worked without supplemental cleaning by compressed air until the first membrane clogging (16 April 2007). The flux during the first month of operation gradually decreased from 3.5 L/m^2 h to 1 L/m^2 h but stabilized during the next month. However, on 16 April 2007 total membrane clogging was observed. The membrane was then cleaned and on 25 April 2007 started operation again. To prevent membrane clogging, the HFMM was subjected to supplemental compressed air injection for a duration of 30 min at 10-min intervals. On 11 May the membrane clogged again and its operation was ended.

Because of the relatively high surface area of the HFMM, the supplemental compressed air injection during the test period was inadequate to prevent clogging. The flux values were too low and it is concluded that the MBR process needs advanced efficiency aeration to clean the MMs to prevent clogging. Table 7 shows the water quality of the influent and the effluent from HFMM installed into the activated sludge tank of WWTP DNV during the fifth phase.

4. Conclusions

In this study, 2 years of data collection and operational experience were obtained for immersed MMs in a domestic WWTP. Based on these data and experiences the following conclusions can be drawn:

- Pilot plant need activated sludge inoculation (during the first stage), because a MM without inoculation became clogged in short time.
- Experimental testing of MMs during pilot plant operations for several months without external cleaning showed that measured water quality parameters and flux were satisfactory. The FSMM was operated without external cleaning for 6 months with a flux of 20–60 L/m^2 h. The operation HFMM was operated for 4 months without external cleaning with flux of 20–45 L/m^2 h.
- Short duration, short interval, external cleaning of hollow fiber membrane by an adequate quantity of compressed air was demonstrated.
- Long term operation of the activated sludge tank without excess sludge draw off occurred due to an effective pretreatment stage (high HRT in the sedimentation tank).

Results of water quality analyses illustrated that the membrane technologies can be used to treat raw municipal wastewater to produce high quality water. Organic matter removal in this system was stable and efficient (up to 90%). Under these conditions, more than 97% of NH_4^+–N was removed, and effluent NH_4^+–N concentration was less then 1.5 mg/L during long term measurement. Nitrogen removal via denitrification was observed during the short periods with low oxygen concentration.

Based on 2 years of experiments and pilot plant operational experience, the first Slovak domestic WWTP (300 PE) using MBR was designed and constructed and is now beginning its first year of operation.

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References

[1] C.V. Smith, D.O. Gregorio, R.M. Talcott, The use of ultrafiltration membranes for activated sludge separation, Proc. 24th Ind. Waste Conf., Purdue University, Ann Arbor Science, Ann Arbor, USA, 1969, pp. 1300–1310.

[2] S. Judd, Fouling control in the submerged membrane bioreactors, Water Sci. Technol., 51 (6–7) (2005) 27–34.

[3] E.H. Bouhabila, R. Ben Aim and H. Buisson, Fouling characterisation in membrane bioreactors, Sep. Purif. Technol., 22–23 (2001) 123–132.

[4] I.S. Chang, P. Le Clech, B. Jefferson and S. Judd, Membrane fouling in membrane bioreactors for wastewater treatment, J. Environ. Eng. ASCE, 128(11) (2002) 1018–1029.

[5] I.S. Chang and C.H. Lee, Membrane filtration characteristics in membrane-coupled activated sludge system — the effect of physiological states of activated sludge on membrane fouling, Desalination, 120(3) (1998) 221–233.

[6] E.H. Bouhabila, R. Ben Aim and H. Buisson, Microfiltration of activated sludge using submerged membrane with air bubbling (application to wastewater treatment), Desalination, 118 (1998) 315–322.

[7] T. Ueda, K. Hata, Y. Kikuoka and O. Seino, Effects of aeration on suction pressure in a submerged membrane bioreactor, Water Res., 31 (1997) 489–494.

[8] C. Cabassud, S. Laborie, L. Durand-Bourlier and J.M. Laine, Air sparging in ultrafiltration hollow fibers: relationship between flux enhancement, J. Membr. Sci., 181 (2001) 57–69.

[9] M. Mercier-Bonin, I. Daubert, D. Leonard, C. Maranges, C. Fonade and C. Lafforgue, How unsteady filtration conditions can improve the process efficiency during cell cultures in membrane bioreactors, Sep. Purif. Technol., 22–23 (2001) 601–615.

[10] I.S. Chang and S. Judd, Air sparging of a submerged MBR for municipal wastewater treatment, Process Biochem., 37(8) (2002) 915–920.

[11] M. Dvořáková, M. Chorvátová, M. Pečenka and I. Růžičková, Separation of activated sludge by means of the membrane technology, Vodní hospodářství, 57(7) (2007) 239–244 (in Czech).

[12] M. Horáková, P. Lischke, A. Grunwald, Chemical and Physical Methods of Water Analysis, Alfa, Bratislava, 1986 (in Czech).

[13] A. Jirka and M.J. Carter, Micro semi-automated analysis of surface and waste waters for chemical oxygen demand, Analyt. Chem., 47 (1975) 1397.

[14] Aprica Studi, S. Rondi, Two years of operation of a 42 000 m^3/d MBR process: the case of ASM Brescia (Italy), 6. Aachener Tagung Siedlungswasserwirtschaft und Verfahrenstechnik, A2-1 2005.