Forward osmosis for concentration of anaerobic digester centrate

Ryan W. Holloway[a], Amy E. Childress[b], Keith E. Dennett[b], Tzahi Y. Cath[c],*

[a]Tahoe-Truckee Sanitation Agency, 13720 Joerger Drive, Truckee, CA 96161, USA
[b]Department of Civil and Environmental Engineering, University of Nevada, Reno, MS 258, Reno, NV 89557-0152, USA
[c]Colorado School of Mines, Division of Environmental Science and Engineering, 1500 Illinois St., Golden, CO 80401, USA

ABSTRACT

The nutrient-rich liquid stream produced during the dewatering of digested biomass (i.e., the centrate) is commonly mixed with the influent raw wastewater at wastewater treatment facilities. This increases the nitrogen and phosphorus loading on biological processes, increases operating costs, and in some cases, results in increased nutrient concentrations in the final effluent. Forward osmosis (FO) is a membrane treatment process that was investigated at bench scale to determine its feasibility to concentrate centrate under both batch and continuous operating conditions. The continuous bench-scale system used FO as pretreatment for reverse osmosis (RO). Results demonstrated that high water flux and high nutrient rejection could be achieved. The combined FO/RO process exhibited sustainable flux over an extended time period. A mathematical model was developed in order to determine the specific energy, power, and membrane area requirements for a larger-scale centrate treatment process. Modeling results indicated that to optimize power and membrane area requirements, the system should be operated at approximately 70% water recovery.

1. Introduction

Excess biosolids generated during wastewater treatment are most often treated in an aerobic or anaerobic digester in order to destroy pathogens, reduce biochemical oxygen demand (BOD), and reduce the volume of the solid waste to be handled. The end products of digestion are biogas and stabilized biomass. When the digested biomass is dewatered using a centrifuge, a biosolids fraction and a liquid fraction (i.e., a centrate) are produced. The centrate is nutrient-rich and has ammonia, total Kjeldahl nitrogen (TKN), and orthophosphate concentrations typically exceeding 1100, 1300, and 200 mg/l, respectively. BOD concentrations up to 2000 mg/l (Metcalf and Eddy, 2003), dissolved solids concen-

trations of 500–1000 mg/l, and suspended solids concentrations of 50–100 mg/l are also typical. Many wastewater treatment plants combine the centrate with the influent raw wastewater and thus the centrate can contribute 15–20% of the influent nitrogen load (Fux et al., 2002) and approximately 8% of the influent phosphorus load (Wild et al., 1997). Treatment or removal of this liquid stream from the treatment plant could greatly reduce operating costs and improve the water quality of the final effluent from the treatment facility. Furthermore, if successfully concentrated, the centrate could be economically transported and beneficially used as a fertilizer.

To mitigate the problems associated with centrate, a process is needed that can either remove both nitrogen and

phosphorus or reduce the volume of centrate. A physiochemical treatment process for centrate was developed by Ueno and Fujii (2001), which uses a fluidized bed reactor to precipitate struvite (a 1:1:1 molar ratio of ammonia, phosphate, and magnesium) and thereby reduce concentrations of phosphorus and ammonia in the centrate. The main limitation of this process is that only a small fraction of the ammonia is converted to struvite. Mulder et al. (2001) investigated a microbial treatment process, referred to as the SHARON (Single reactor High activity Ammonia Removal Over Nitrite) process. In this process, ammonium is converted to nitrite under aerobic conditions and then nitrite is reduced to nitrogen gas by denitrifying bacteria. Mulder et al. were able to remove up to 90% of the nitrogen from centrate using this process. However, because the SHARON process is designed only to remove nitrogen, phosphorus remains in the centrate.

The focus of the current investigation is on forward osmosis (FO) pretreatment for reverse osmosis (RO) concentration of centrate. FO is the engineered application of osmosis whereby a feed solution is concentrated and its volume is minimized when water diffuses from the feed solution, through a membrane, and into a solution of higher solute concentration. The higher concentration solution is referred to as the draw solution (DS). As the water diffuses through the membrane, the concentration and osmotic pressure of the DS decline. RO can then be used to continuously reconstitute the DS (Cath et al., 2006). Thus, FO can be considered as a pretreatment for RO (Fig. 1). Because FO is an osmotically driven process that requires very low hydraulic pressure, the effects of fouling are reduced and membrane support and compaction are of less concern.

The driving force in FO is the gradient ($\Delta\pi$) between the osmotic pressure of the DS (π_{DS}) and the osmotic pressure of the feed solution (π_F):

$$\Delta\pi = \pi_{DS} - \pi_F. \tag{1}$$

Thus, the concentration and composition of the DS have strong effects on the FO process. Water flux (J_w) through the membrane in FO is the product of the membrane permeability coefficient (A_w) and the osmotic pressure driving force:

$$J_w = A_w\Delta\pi. \tag{2}$$

FO/RO has been tested for direct potable reuse of wastewater (Cath et al., 2005a, b; Cartinella et al., 2006). FO/RO has also been proven to be capable of treating low-quality feed solutions such as landfill leachate and industrial wastes (Anderson, 1977; Beaudry and Herron, 1997; York et al., 1999; Cath et al., 2005b). York et al. used a combined FO/RO system for the treatment of landfill leachate, which is similar to centrate in that it is a challenging stream to treat because it contains a wide range of pollutants including organic compounds, heavy metals, organic and inorganic nitrogen, suspended solids, and TDS. York et al. were able to achieve water recoveries up to 95%. Also, the quality of the final product water met stringent permit limits, which were based on state and federal ambient chronic freshwater quality standards.

The overall goal of the current investigation was to evaluate the feasibility of using FO/RO for the treatment of anaerobic digester centrate. More specifically, the advantages, limitations, and economics of using FO as pretreatment for RO were investigated. The objectives leading to this goal were (1) to evaluate FO process performance under a wide range of operating conditions, (2) to compare the performance of FO with that of RO for the treatment of centrate, and (3) to develop a mathematical model to determine the economic viability of FO pretreatment for centrate treatment.

2. Materials and methods

2.1. Evaluation of FO at bench scale

Bench-scale FO batch experiments were conducted to determine if FO is a viable process for centrate treatment. FO was evaluated using different centrate feed solutions and DS concentrations. The performance of FO was evaluated based on sustainable flux, recoverable flux (flux after cleaning), and nutrient rejection.

The membrane used in all FO experiments was a cellulose triacetate (CTA) membrane (Hydration Technologies, Inc., Albany, OR). The physical characteristics of this specific CTA membrane are unique compared with other commercially available semi-permeable membranes and it has been determined to be the best available membrane for most FO applications (Cath et al., 2005b; McCutcheon et al., 2006). The CTA membrane is relatively hydrophilic and negatively charged at the pH of feed solutions used in the FO experiments (pH 7.2) (Cartinella et al., 2006).

Two feed solutions were used in the investigation: raw centrate and filtered centrate. Additionally, deionized (DI) water was used as the feed solution in several batch FO experiments. Raw centrate was collected directly from the dewatering centrifuges at the Truckee Meadows Water Reclamation Facility (TMWRF) (Reno, NV) and was processed through a 106-μm sieve (mesh #140) to produce filtered
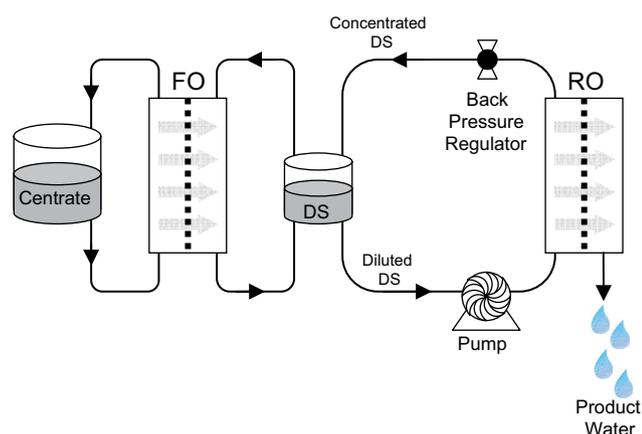

Fig. 1 – A continuous FO/RO bench-scale process setup. The first stage is an FO subsystem for pretreatment of the centrate and the second stage is an RO subsystem that concentrates the DS for the FO stage and produces purified water.

centrate. The DS was prepared using ACS grade NaCl (Fisher Scientific, Pittsburgh, PA) dissolved in DI water.

A schematic drawing of the bench-scale FO system is shown in Fig. 2. The CTA membrane was installed in a modified SEPA cell (GE Osmonics, Minnetonka, MN) that has symmetric channels on both sides of the membrane. This allowed for both the feed solution and the DS to flow tangential to the membrane. The effective membrane surface area was 139 cm^2. Mesh spacers placed in the feed and the DS channels supported the membrane and enhanced mixing. Two variable-speed peristaltic pumps (Cole-Palmer, Vernon Hills, IL) were used to recirculate the feed solution and the DS on opposite sides of the membrane at 1.5 l/min. The temperatures of the feed solution and the DS were monitored with thermocouples installed at the inlets of the SEPA cell.

For each feed solution evaluated, two sets of experiments were conducted: one at constant feed solution concentration and the other at increasing feed solution concentration. The initial DS concentration for all experiments was 70 g/l NaCl, and as water diffused through the membrane, the DS concentration slowly decreased.

In experiments conducted with increasing feed concentration, a 4-l Erlenmeyer flask was used as the feed reservoir and the feed volume was allowed to decrease over time. In experiments conducted with constant feed concentration, DI water was constantly added to the feed reservoir in order to maintain a constant feed volume. In all experiments, the DS reservoir was a 1-l filtration flask. Water that diffused through the membrane into the DS overflowed to a beaker placed on an analytical balance (PB5001-S, Mettler-Toledo, Columbus, OH) interfaced with a computer. The change of weight on the balance was recorded and the data were used to calculate water flux through the membrane.

For each set of experiments, five cycles were conducted. Cycles were conducted for approximately 8 h, at which time the DS concentration reached approximately 30 g/l NaCl. At the end of each cycle, the DS was flushed out of the system and replaced with fresh 70 g/l NaCl DS, and the centrate was replenished with DI water to its original volume without removing or cleaning the membrane. After the fourth cycle, the system and the membrane were cleaned. Membrane cleaning was accomplished by flushing the feed side with 4 l of DI water, recirculating a sodium hydroxide solution adjusted to pH 11 for 30 min, and flushing with 4 l of DI

water. After cleaning, the system was filled with fresh DS and centrate and the fifth cycle was conducted similar to the first four cycles. At the end of each of the five cycles, samples from the DS and the centrate were collected and analyzed for ammonia, TKN, and orthophosphate, and the rejection was calculated for each.

2.2. Evaluation of combined bench-scale FO/RO system

An FO/RO system was utilized to perform continuous bench-scale experiments. This system made it possible to operate the FO process with the DS at a constant concentration. The continuous FO/RO system was an integration of two systems: the bench-scale FO subsystem and a pilot-scale RO subsystem. The RO subsystem was described in a previous publication (Cath et al., 2005b).

Experiments with constant feed solution concentration were conducted with filtered centrate as the feed solution and concentrated NaCl as the DS (50±2 g/l NaCl). The flow rates of both the feed solution and the DS in the modified SEPA cell were 1.5 l/min. The feed temperature was maintained at 23±1 °C with a heat exchanger, and the DS temperature was allowed to equilibrate to 25±1 °C by heat transfer through the membrane. An experiment with increasing feed solution concentrations was conducted in four cycles; each cycle was terminated once 70–80% feed recovery was achieved. At the termination of each cycle, the feed solution was drained from the system, 4 l of fresh filtered centrate was added to the feed reservoir, and the next cycle was started without cleaning or replacing the membrane. At the end of the second cycle, both the feed solution and the DS were drained from the system and the membrane was osmotically backwashed for 10 min.

During osmotic backwashing, water from the support side of the membrane diffused through the membrane toward the active side, thereby backwashing the membrane (Sagiv and Semiat, 2005). This was accomplished by circulating a 50 g/l NaCl solution on the active (feed) side of the membrane and DI water on the support (DS) side of the membrane. At the end of osmotic backwashing, the feed side of the system was flushed with filtered centrate and the DS side of the system was flushed with concentrated brine from the RO subsystem.

After osmotic backwashing, the feed reservoir was filled with 4 l of filtered centrate, the DS reservoir was filled with concentrated brine, and the third cycle was started. After the third cycle was completed, a 20-min osmotic backwash was performed and the fourth cycle was started.

2.3. Comparison between RO and FO for centrate treatment

RO experiments were carried out to compare the performance (fouling rate and contaminant rejection) of RO and FO. Two semi-permeable membranes were evaluated: the LFC-1 (Hydranautics, Oceanside, CA) and the FO CTA membrane. Although the CTA membrane is typically used for FO separation, it was tested here in RO mode for comparison purposes. The LFC-1 membrane is a low-fouling hydrophilic membrane marketed for wastewater applications.
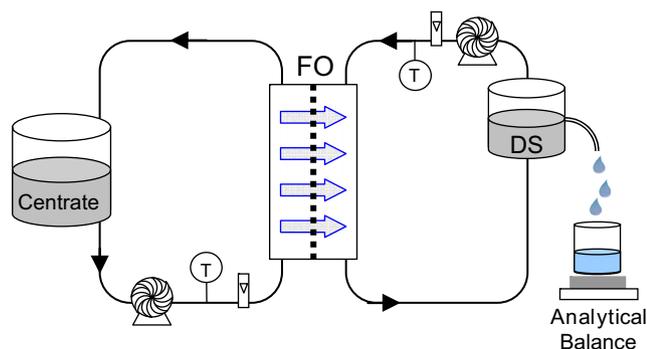

Fig. 2 – Schematic of the bench-scale FO apparatus.

RO experiments were conducted using a stainless-steel SEPA cell. Initial feed pressure was regulated to achieve a flux similar to the initial flux in the FO experiments with a DS concentration of 50 g/l NaCl. The feed pressure was 8.7 bar (120 psig) during the experiment with the LFC-1 membrane and 13.8 bar (200 psig) during the experiment with the CTA membrane. RO experiments were conducted for approximately 60 h with each membrane. At the beginning of each experiment, 10 l of filtered centrate was added to the feed reservoir. After every 8 h, a sample of permeate was taken for analysis, the centrate was drained from the system, and fresh centrate was added to the feed tank. After 48 h, the system was shut down for membrane cleaning. The cleaning cycle was similar to the procedure described in Section 2.1. The experiment was continued for 8 more hours after cleaning.

2.4. Solution chemical analyses

Ammonia concentrations were determined using a Hach spectrophotometer (DR/2400, Hach Company, Loveland, CO) and Hach Method 10031. TKN concentrations were determined using Hach Digestion Method 8075. Orthophosphate concentrations were determined using Standard Methods 4500-P (Franson, 1998). Appropriate dilutions were performed on all samples to allow measurements within the calibration range.

2.5. Process modeling

A model was developed in order to estimate the energy and major hardware required to treat 190 m^3 (50,000 gal) of centrate per day at water recoveries of 50–90%. The model is a combination of two models: an FO model developed in the laboratory and an RO model available from DOW-Filmtec (Midland, MI) called RO System Analysis (ROSA).

The FO model was developed with three FO membrane modules arranged in parallel. Each module was divided into 20 area elements: within each element, the flows, concentrations, and osmotic pressures of both the centrate and the DS were calculated based on values in the upstream element. The average flux in each element was calculated using the flowrate difference between adjacent elements and the surface area of the element. The six input parameters required were the membrane permeability coefficient (determined to be 8.678×10^{-8} m/s bar), the membrane element area, the centrate inlet flowrate and concentration, and the DS inlet flowrate and concentration. The FO model used an iterative numerical method to calculate the system outputs, including feed solution and DS concentrations and flowrates, and the permeation rate. The model did not account for concentration polarization and fouling effects, these were incorporated into the water permeability coefficient.

The outlet DS concentration and flowrate were used as inputs into ROSA. The RO membrane used in ROSA was the SW30HR-320 membrane (DOW Filmtec). The membrane array was concentrate staged with three passes and two membrane elements in each pass. For 90% recovery, an additional element was added to pass 3. The applied pressure used in the model varied based on the recovery of the system. The permeate from each stage was mixed to produce the final product water.

After balancing the two models at each recovery, the specific energy, power, and membrane area requirements were calculated for water recoveries of 50%, 60%, 70%, 80%, and 90% for the FO system (to treat the centrate), for the RO system (to concentrate the brine), and for the combined FO/RO system. The specific energy is the amount of energy required to produce a given volume of water. The specific energy requirement for the FO process was calculated by dividing the energy required to pump 190 m^3/d at 1 hp by the volume of product water produced. The specific energy for the RO process was calculated by ROSA for each of the recoveries. The power consumption was determined by multiplying the specific energies by the production rates at each recovery.

3. Results and discussion

3.1. Evaluation of batch FO at bench scale

3.1.1. Flux analysis

Four sets of batch bench-scale experiments were conducted in order to compare water flux and nutrient rejection for raw and filtered centrate at constant and increasing feed concentrations. Water flux as a function of DS concentration for the raw centrate experiments is illustrated in Fig. 3. Results from each of the five cycles conducted during each experiment are shown. Results from an experiment conducted with DI water as the feed solution are also shown. DI water has negligible osmotic pressure and no fouling potential and, therefore, provides a baseline to which experimental results can be compared.

For both constant feed concentration experiments (Fig. 3a) and increasing feed concentration experiments (Fig. 3b), water flux declined with decreasing DS concentration. As water diffused through the membrane, the DS was diluted and the osmotic pressure driving force and water flux were reduced. Comparing Fig. 3a and b, water flux was lower when the feed concentrations increased during the experiment. The osmotic pressure driving force is a function of the solute concentration on both sides of the membrane. Thus, water flux declines with increased feed TDS concentration. In both cases, it can also be seen that water flux declined between each cycle. The decline in flux between cycles was due to the membrane fouling that occurred during the previous cycle. After cycle 4 the membrane was cleaned and cycle 5 was performed. Water flux during cycle 5 exceeded that of cycles 2, 3, and 4. This increase in flux after cleaning was due to the removal of foulants deposited on the membrane surface. One of the immediate conclusions from the batch experiments is that much of the membrane fouling during treatment of centrate with FO is not permanent and can be readily reversed by proper cleaning.

Results from experiments conducted with filtered centrate are shown in Fig. 4. Comparing results from these experiments with results from the raw centrate experiments (Fig. 3), the flux decline between cycles was substantially less for the filtered centrate than for the raw centrate. This indicates that
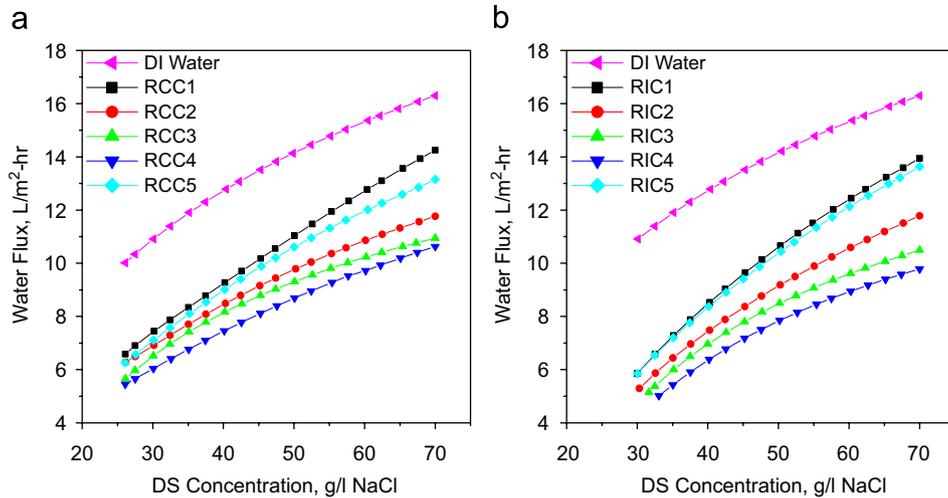
Fig. 3 – Water flux as a function of DS concentration for a batch FO experiment conducted with (a) raw centrate at constant feed solution concentration (RCC) and (b) raw centrate at increasing feed concentrations (RIC). Results from a DI water experiment are provided as a baseline.
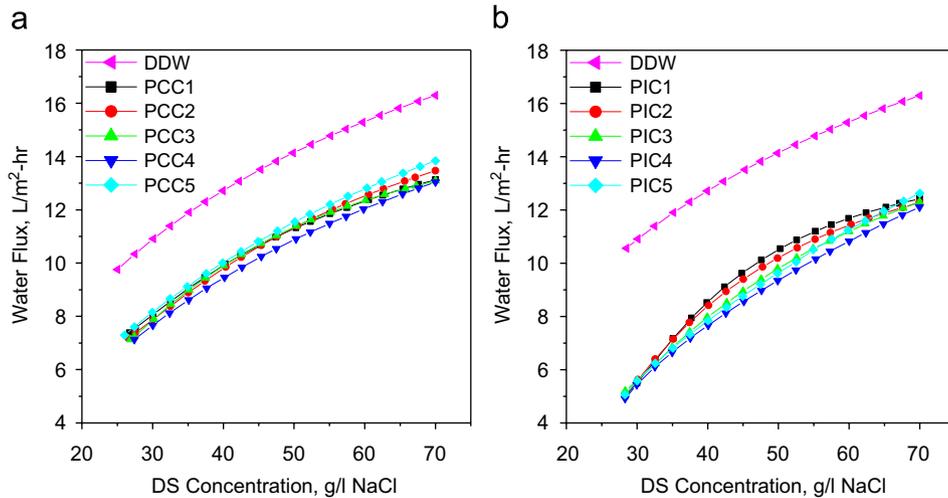

Fig. 4 – Water flux as a function of DS concentration for a batch FO experiment conducted with (a) pretreated centrate at constant feed concentration (PCC) and with (b) pretreated centrate at increasing feed concentrations (PIC). Results from a DI water experiment are provided as a baseline.

pretreatment of the centrate by improving the centrifuge process or by filtration will improve the performance of the FO process.

3.1.2. Nutrient rejection
Nutrient rejections were measured during each of the batch FO experiments. Rejections were calculated using initial concentrations in the feed solution and final concentrations in the DS. Concentrations of ammonia, TKN, and orthophosphate in the raw centrate were relatively constant and averaged 1300, 1400, and 240 mg/l, respectively. The concentrations of these constituents in the DS varied based on the conditions in each experiment.

Results in Table 1 indicate that ammonia and TKN rejections were similar for the majority of the batch FO experiments, most likely due to the fact that the TKN in the centrate was approximately 92% ammonia. Also, ammonia

and TKN rejections were lower than those of orthophosphate for each of the experiments. For semi-permeable membranes, such as the CTA membrane, rejection is generally based on the charge and hydrated radius of the molecules. The pH of centrate during all experiments was 7–8; in this range, the CTA membrane is negatively charged (Cartinella et al., 2006) and therefore, negatively charged ions will have greater rejection than positively charged ions. Thus, the higher rejection of orthophosphate was most likely due to its larger hydrated radius (0.49 nm compared with 0.11 nm of ammonium (Voorthuizen et al., 2005), and its negative charge ($H_2PO_4^-$ and HPO_4^{2-}). Ammonia rejection, although less than orthophosphate, was still greater than 80% because a large fraction of the ammonia is present as the positively charged ammonium ion (NH_4^+), which is rejected better than the uncharged ammonia molecule (NH_3). Similar results were reported by Voorthuizen et al. (2005) and Oron et al. (2006)

Table 1 – Ammonia, TKN, and orthophosphate rejections for the FO and RO experiments

Experiment	Rejection (%)		
	Ammonia	TKN	Orthophosphate
FO-raw centrate increasing feed concentrations	92.1	91.0	99.9
FO-raw centrate constant feed concentration	84.7	85.0	99.6
FO-treated centrate increasing feed concentrations	82.9	91.6	99.8
FO-treated centrate constant feed concentration	88.0	89.1	99.8
RO/CTA-treated centrate constant feed concentration	95.4	96.5	99.9
RO/LFC-1-treated centrate constant feed concentration	84.3	84.7	99.8

Initial feed concentrations of ammonia, TKN, and orthophosphate averaged 1300, 1400, and 240 mg/l, respectively.

using RO membranes for nutrient removal from secondary effluent and sanitary systems.

3.2. Evaluation of continuous FO/RO at bench scale

All FO experiments up to this point were performed in a batch mode with the DS being diluted as the experiments proceeded. In order to evaluate the process in a continuous mode and to eliminate the influence of declining DS concentration on flux, the FO process was combined with an RO subsystem that supplied a stream of DS at a constant concentration. Two sets of experiments were performed: increasing feed concentration experiments and constant feed concentration experiments.

In the increasing feed concentration experiments (Fig. 5), the experiments were conducted in four cycles; after each cycle, a new batch of centrate was loaded and after cycles 2 and 3, osmotic backwashing cycles were performed for 10 and 20 min, respectively. For each cycle, water flux decreased with time. The decline in flux was due to both an increase in the osmotic pressure of the feed solution and membrane fouling. It is difficult to differentiate between the effects of increased osmotic pressure and membrane fouling on flux decline in a given cycle, but the effect of fouling between cycles is apparent. The initial flux of cycle 1 was approximately 10.5 l/m²h, whereas the initial flux of cycle 2 was approximately 9 l/m²h. The decrease in initial flux between cycles 1 and 2 was clearly due to the membrane fouling that occurred during cycle 1.

It was anticipated that the initial flux in cycle 3 (after the 10-min osmotic backwashing) would be greater than the initial flux in cycle 2 (prior to any membrane cleaning) because the cleaning should have removed foulants that had been deposited on the membrane surface. However, this was not the case; the 10-min osmotic backwash at the end of cycle 2 appears not to have removed the foulants as well as the 20-min osmotic backwash at the end of cycle 3, which restored the initial flux of cycle 3 but did not necessarily improve it. However, it is interesting to note that the flux declined less rapidly after the 10- and 20-min osmotic backwashes (cycles 3 and 4) than it did in the previous two cycles (cycles 1 and 2). A precision analysis was conducted on the slope of each cycle to confirm this; results (not shown)
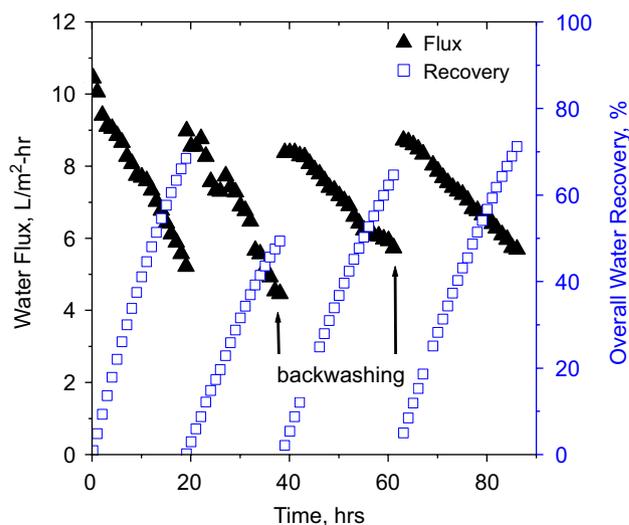

Fig. 5 – Water flux and recovery as a function of time for FO experiments conducted with increasing feed concentration. The feed solution was filtered centrate and the DS was NaCl solution at a relatively constant concentration of 50 ± 2 g/l. The feed solution was replenished prior to each cycle. The membrane was osmotically backwashed for 10 min prior to cycle 3 and 20 min prior to cycle 4.

revealed that the slopes of cycles 1 and 2 are steeper and statistically different from those of cycles 3 and 4.

During operation, both pore blocking and surface fouling reduce the performance (flux) of the process. Chemical cleaning typically removes the cake layer only and cannot remove foulants in the membrane pores. It is hypothesized that during osmotic backwashing these foulants are flushed out while the cake layer may still be partially intact and adhered to the membrane surface. Upon resumption of the process, the cake layer continues to act as a secondary membrane and reduces the recurrence of pore plugging—hence, the slower flux decline after osmotic backwashing. Research is under way to confirm this hypothesis. Nevertheless, these preliminary results indicate that osmotic backwashing may be an efficient method of membrane cleaning in osmotic-driven membrane processes and that the duration of the backwashing cycle may play an important role in its efficiency.

Constant feed concentration experiments were used to isolate and more clearly evaluate the effects of membrane fouling on flux decline. In these experiments, the feed solution was filtered centrate that was replaced approximately every time 2 l of water permeated through the membrane (approximately every 10 h). The DS was an NaCl solution at a relatively constant concentration of 50 ± 2 g/l. The initial flux was approximately 10 l/m^2 h (a relative flux of 1.0 is shown in Fig. 6), and was relatively stable for the first 20 h (approximately 4 l cumulative permeate and prior to the second feed replacement). After the second replenishment (~4 l cumulative permeate), water flux noticeably declined at a constant rate until membrane cleaning was performed. Because both the DS and the feed solution were maintained at constant concentrations, the observed flux declines were primarily due to membrane fouling. Increased flux decline with consecutive replenishments was the result of additional foulants in the system. Prior to the fourth replenishment, mild cleaning with an NaOH solution at pH 11 improved the flux for the CTA membrane in RO mode and almost completely restored the initial water flux for the CTA membrane in the FO mode. The LFC-1 membrane suffered severe fouling and very minimal flux restoration was observed.

3.3. Evaluation and comparison of FO and RO at bench scale

Fig. 6 can also be used to compare the performances of FO and RO for centrate treatment. For both membranes and pro-

cesses, water flux declined as more water permeated through the membrane. From the four possible causes for flux decline in RO processes (i.e., membrane compaction, fouling, increased feed osmotic pressure, and scaling), it can be deduced that fouling is the main contributor to the observed flux decline. First, both membranes were compacted with DI water prior to the start of the centrate experiments. Second, the feed solutions in both RO experiments were maintained at constant concentrations by adding DI water to the feed solution as water crossed the membrane; therefore, no changes in osmotic pressures was expected and the driving force should have been constant. And third, scaling is also unlikely for the same reason (i.e., no conditions for salts to reach saturation).

In RO, the deposition of foulants on the membrane surface occurs very rapidly and is controlled by permeate flux rate and membrane–foulant interactions (Lee and Elimelech, 2006). As a result, a cake layer forms on the membrane, the hydraulic resistance of the membrane increases, and water flux decreases. Results in Fig. 6 indicate that the effects of fouling differed for the two membranes investigated under RO conditions. The difference in flux decline is mostly apparent during the permeation of the first 2 l (~10 h), when the rate of flux decline of the CTA membrane was greater than that of the LFC-1 membrane. The greater flux decline in the CTA experiment can be associated with both the applied pressure and the membrane surface characteristics.

The hydrostatic pressure applied during the CTA-RO experiment was 80 psi greater than that applied during the LFC-1 RO experiment; the greater compaction of foulants on the CTA membrane may have resulted in greater flux decline. After the first 10 h of operation, the rate of flux decline was approximately the same for both membranes. It is likely that at this point membrane–foulant interactions were no longer the predominant fouling mechanism.

In general, cellulose acetate membranes used for RO are considered to be smooth. However, the CTA membrane which is cast over a polyester fine mesh (McCutcheon et al., 2006), has dented areas where the CTA polymer is not in contact with the mesh. Attempts to measure the surface roughness of the CTA membrane using atomic force microscopy were inconclusive due to the undulation of the membrane surface. However, it is believed that these elevation differences on the CTA membrane can attract and trap foulants on the membrane and enhance cake formation (Vrijenhoek et al., 2001). Further research into the fouling mechanisms of such a unique membrane is necessary in order to fully understand the causes of flux decline under RO conditions.

After approximately 50 h of operation (6–7 l of water recovered), the membranes were cleaned with an NaOH solution at a pH of 11. After cleaning of the CTA membrane, the water flux increased by approximately 24%, and after cleaning of the LFC-1 membrane, the water flux increased by less than 5%. Although the flux declined more rapidly during the initial stages of the CTA experiment, it appears that the foulants were more easily removed and the flux was more easily restored after cleaning compared with the LFC-1 membrane.

Comparing the performances of RO and FO, it is clear that water flux declined more rapidly in the CTA-RO experiment
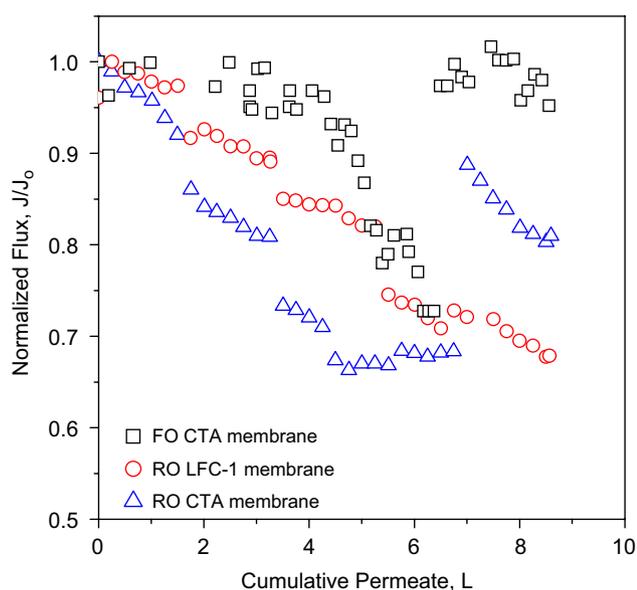

Fig. 6 – Relative water flux as a function of water produced for two RO experiments and one FO experiment. The three experiments were conducted with a feed solution of filtered centrate at constant concentration and a DS of NaCl solution at a relatively constant concentration of 50 ± 2 g/l. Each experiment was conducted with three centrate replenishments and one chemical membrane cleaning.

than in the CTA-FO experiment. This was most apparent for the first 4l of water produced; during this period, the flux declined by 3 and less than $1 l/m^2 h$ for the CTA-RO and CTA-FO experiments, respectively. Because there was no applied pressure in FO, cake-layer compaction was reduced or eliminated and flux decline was not as severe. After 4.5l of water was produced, the flux in the FO experiment started declining due to accumulated fouling. The total flux decline in the FO experiment, from the beginning to just prior to cleaning, was approximately $2 l/m^2 h$, compared with $4 l/m^2 h$ for the CTA-RO experiment.

After membrane cleaning, the CTA-FO experiment exhibited close to 100% flux restoration, whereas the CTA-RO experiment exhibited approximately 24% flux restoration. The reason for the higher flux restoration of the CTA-FO experiment is likely because the foulants that were deposited on the membrane during the FO experiments were more easily removed than those in the RO experiment. During the RO experiments, the foulants were compacted because of the hydraulic pressure and because pore blockage was very likely; in FO, hydraulic pressure was not applied on the membrane, and therefore, less effort was required to remove the foulants from the membrane surface.

3.3.1. RO rejection analysis

Rejections of ammonia, TKN, and orthophosphate were calculated for both RO experiments (Table 1). Similar to the results for FO, orthophosphate rejection was high and ammonia and TKN rejections were similar. Ammonia and TKN rejections were lower for the experiment conducted with the LFC-1 membrane. This may have been due to the fact that the LFC-1 membrane is a looser RO membrane than the CTA membrane or that the ammonia has a higher affinity to diffuse through the polymer used to make the LFC-1 membrane.

3.4. Process modeling

Pilot system modeling was performed in order to estimate the specific energy, power, and membrane area requirements to treat $190 m^3/d$ (50,000 gal/d) of centrate. Specific energy as a function of percent water recovery is illustrated in Fig. 7. Specific energy is defined as the power demand per unit water production rate (e.g., $1 m^3$ of water). Results in Fig. 7 indicate that the total specific energy requirement for the system is approximately $4 kWh/m^3$, with the FO system requiring approximately 25% and the RO system requiring approximately 75% of the total specific energy demand. Results in Fig. 7 also indicate that there may exist an optimum operating condition between 70% and 75% water recovery.

The FO system had lower specific energy requirements than the RO system because the FO system requires minimal pressure to treat the water. Also, as the percent recovery increases, the specific energy for the FO system decreases because more water is produced due to increase in surface area and not due to pumping or applied pressure. In the RO system, the specific energy increases with increasing recovery because higher pressure is needed to increase water production rates.
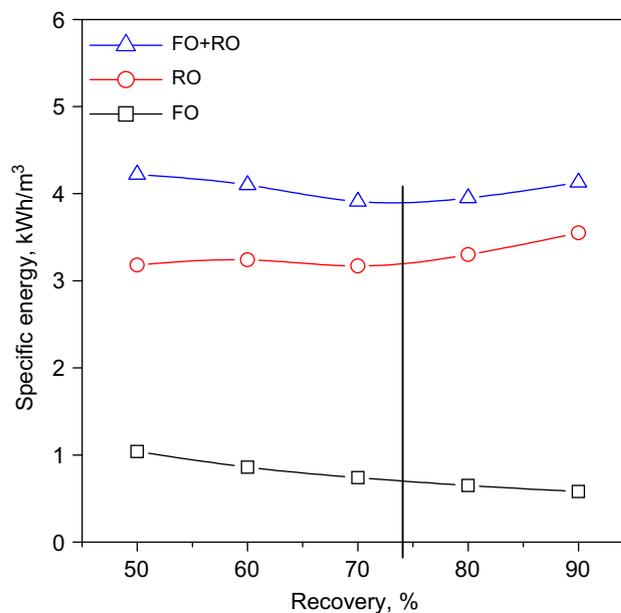

Fig. 7 – Specific energy as a function of recovery for the FO, RO, and combined FO/RO systems. The specific energy was based on the energy needed to treat $190 m^3/d$ of centrate at 50%, 60%, 70%, 80%, and 90% recoveries. The vertical line represent the minimum specific energy at approximately 74% recovery.
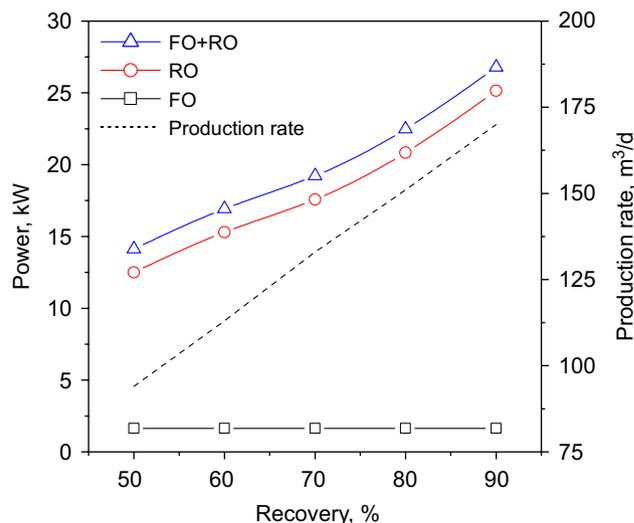

Fig. 8 – Power requirement and production rate as a function of recovery for the FO, RO, and FO/RO systems. The power requirements were based on the power needed to treat $190 m^3/d$ of centrate at 50%, 60%, 70%, 80%, and 90% recoveries. The production rate was the total amount of water produced per day at the given recoveries.

In order to determine the total power requirements for the FO, RO, and FO/RO pilot systems, the specific energy was multiplied by the total amount of water produced at each recovery. Power requirements and water production rates as a function of percent water recovery are summarized in Fig. 8.

Results indicate that the total power requirement for the combined FO/RO pilot-scale system increases from approximately 14 kW at 50% recovery to approximately 27 kW for 90% water recovery. The increase in total power was primarily due to increasing power requirements of the RO system at elevated recoveries. In FO, however, the power requirements remained constant throughout the range of recoveries mainly due to the fact that the feed solution flowrate was constant for all recoveries and thus, the power requirements did not change. Both the specific energy and power requirements for the FO system were modeled at a constant feed solution flowrate and initial DS concentration.

In FO, as with other membrane processes, the amount of recovered water depends on the surface area of the membrane that is in contact with the solution. Therefore, in order to obtain recoveries of 50%, 60%, 70%, 80%, and 90% in the model, the membrane area was modified until the given recovery was obtained. The membrane area (m^2) required to achieve the range of recoveries is illustrated in Fig. 9. Results in Fig. 9 indicate that the required membrane area for the FO/RO system increased from approximately 800 m^2 for 50% water recovery to approximately 1900 m^2 for 90% water recovery. This is primarily due to the increase in membrane area needed for the FO system to obtain the given recoveries. For a given recovery, the FO system required more membrane area because the driving force for mass transport constantly decreased as the feed solution became more concentrated and the DS stream was diluted. At higher recoveries, considerably more FO membrane area was required because the driving force was further reduced. Recovery in the RO system, however, was primarily increased by increasing the pressure (driving force); and therefore, the required membrane area was fairly constant for each recovery.
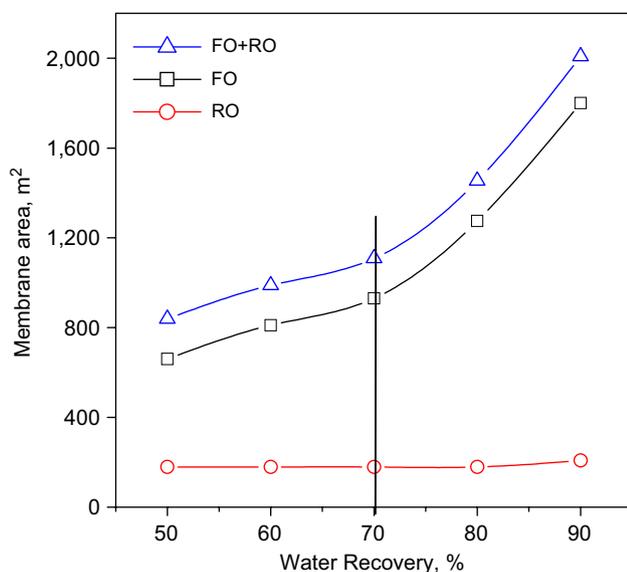
The results in Fig. 9 also indicate that the slope of required membrane area versus water recovery substantially increased above approximately 70% recovery. Therefore, in order to optimize the FO process, the system should be operated in a range that allows for high recovery but does not require excess amounts of membrane. From the results in Fig. 9, this optimum occurs at approximately 70% recovery.

4. Conclusions

It was demonstrated from the bench-scale FO evaluation that FO is capable of concentrating both raw and pretreated centrate and providing high rejection of nutrients of interest. The performance, and specifically flux, increased when the centrate was pretreated prior to the FO operation. Furthermore, FO was effective at concentrating centrate over a wide range of DS concentrations; however, higher fluxes were achieved with higher DS concentrations. The combined FO/RO system maintained long-term (~50 h) process performance. The water flux of the combined FO/RO system was influenced by both centrate replenishment and centrate concentration; it decreased with increased replenishments and concentrations. The combined FO/RO system outperformed the stand-alone RO process for centrate concentration.

Results from FO and RO modeling indicated that the RO system required the majority of the power, and the FO pretreatment process required the majority of the membrane area. Preliminarily results indicated that operating a pilot-scale system at approximately 70% recovery will optimize power and membrane area requirements. For a 190 m^3/d (50,000 gal/d) FO/RO pilot-scale system, 133 m^3/d (35,000 gal/d) of high-quality product water could be generated and used for beneficial purposes.

Fig. 9 – Membrane area as a function of recovery for the FO, RO, and FO/RO systems. The membrane area was based on the area needed to treat 190 m^3/d of centrate at 50%, 60%, 70%, 80%, and 90% recoveries.

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