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The forward osmosis membrane bioreactor:
A low fouling alternative to MBR processes

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Abstract

A novel osmotic membrane bioreactor (OsMBR) is presented. The system utilizes a submerged forward osmosis (FO) membrane module inside a bioreactor. Through osmosis, water is transported from the mixed liquor across a semi-permeable membrane, and into a draw solution (DS) with a higher osmotic pressure. To produce potable water, the diluted DS is treated in a reverse osmosis (RO) unit; the by-product is a reconcentrated DS for reuse in the FO process. Preliminary results from experiments conducted with a flat-sheet cellulose triacetate FO membrane demonstrated high sustainable flux and relatively low reverse transport of solutes from the DS into the mixed liquor. Membrane fouling was controlled with osmotic backwashing. The FO membrane was found to reject 98% of organic carbon and 90% of ammonium-nitrogen; the OsMBR process (bioreactor and FO membrane) was found to remove greater than 99% of organic carbon and 98% of ammonium-nitrogen, respectively; suggesting a better compatibility of the OsMBR with downstream RO systems than conventional membrane bioreactors.

Keywords: Membrane bioreactor; Forward osmosis; Membrane fouling; Wastewater treatment; Nutrient removal

1. Introduction

More stringent regulations and the potential to produce high quality effluent make membrane bioreactors (MBRs) an attractive process for domestic wastewater treatment [1]. In a submerged MBR, microporous [microfiltration (MF) or ultrafiltration (UF)] membranes are immersed in a bioreactor and water is filtered through the membranes using vacuum; suspended solids are retained in the system and high levels of treatment (including nutrient removal) can be achieved [2]. The MBR replaces the two stages of the conventional activated sludge process (biotreatment and clarification) with a single, integrated process. MBR effluent may be suitable for

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use as irrigation water, process water, or a source of potable water. For potable reuse (e.g., indirect reuse through aquifer recharge), advanced treatment [e.g., reverse osmosis (RO), nanofiltration (NF), or advanced oxidation] is necessary after the MBR [3]. The advantages of MBRs over conventional treatment have been thoroughly reviewed [1] and they include product water consistency, reduced footprint, reduced sludge production due to a high biomass concentration in the bioreactor, and essentially complete suspended solids removal from the effluent. The main problem associated with MBRs is membrane fouling. Fouling reduces permeate flux and increases the frequency of membrane cleaning and replacement [4]. Membrane fouling can occur in the MBR itself, and also in the downstream RO system [5]. Specifically, high concentrations of dissolved organic compounds in the MBR effluent can cause severe fouling of RO membranes; this leads to reduction of water flux and deterioration of treated water quality [6].

In order to operate conventional MBRs at constant flux, physical membrane cleaning techniques are utilized; they include backwashing, relaxation, or a combination of the two, depending on the membrane configuration (flat-sheet or tubular). Chemicals are often added to enhance physical cleaning [7]. During backwashing, the permeate is pumped in the opposite direction through the membrane. Backwashing effectively removes most of the reversible fouling due to pore blocking. The efficiency of backwashing has been studied in detail and the key parameters have been found to be frequency, duration, and intensity [8–10]. During membrane relaxation, permeate suction is stopped and the back-transport of foulants is naturally enhanced as reversibly attached foulants diffuse away from the membrane surface due to the concentration gradient. Tubular and hollow fibers membranes undergo regular backwashing and sometimes relaxation [8–12]. Flat-sheet membranes cannot be backwashed due to their inability to withstand

pressure in the opposite direction of the operating flow; for this reason, relaxation is used to control the fouling of these membranes [4].

A novel MBR system that utilizes a submerged forward osmosis (FO) membrane in the bioreactor is investigated in the current study. FO, or simply osmosis, is the transport of water across a selectively permeable membrane from a solution of higher water chemical potential (low osmotic pressure) to a solution of lower water chemical potential (higher osmotic pressure). Typically, the FO process results in concentration of the feed stream and dilution of a highly concentrated stream [referred to as the draw solution (DS)] [13]. In an osmotic MBR (OsMBR) system (Fig. 1), wastewater is fed into a reactor which is continuously aerated to supply oxygen for the biomass and to scour the membrane. Through osmosis, water diffuses from the bioreactor, across a semi-permeable membrane, and into a lower water chemical potential DS. The FO membrane acts as a barrier to solute transport and provides high rejection of the contaminants in the wastewater stream [14–16]. The diluted DS is sent to a reconcentration process (e.g., RO or distillation) which reconcentrates the DS and generates a high-quality product water. Thus, in most wastewater treatment applications, FO is not the ultimate process but rather a high-level pretreatment step before an ultimate reconcentration process. Compared to the MF or UF process in a conventional MBR, the FO process in the OsMBR offers the advantages of much higher rejection (semi-permeable membrane versus microporous membrane) at a lower hydraulic pressure. FO processes are also likely to have lower fouling propensity than pressure-driven systems [13], and therefore, require less frequent backwashing. When comparing an OsMBR system (OsMBR followed by RO) with a conventional MBR followed by RO, the high rejection of the FO membrane will result in an RO influent with lower fouling propensity and may lead to a higher quality RO product
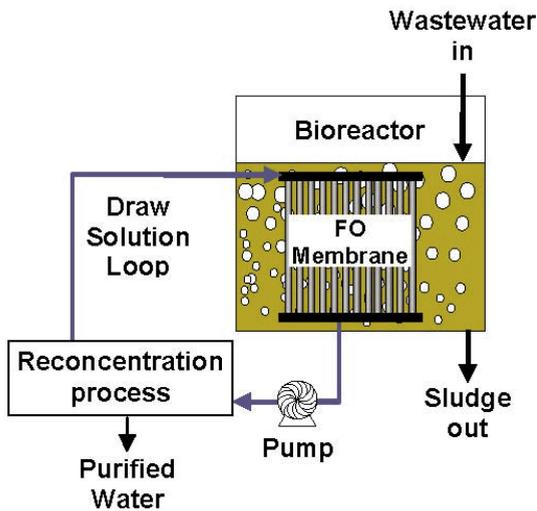
Fig. 1. Schematic diagram of an OsMBR system.

water. Furthermore, FO followed by the RO treatment scheme represents a dual barrier purification system.

In FO membrane processes, osmotic backwashing can be performed to remove foulants from the surface of the membrane. During osmotic backwashing, water flows from the support side of the membrane to the active side, thereby reversing the direction of flow through the FO membrane and potentially removing foulants [17,18].

A key consideration in developing an OsMBR system is selection of an appropriate DS [13]. The main criterion is that the DS has a higher osmotic pressure than the feed solution. Another important criterion in some FO applications is the availability of a suitable process for reconcentrating the draw solution after it has been diluted in the FO process. Very often an NaCl solution is selected because it has high solubility and is relatively easy to reconcentrate to high concentrations using a conventional desalination process (e.g., RO or distillation) without risk of scaling [13].

Transport of the draw solute (e.g., Na^+ or Cl^- ions) into the bioreactor through the membrane

must also be considered in the OsMBR process. This "reverse salt transport" is expected to occur because of the difference in solute concentration between the DS and the reactor solution. Reverse salt transport from the DS not only results in a reduced driving force, but may also have inhibitory or toxic effects on the microbial community inside the reactor.

In this paper, the feasibility of the OsMBR system to treat domestic wastewater is evaluated. The specific objectives of this paper are to present the novel OsMBR system and to report the results of preliminary experiments focused on water flux, reverse salt transport, fouling propensity, and nutrient removal. Furthermore, the operation of the OsMBR process is compared to conventional MBR processes in terms of filtration time, backwashing time, net flux, and TOC removal efficiencies.

2. Materials and methods

2.1. FO membranes

Three flat-sheet cellulose triacetate (CTA) FO membranes (Hydration Technologies, Albany, OR), designated membranes A, B, and C, were used in the experiments. These membranes are unique compared to other semi-permeable membranes (e.g., RO membranes), and have been determined to be the best available membranes for current FO applications [13,19]. These membranes have different selectivities and permeabilities; however, specific differences are proprietary.

2.2. Batch abiotic experiments

An abiotic batch apparatus was used to characterize the water flux and reverse salt transport of the three membranes (Fig. 2). In this apparatus, a unique plate-and-frame FO membrane module was immersed in a 14 L reactor containing doubly deionized water (DDW). In the
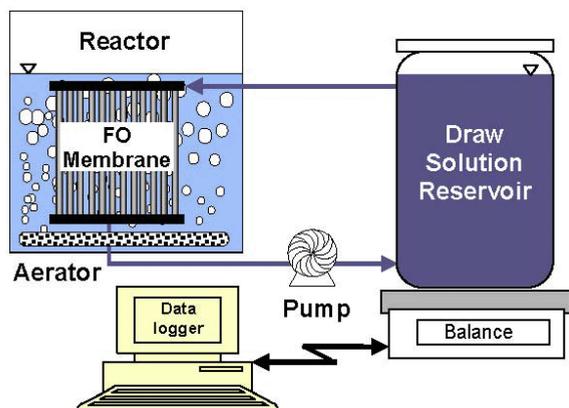
Fig. 2. Schematic diagram of the batch laboratory-scale OsMBR apparatus.

FO membrane module, two flat-sheet membranes were held in place by frames bolted to an inner plate. The DS flowed between the membranes through inlet and outlet channels located on opposite sides of the plate. The membranes were oriented with their active sides facing the reactor and their support sides facing the DS. The total membrane surface area was 0.0173 m^2. Mesh spacers were placed between the membrane and the central unit to both support the membrane and increase the turbulence of the DS as it passed through the membrane module. An aerator was placed at the bottom of the reactor to continuously agitate the reactor contents.

The DS reservoir was placed on an analytical balance (PB5001-S, Mettler-Toledo, Columbus, OH) linked to a computer; a variable-speed gear pump (Cole-Palmer, Vernon Hills, IL) was used to recirculate the DS at 1.5 L/min. The DS concentration was allowed to decrease from 70 to 30 g NaCl/L as water diffused through the membrane from the reactor to the DS reservoir. Flux through the membrane was calculated based on the change of weight of the DS. The conductivity of the reactor solution was continuously monitored (Accumet Basic, Fisher Scientific, Hampton, NH) and recorded to calculate the reverse salt transport.

The three CTA membranes were tested under the same conditions and using NaCl as the DS. The temperature was held constant at 23±1°C.

2.3. Continuous-flow OsMBR experiments

A continuous-flow apparatus was used to evaluate the OsMBR system under continuous feed and constant DS conditions (Fig. 3). Water flux, reverse salt transport, and removal efficiencies of the biological, FO and RO processes were investigated.

The reactor was continuously fed from a stock solution feed reservoir. The synthetic feed solution consisted of 5 g/L meat extract, 1 g/L $C_6H_{12}O_6$, 0.6 g/L $(NH_4)_2SO_4$, and 0.14 g/L K_2HPO_4. Between 1 and 2 g/L $NaHCO_3$ was also added to maintain neutral pH. This solution had a chemical oxygen demand (COD) of 4.5± 0.2 g/L, a C:N:P ratio of 100:5:1, and a food-to-microorganism (F/M) ratio of 0.25 kg COD/(kg MLSS·d). Due to the large volume of the bioreactor (14 L) and the associated large hydraulic retention time (HRT) of the system (3.5 d), the feed concentrations were higher than those found in typical domestic wastewaters. These concentrations were necessary to keep the F/M ratio within the range of 0.10 to 0.40 kg COD/(kg MLSS·d), a range that is typical for MBR treatment processes [1]. Similar to the batch apparatus, the plate-and-frame FO membrane module was immersed in a 14 L reactor. However, in the continuous-flow apparatus, the reactor was filled with 5.5 g mixed liquor suspended solids per liter (g MLSS/L) that were collected from the aeration basin of a conventional wastewater treatment facility (Truckee Meadows Water Reclamation Facility, Reno, Nevada). The mixed liquor was acclimated to the new feed for a period of 2 weeks prior to being added to the OsMBR system. The mixed liquor level in the reactor was held constant by a float valve. The aerator placed at the bottom of the reactor agitated the solution,
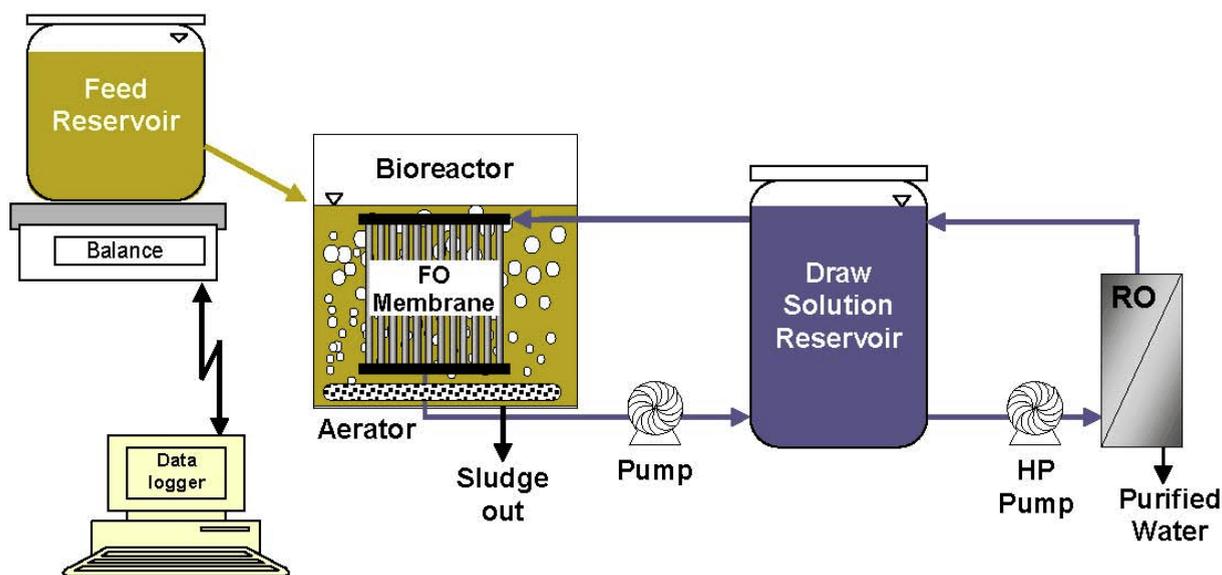
Fig. 3. Schematic diagram of the continuous-flow laboratory-scale OsMBR apparatus.

provided oxygen to the microbial community, and lightly scoured the membrane surface.

The feed reservoir was placed on an analytical balance linked to a computer. Flux through the membrane was calculated based on the change of weight of the feed solution transferred to the reactor. A variable-speed gear pump was used to recirculate the DS at 1.5 L/min. The conductivity of the mixed liquor was monitored and recorded in order to calculate the reverse salt transport and the total accumulation of salt in the bioreactor.

The DS concentration was held constant at 50±1 g NaCl/L by continuous reconcentration with an RO system. The RO system was comprised of a SEPA-CF membrane cell (GE Osmonics, Minnetonka, MN), utilizing an SW30HR membrane (Dow Chemical Company, Midland, MI), and driven by a high pressure positive displacement pump (Hydracell M03, Wanner Eng., Minneapolis, MN).

The continuous-flow experiments were conducted utilizing membrane B as the FO membrane. The solids retention time (SRT) of the

OsMBR was 15 days, thus guaranteeing the development of nitrifying bacteria [20]; SRT was maintained by daily wasting of excess sludge. The temperature was held constant at 23±1°C.

2.4. Fouling verification and cleaning experiments

After 14, 21, and 28 days of continuous operation, the membrane module was removed from the continuous-flow apparatus and placed in the batch apparatus. An abiotic batch experiment was performed for approximately 8 h in order to quantify the water flux loss due to fouling only. Reverse salt transport was also measured. Water flux loss in the continuous system is expected to be caused by a combination of fouling and reduced driving force due to the salt accumulated in the bioreactor through reverse salt transport. Because DDW is used as the reactor solution in the batch abiotic experiments, and the experiment duration was only 8 h, the reverse salt transport over the duration of the experiments did not result in substantial salt accumulation in the bioreactor

approximately 100 mg NaCl/L at the end of the experiment). Considering this, and also considering that fouling does not occur in the abiotic experiments, flux decline due to only membrane fouling can be isolated. The batch cleaning experiments were designed to mirror the abiotic batch experiments that were used to initially characterize the flux and rejection of the membranes; the experiments lasted as long as it took for the DS concentration to decrease from 70 to 30 g NaCl/L (approximately 8 h). After 8 h, an osmotic backwash was performed in batch mode for 1 h. The osmotic backwash was accomplished by filling the reactor with a 5 g NaCl/L solution (a concentration similar to that found inside the continuous-flow reactor) filling the DS reservoir with DDW, and recirculating the DDW through the membrane module at a flow rate of 1.5 L/min with a variable-speed gear pump. Following the osmotic backwash and prior to returning the membrane module to the continuous-flow system, the membrane was operated in abiotic batch mode for an additional 8 h (as the DS again decreased from 70 to 30 g NaCl/L). This step isolated the effect of osmotic backwashing on water flux and reverse salt transport.

2.5. Analytical methods

Daily grab samples were collected from the feed reservoir, the reactor, the DS reservoir, and the product water. The samples were first centrifuged (Centrifuge 5415C, Eppendorf AG, Hamburg, Germany) at 9,000 rpm for 15 min to separate the biomass from the solution. The supernatant was then filtered through a 0.22-µm filter to remove the fine particles and biomass residuals. Samples from the DS and product water were tested without centrifugation. The samples were analyzed for ammonium-nitrogen (NH_4^+-N) according to Standard Methods (APHA, 1999) and for total organic carbon (TOC) using a TOC analyzer (TOC-Vcsh, Shimadzu Corp., Kyoto, Japan).

3. Results and discussion

3.1. Batch experiments

Water flux as a function of DS concentration is illustrated in Fig. 4 for the three membranes tested in this study. For all three membranes, water flux decreases as DS concentration decreases because of the decreasing osmotic pressure difference between the DS and the reactor solution. Water flux through membrane A is at least 29% greater than the water flux through membrane B, and water flux through membrane B is at least 71% greater than the water flux through membrane C. Thus, it appears that membrane A is the loosest, membrane B is in the middle, and membrane C is the tightest membrane.

The rate of reverse salt transport as a function of DS concentration is depicted in Fig. 5. For all three membranes, reverse salt transport decreases linearly as the DS concentration decreases. The magnitude of reverse salt transport for membrane A is at least 2.2 times greater than that for membrane B and the reverse salt transport for membrane B is at least 7.3 times greater than that for membrane C. Thus, the membrane with the highest water flux (membrane A) also showed the highest reverse salt transport and the membrane with the lowest water flux (membrane C) showed the lowest reverse salt transport. These results are expected as the loosest membrane matrix would be expected to have the advantage of high water flux and the disadvantage of high reverse salt transport. Membrane B was selected for the continuous flow experiments because it exhibited a favorable water flux while maintaining a reverse salt transport rate that would not lead to a salt concentration inside the bioreactor high enough to adversely affect the biological process.

3.2. Continuous-flow OsMBR experiments

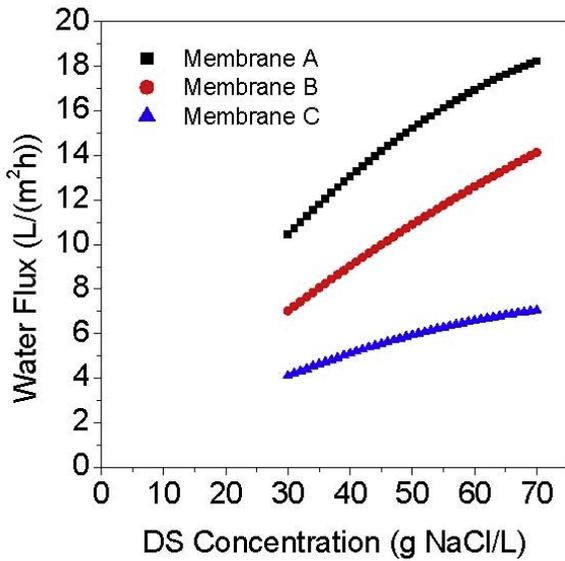TOC and NH_4^+-N concentrations at various locations in the OsMBR system are summarized

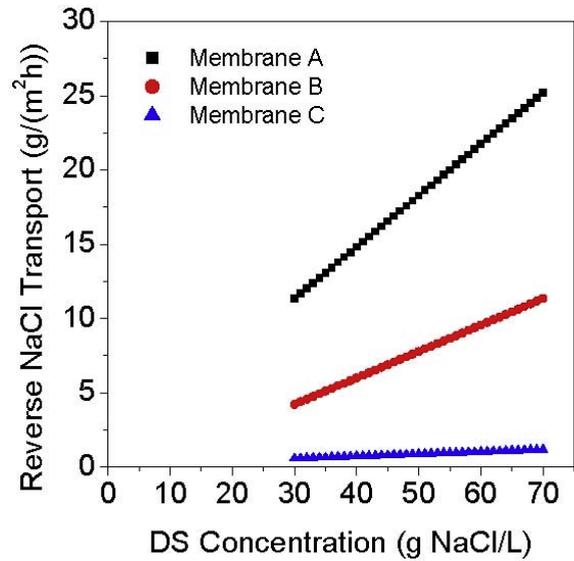Fig. 4. Water flux as a function of DS concentration for batch FO experiments with three CTA membranes.

Fig. 5. Reverse salt transport as a function of DS concentration for batch FO experiments with three CTA membranes.

Table 1
TOC and NH_4^+-N concentrations at various locations in the OsMBR system

	Feed solution	Bioreactor	Draw solution	Product water
TOC (mg/L)	1,325 ± 25	140 ± 5	3 ± 0.5	2.5 ± 0.5
NH_4^+-N (mg/L)	65 ± 5	15 ± 2	1.5 ± 0.5	0.4 ± 0.1

in Table 1. The difference between the feed solution concentrations and the bioreactor concentrations represents the effect of both biological degradation and the FO process. Biological degradation reduces the TOC and NH_4^+-N concentrations in the bioreactor, while the FO process increases these constituents, as the semi-permeable membrane rejects and retains TOC and NH_4^+-N inside the bioreactor. The data in Table 1 reveal that biological degradation works to greatly reduce the concentrations of TOC and NH_4^+-N, substantially overcoming the concentration process of the FO membrane. In the DS and in the product water, TOC and NH_4^+-N concentrations further decrease due to the FO and RO membrane rejection, respectively.

Removal efficiencies of the FO membrane, the OsMBR process (bioreactor and FO membrane), and the overall system (OsMBR followed by RO) are summarized in Table 2. For removal only by the FO membrane, both the TOC and NH_4^+-N rejections are comparable to published results for semi-permeable membranes [17,21]. For the OsMBR process (bioreactor and FO membrane), the TOC and NH_4^+-N removals are substantially higher than those obtained in conventional MBRs, where removals up to 95% are typical [2]. For the OsMBR system (OsMBR followed by RO), greater than 99% removal is achieved for both TOC and NH_4^+-N.

Water flux and salt concentration in the OsMBR process are illustrated in Fig. 6. The

Table 2
TOC and NH_4^+-N removal efficiencies at various stages of the OsMBR system

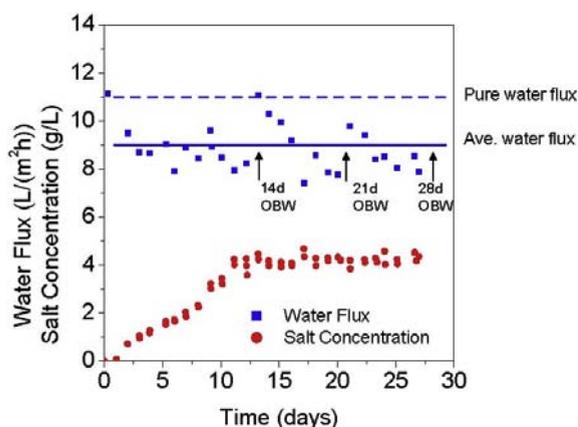|	% rejection of FO membrane	% removal of OsMBR process (bioreactor + FO)	% removal of overall system (OsMBR + RO)
TOC (mg/L)	97.9	99.8	99.8
NH_4^+-N (mg/L)	90.0	97.7	99.4

Fig. 6. Water flux through the FO membrane and salt concentration in the bioreactor as a function of time for continuous-flow experiments with membrane B. Osmotic backwashings (OBWs) are indicated at 14, 21, and 28 d.

average water flux during the experiment was approximately 9 L/(m^2·h). This flux value was approximately 18% lower than the pure water flux observed in the batch experiments (11 L/(m^2·h) at 50 g NaCl/L). These results confirm the lower fouling propensity of the FO process compared to pressure-driven membrane processes, where severe fouling causes the operating water flux to be a small percent of the initial water flux (75% lower than the initial flux in one investigation [11]). The likely cause of flux decline is increased hydraulic resistance caused by foulants deposited on the membrane surface. In order to determine how much of the 18% flux decline is due to membrane fouling and how much is due to increasing salinity in the reactor, the NaCl concentration in the OsMBR process must be

considered. As can be seen in Fig. 6, the salt concentration in the bioreactor steadily increases due to reverse salt transport over the first 14 d. This reduces the net difference between the DS concentration and the reactor salt concentration and results in a lower driving force. Eventually it is expected that the reactor salt concentration will reach a constant value depending on the operating SRT. The operating SRT determines the amount of daily sludge wasting and hence, the removal of salt from the bioreactor. NaCl that enters the bioreactor due to reverse salt transport will only be removed from the bioreactor by sludge wasting; thus the SRT regulates its concentration inside the bioreactor. From Fig. 6 it appears that the salt concentration in the bioreactor stabilized at approximately 4 g/L around 14 d of operation. Thus, reverse salt transport plays a substantial role in flux decline over the first 14 d. After that time, when the salt concentration in the reactor remains relatively constant, observed flux decline is likely due to membrane fouling only. It is also worth noting that the level of salinity observed in the reactor is not a concern in terms of inhibition or toxic effects on the biological processes [22].

The arrows at 14, 21, and 28 d in Fig. 6 indicate times when osmotic backwashing was performed. Figs. 7 and 8 summarize the results of the offline batch experiments performed to study membrane fouling and backwashing at 14, 21, and 28 d. In Fig. 7, the left and right bars of each pair depict the effects of membrane fouling and osmotic backwashing on water flux, respectively. After 14 d of operation, the water flux decreased
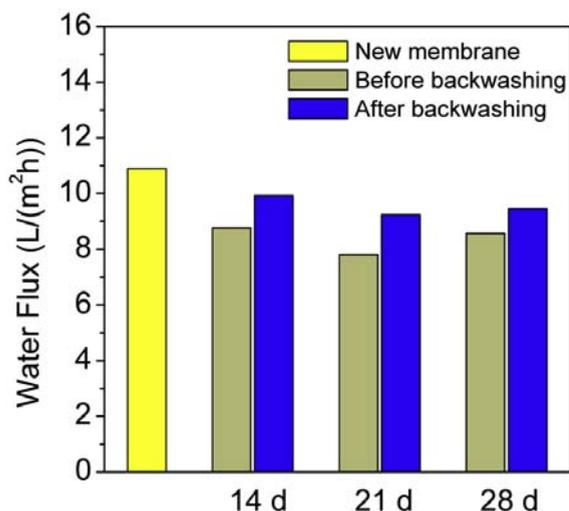
Fig. 7. Water fluxes of membrane B. Comparison between water flux for the new membrane and before/after osmotic backwashings at 14, 21, and 28 d. The concentration of the draw solution was 50 g NaCl/L.
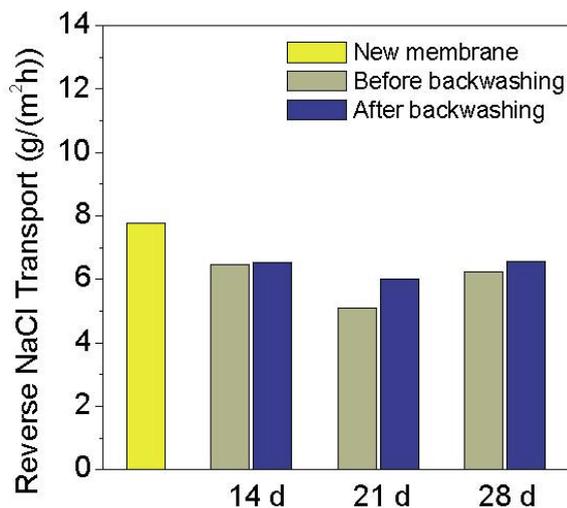

Fig. 8. Reverse salt transport through membrane B. Comparison between reverse salt transport for the new membrane, and before/after osmotic backwashings at 14, 21, and 28 d. The concentration of the draw solution was 50 g NaCl/L.

by approximately 20% from the initial flux of the new membrane. Osmotic backwashing was able to restore approximately 50% of this flux loss, leaving an overall 10% lower flux due to irreversible fouling. Similar trends were observed after 21 and 28 d. This suggests that after an initial phase of irreversible fouling (occurring in the first 14 days), later phases of fouling are more reversible and the water flux can be maintained at a constant level over time. This reinforced the earlier observation (in Fig. 6) of the low fouling effects of the OsMBR process. Additionally, these preli-minary results indicate that osmotic backwashing is an effective way to control membrane fouling. Comparing Fig. 7 with Fig. 6, it is worth noting that the values of water fluxes taken from the con-tinuous experiments at the end of each cycle (the points just before each arrow in Fig. 6) match with the water fluxes of the abiotic batch experi-ments prior to osmotic backwashing (the left bars in Fig. 7), showing that no unintentional cleaning occurred during the abiotic batch experiments, and that the cleaning was in effect due to the osmotic backwashing.

Furthermore, integrating the results of the offline batch experiments, the resulting average water flux is 9.4 L/(m^2·h), approximately 15% lower than the pure water flux observed in the batch experiments (11 L/(m^2·h) at 50 g NaCl/L). This is the hypothetical average water flux influenced by membrane fouling only. Comparing this to the flux loss due to membrane fouling and salt concentration in the bioreactor (18% from Fig. 6), the flux loss is almost completely due to membrane fouling; only minimal flux decline is attributed to the reduced driving force resulting from salt accumulation in the bioreactor.

The effects of fouling and osmotic backwashing on reverse salt transport are illustrated in Fig. 8. The reverse salt transport of the used membrane is approximately 16% less than that of the new membrane, and it remains fairly constant [between 5 and 6.5 g/(m^2·h)] over the duration of the experiment. Generally, there is no obvious effect of membrane backwashing on the reverse salt transport in the process; and as would be expected, the limited fouling that occurs at the

Table 3
Operation comparisonbetween conventional MBRs and the OsMBR process

Membrane	Material	Pore size, µm	Flux, L/(m²·h)	Filtration–cleaning cycle		Cycles/day	Net flux[a], L/(m²·h)	Ref.
				Filtration (min)	Backwashing (min)			
Hollow fiber	Polyethylene–polysulfone	0.1	20	5–45	0.25–15[b]	24–274	15–18	[8,11, 12]
Tubular	Polypropylene	0.2	8–22	30	0.25	48	—[c]	[9,10]
Flat sheet	C-PVC–Stainless steel	0.2–0.4	17–22	3–8	1–4[b]	120	8.7–11	[23, 24]
OsMBR	CTA	—[d]	9.0	10,020	60	0.14	8.9	This study

[a]Net flux is either reported by the authors of the articles or calculated based on the information contained in the articles. Calculations are based on relaxation time, backwashing time, and backwashing water flux. In the case of relaxation only, the net flux value is calculated by multiplying the flux by the ratio between the filtration time and the total cycle time. In the case of backwashing, the net flux value is calculated by subtracting the backwashing flux times the ratio between the backwashing time and the total cycle time from the filtration flux where the filtration flux is calculated the same as the relaxation flux.
[b]Relaxation time.
[c]Data not available. The authors of the articles did not report the net flux, nor the flux during backwashing.
[d]The CTA is a semi-permeable membrane.

membrane surface has a beneficial effect of reducing the reverse salt transport from approximately 7.7 g/(m²·h) for the new membrane to approximately 6.4 g/(m²·h) for the used membrane.

The operation of the novel OsMBR process is compared to conventional MBR processes in Table 3. In microporous MBR systems, the length of the filtration-cleaning cycle varies between 4 and 90 min and the cycles repeat between 24 and 274 times per day depending on the membrane configuration and operating conditions. During the experiments reported here, the OsMBR process required substantially fewer backwashing cycles (once per week) than the conventional MBR process. In the osmotically-driven MBR, the lack of hydraulic pressure across the membrane means that chemical or particulate foulants are not forced onto the membrane surface. Because cleaning duration and intervals markedly

affect energy requirements, operational costs, system productivity, and membrane integrity [7], the need for less frequent backwashing is a substantial benefit of the OsMBR process. The net flux of conventional MBRs is greatly influenced by the downtime caused by cleaning procedures, with a reduction up to 50% from the instantaneous flux (e.g., a net flux of 11 L/(m²·h) compared to an instantaneous flux of 22 L/(m²·h) in Table 3). The fewer backwashing cycles required by the OsMBR process leads to a net flux essentially equal to its instantaneous flux, and in some cases, greater than the net flux of conventional MBRs.

A further advantage of the OsMBR process (bioreactor and FO) compared to a conventional MBR is the higher TOC removal efficiencies that can be achieved by an FO membrane compared to a microporous membrane (Table 4). Soluble organic matter is responsible for most of the

Table 4
TOC removal efficiencies of conventional MBR membranes and the FO membrane in the OsMBR system

Membrane	Material	Pore size, µm	Removal, %	Ref.
Hollow fiber	Polyethylene	0.1	75	[12]
Hollow fiber	Polysulfone	0.1	68[a]	[25]
Flat sheet	Stainless steel	0.2	58[a]	[24]
Flat sheet	C-PVC	0.4	28–79[a]	[26]
Flat sheet	Polyethylene	0.4	87[a]	[27]
OsMBR	CTA	—[b]	98	This study

[a]Based on COD.
[b]The CTA is a semi-permeable membrane.

fouling of RO membranes downstream of conventional MBRs [5,6]. Microporous membranes do not reject soluble matter due to their porous nature, but due to the formation of a cake layer on their surface, the membrane plus cake layer system can typically retain 28–87% of soluble organic matter. The semi-permeable FO membrane has been shown to reject 98% of TOC due to its non-porous composition.

4. Conclusions

A novel submerged OsMBR system was presented. Long-term water fluxes for experiments with activated sludge operated at a solids concentration of 5.5 g MLSS/L were only 18% lower than water fluxes for experiments using DDW feed. Most of the reduction in water flux was due to membrane fouling; only a small part was due to increased salt concentration inside the bioreactor due to reverse salt transport.

Osmotic backwashing was conducted once per week to restore water flux to approximately 90% of the initial water flux. Compared to a conventional MBR, this system required substantially less backwashing. OsMBR process removal efficiencies for TOC and NH_4^+-N were greater than 99% and 98%, respectively, suggesting a better compatibility of the OsMBR with downstream RO systems than conventional MBRs.

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