Evaluation of Membrane Processes for Reducing Total Dissolved Solids Discharged to the Truckee River

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Abstract: Truckee Meadows Water Reclamation Facility (TMWRF) is a 150,000 m^3/day (40 mgd) tertiary wastewater treatment facility that serves the cities of Reno and Sparks, Nev. The effluent from TMWRF is discharged into the Truckee River which flows to Pyramid Lake—a very sensitive ecosystem and habitat for endangered species. Reverse osmosis (RO) and nanofiltration (NF), in conjunction with ultrafiltration (UF) pretreatment, were evaluated for total dissolved solids (TDS) and nutrient removal from the effluent of TMWRF at bench and pilot scale. Results from short-term pilot-scale tests showed that RO and NF membrane processes can successfully remove both TDS and nutrients from the effluent when paired with coagulation-enhanced UF pretreatment. NF membranes were able to achieve the necessary removal while maintaining higher fluxes and lower specific power consumption.

DOI: 10.1061/(ASCE)0733-9372(2007)133:12(1136)

CE Database subject headings: Water reclamation; Wastewater treatment; Salinity; Membranes; Filtration; Fouling.

Introduction

Truckee Meadows Water Reclamation Facility (TMWRF) is a 150,000 m^3/day wastewater treatment facility that discharges into the Truckee River—the primary source of water for Pyramid Lake (Fig. 1). Conservative substances, such as salt, discharged to the river, accumulate in Pyramid Lake (a terminal lake), and have caused the salinity of the water to almost double since 1933 to approximately 5,000 mg/L total dissolved solids (TDS) today. The total maximum daily load (TMDL) for TMWRF's discharge of TDS has been set at 54,500 kg/day. Nutrients discharged to the river can promote eutrophication, destroy sensitive ecosystems, and threaten endangered species (e.g., the cutthroat trout and cui-ui fish) in the river and lake. The TMDLs for TMWRF's discharge of nitrogen and phosphorous have been set at 227 and 61 kg/day, respectively.

To accommodate future growth in the Reno/Sparks metropolitan area, the facility is expected to reach a new capacity of 176,000 m^3/day in approximately 15 years. Using the current average effluent TDS concentration of 375 ppm (TMWRF 2006), the expansion would result in a TDS discharge of approximately 66,000 kg/day. To satisfy the TMDL for TDS discharge to the Truckee River, TMWRF would have to reduce its TDS loading to the Truckee River by approximately 11,500 kg/day. Thus, the average TDS concentration would have to be reduced to approximately 310 ppm for the design flow of 176,000 m^3/day. Eventually, nitrogen and phosphorous may also have to be removed from the effluent in order to maintain their TMDLs.

Reverse osmosis (RO) and nanofiltration (NF) are pressure-driven membrane processes used to remove TDS and nutrients from water. Many studies and full-scale applications have established RO as a suitable membrane process for wastewater reclamation and reuse. Examples of successful RO wastewater reclamation projects are summarized in two comprehensive studies (Reardon et al. 2005; Wintgens et al. 2005). However, fewer studies (Bellona and Drewes 2005; Voorthuizen et al. 2005; Nghiem and Schäfer 2006) have been published on the use of NF for wastewater reclamation, and particularly for removal of TDS and nutrients from secondary or tertiary effluent.

When using membrane processes in wastewater treatment, the quality of the feedwater is of critical importance due to problems associated with membrane fouling (Abdel-Jawad et al. 1997, 2002; Wilf and Alt 2000; Wintgens et al. 2005). Dissolved matter, colloids, suspended particles, and microorganisms that are present in secondary or tertiary effluent will accumulate on the membrane surface and result in loss of performance, possible membrane damage, and increased operating costs. Feed pretreatment prior to RO or NF membrane processes may reduce the fouling potential.

Studies (Chakravorty and Layson 1997; Reardon et al. 2005) have shown that granular media filtration is typically not an effective pretreatment for RO or NF processes. Specific contaminants, such as total suspended solids (TSS) and total organic carbon (TOC) can exist in the filtered effluent and may compromise downstream membrane operations. Knowledge of the filter effluent water composition is important in determining if filtration is an appropriate pretreatment process. The data in Table 1 imply that the filter effluent quality at TMWRF might be too low for membrane processes such as RO and NF.

Microfiltration (MF) and ultrafiltration (UF) are pressure-driven membrane processes that can be used in place of or in

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Note. Discussion open until May 1, 2008. Separate discussions must be submitted for individual papers. To extend the closing date by one month, a written request must be filed with the ASCE Managing Editor. The manuscript for this paper was submitted for review and possible publication on April 12, 2006; approved on April 26, 2007. This paper is part of the ***Journal of Environmental Engineering***, Vol. 133, No. 12, December 1, 2007. ©ASCE, ISSN 0733-9372/2007/12-1136–1144/ $25.00.

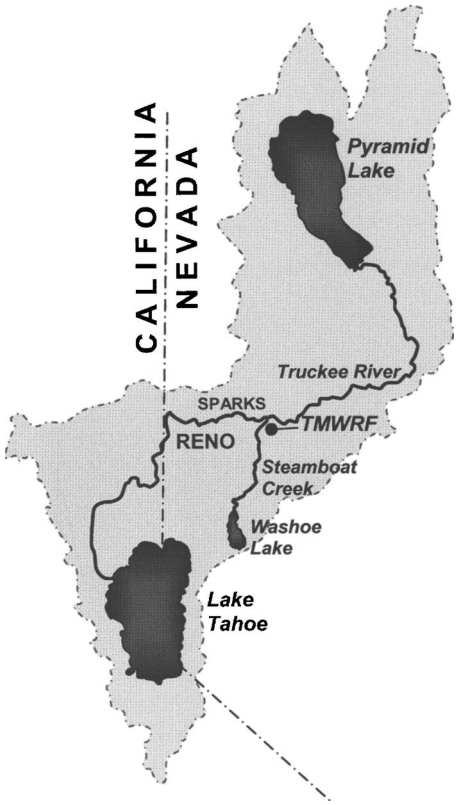

Fig. 1. Illustrative map of Truckee River watershed (adapted from Oden 2003, with permission from *Reno Gazette-Journal*)

addition to conventional media filtration for RO or NF pretreatment (Chakravorty and Layson 1997; Ebrahim et al. 1997). The advantage of using MF or UF systems is that they provide a consistent and high quality feedwater for RO or NF (Van Hoof et al. 1999; Durham et al. 2001; Brehant et al. 2002). For example, most RO and NF membrane manufacturers use silt density index (SDI) as a measure of the feedwater quality required for optimal operation of their membranes. An SDI of 5 or less is generally required for optimal membrane performance (Abdel-Jawad et al. 1997). It has been shown that effluent from conventional wastewater treatment processes has an SDI of 5–6, whereas effluent from MF can have an SDI as low as 1–2 (Durham et al. 2001).

Gabelich et al. (2003) compared conventional treatment (coagulation, flocculation, sedimentation, and dual-media filtration) and conventional treatment with preozonation and biologically active dual-media filtration to MF for use as pretreatment for RO. Following testing, the RO membranes were examined using scanning electron microscopy (SEM) and atomic force microscopy (AFM). Based on the SEM and AFM data, the RO membranes exposed to conventionally pretreated water had a much higher concentration of particles deposited on the membrane surface than the RO membranes exposed to MF-pretreated water. Consequently, Gabelich et al. concluded that MF and UF pretreatment will be more widely applied and will likely become the industry standard for RO pretreatment in wastewater reclamation applications (Ebrahim et al. 1997, 2001; Ghayeni et al. 1998; Durham et al. 2001).

The objective of the current investigation was to evaluate the feasibility, efficiency, and limitations of using RO or NF membrane processes to reduce TDS and nutrient loading from TMWRF into the Truckee River. Specifically, this study focused on selecting appropriate commercially available RO and NF membranes to efficiently remove TDS from TMWRF's effluent water; to evaluate the performance of NF membranes compared to RO membranes in treating reclaimed wastewater under varying operating conditions; to determine if and what additional pretreatment of the feedwater is necessary in addition to the existing dual media filtration used in the treatment plant; and to calculate a preliminary cost estimate for implementing membrane processes at TMWRF.

Materials and Methods

Membrane Selection

Three NF membranes and seven RO membranes were selected for preliminary screening. The characteristics of these membranes are summarized in Table 2. Stirred-cell (HP4750, Sterlitech, Kent, Wash.) tests were performed prior to bench- and pilot-scale testing to obtain short-term performance data and to select membranes for extended bench-scale testing. Each of the NF and RO membranes were tested for 90 min under a feed pressure of 10.4 and 17.2 bars (150 and 250 psi), respectively. The stirring rate was constant at 250 rpm. Water flux and TDS rejection were used as the criteria for selecting the membranes for bench-scale testing.

Bench-Scale Studies

The objective of the bench-scale testing was to obtain longer-term performance data and to select the three most suitable membranes for the pilot-scale testing. A closed-loop bench-scale test apparatus was used in evaluating the membranes selected during the stirred-cell experiments. The feed solution was contained in a 30-L temperature-controlled reservoir (Polystat, Cole Parmer, Vernon Hills, Ill.). The feed solution was fed into two flat-sheet membrane test cells (each with 1,935 mm^2 active area) operating in parallel using a positive displacement pump (Model 1605A, Procon Pumps, Murfreesboro, Tenn.). The feed flowrate was controlled using a bypass valve on the feed line. Transmembrane pressure was controlled using a valve installed on the concentrate stream. The flowrate, conductivity, pressure, and pH of both the concentrate and permeate were measured using in-line digital probes (Cole Parmer, Vernon Hills, Ill.) connected to a computer. The concentrate and permeate were recirculated back to the refrigerated reservoir.

Table 1. Average Concentrations of Major Constituents in TMWRF's Effluent Water (Adapted from TMWRF 2006)

Annual average	BOD (ppm)	TSS (ppm)	Total P (ppm)	TDS (ppm)	NH_3 (ppm)	NO_3 (ppm)	Total N (ppm)	pH
2003	4.0	2.0	0.29	372	0.13	0.2	1.74	7.23
2004	5.0	2.0	0.33	376	0.23	0.56	2.22	7.30
2005	5.0	2.0	0.34	393	0.70	0.15	2.69	7.29

Table 2. Selected Membrane Properties Provided by Manufacturers for NF and RO Membranes Used in This Study

	Membrane designation	Membrane type	Manufacturer	Average rejection (%)	Maximum pressure (psi)	Typical operating pressure (psi)	Range (pH)
A	HL	NF	Osmonics	98% $MgSO_4$	400	70–300	2.0–9.0
B	NF-90	NF	FilmTec	99% $MgSO_4$	250	150	2.0–9.0
C	DK	NF	Osmonics	98% $MgSO_4$	500	70–400	2.0–11.0
D	FT-30	RO	FilmTec	99.0% NaCl	1,000	250	2.0–11.0
E	CD	RO	Desal. Systems	98.5% NaCl	450	140–400	5.0–6.5
F	CE	RO	Desal. Systems	97.5% NaCl	450	140–400	5.0–6.5
G	CG	RO	Desal. Systems	89.0% NaCl	450	60–200	5.0–6.5
H	LFC-1	RO	Hydranautics	99.5% NaCl	600	225	3.0–10.0
I	SG	RO	Osmonics	98.5% NaCl	500	200	2.0–11.0
J	TFC-HR	RO	FilmTec	99.7% NaCl	500	225	3.0–11.0

Prior to each experiment, the membrane was compacted by filtering deionized water for 6 h until no further flux decline was observed. Membranes were compacted at the manufacturers' recommended operating pressures. The compression process ensured that flux decline observed during testing was due to membrane fouling and not membrane compaction. Each membrane was tested for approximately 38 h. Due to the length of the experiments, membrane samples were not run in duplicate. However, prior to performing the experiments reported in this paper, duplicate experiments were run and were found to agree within 10–15%. Feed flow rate was kept constant at 1.25 L/min (0.33 gpm) through each cell (with a cross-flow velocity of approximately 0.28 m/s) and temperature was maintained at 20°C. Because constant pressure was maintained throughout each experiment, a loss of permeate flux with time indicated membrane fouling.

Pilot-Scale Studies

The main objective of pilot-scale testing was to evaluate TDS removal using actual continuous-flow source water and commercial-size membrane elements. Pilot-scale tests were performed on a commercial system (E4-6600-DLX-60, GE Osmonics, Minnetonka, Minn.) equipped with three standard 4040 membrane elements [0.1 m (4 in.) in diameter and 1.02 m (40 in.) long] connected in series.

Each membrane was tested for 48 h at each of three feed pressures (12.0, 14.5, and 17.6 bars), for a total of 144 h. The membranes were chemically cleaned between each test using the manufacturers' cleaning procedures for organic fouling. A constant water recovery of 50% was maintained throughout the experiments so that flux and rejection results from each membrane could be easily compared. Permeate flux and conductivity were monitored routinely.

Implementation of RO or NF membrane technology at TMWRF will also enhance the removal of TOC and nutrients from the treated wastewater effluent. The rejection of TOC, total nitrogen, and total phosphorous were measured for the three membranes tested at pilot scale. Samples were collected from the feed and permeate approximately 24 h after the beginning of each experiment. TOC was measured using an ultraviolet (UV) persulfate method TOC analyzer (detection limit 0.5 μg/L) (Tekmar-Dohrmann Phoenix 8000, Cincinnati, Ohio) and total nitrogen and total phosphorous were measured using colorimetric methods [Standard method SM-4500-NO for nitrogen (detection limit 0.5 mg/L) and EPA method 365.4 for phosphorous (detection limit 0.02 mg/L)].

Feedwater Characteristics

The feedwater used for all experiments was TMWRF effluent taken from the plant outfall, immediately after dechlorination. For stirred-cell and bench-scale experiments, fresh samples were collected and brought to the laboratory. Pilot-scale experiments were conducted at TMWRF using feedwater pumped directly from the outfall. The average concentrations of the major constituents in TMWRF's effluent are summarized in Table 1.

Pilot-Scale UF Pretreatment System

A fully automated pilot-scale UF pretreatment system was designed, built, and tested at TMWRF. The treatment train (Fig. 2) consisted of a 100 μm disk prefilter followed by a hollow fiber UF membrane element (Hydracap LD-40, Hydranautics, Oceanside, Calif.). In the UF system, water is pumped from the effluent stream through the UF element and the filtrate flows to the RO feed tank. A bypass valve is located at the pump to control flow; a backpressure valve is located downstream of the UF element to control transmembrane pressure. Flow can be reversed to provide clean water from the RO product tank for backwashing. A control

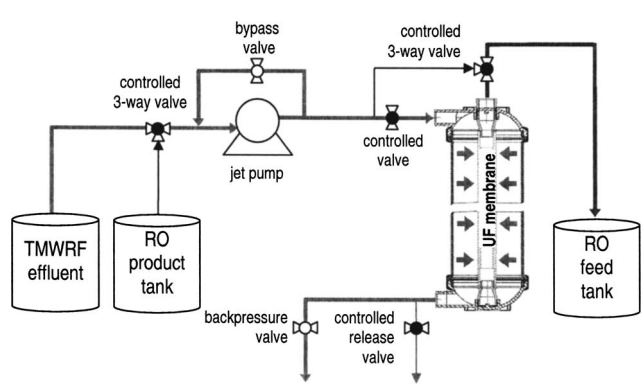

Fig. 2. Flow diagram of pilot-scale UF pretreatment system

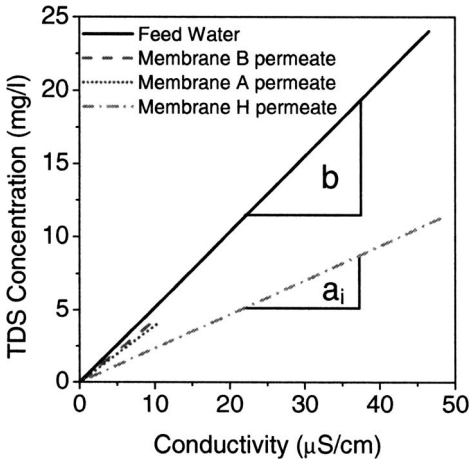

Fig. 3. Correlations between conductivity and TDS concentration for feed and permeate streams from Membranes A, B, and H. For feedwater, R^2=0.972; for Membrane A, R^2=0.96; for Membrane B, R^2=0.95, for Membrane H, R^2=0.98.

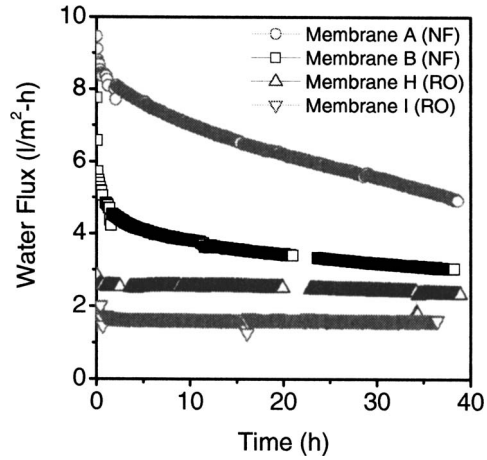

Fig. 4. Specific water flux as function of time for four membranes tested at bench scale. Specific flux is used for comparison because membranes were tested at manufacturers recommended operating pressures. Tests were run for approximately 38 h.

system (Delta, Surrey, BC, Canada) was incorporated to automatically backwash the membrane at preset time intervals.

The operating conditions of the UF system, including transmembrane pressure, feed flow rate, and backwash frequency were optimized to provide adequate and continuous supply of pretreated water to the RO/NF pilot system. To confirm that the UF system adequately reduces fouling potential, TSS [standard EPA method 160.2 (EPA 1983)] and TOC analyses were conducted on the UF feed and product water.

Previous studies (Wilf and Alt 2000; Judd and Hillis 2001) have shown that in certain circumstances, coagulant addition to the feedwater of microporous membrane processes can improve operations. Thus, alum addition prior to UF was also evaluated. ACS grade aluminum sulfate (alum) (Fisher Scientific, Pittsburgh, Penn.) was used for coagulation in the current study.

TDS Measurements

Electrical conductivity (EC) can be used to estimate the TDS or concentration of ions in water by measuring a solution's ability to carry a charge. Because of the ease of using conductivity to calculate TDS rejection and because it has been previously reported in the literature (Sapari et al. 1996; Shaheen 2000; Hench et al. 2003), this method was used to provide continuous and real-time TDS measurement in the current investigation.

However, it was noticed during preliminary data analysis that under certain circumstances (most likely when uncharged particles and organic matter are present) the TDS rejection based on conductivity measurement could be as much as 30% less than that based on direct TDS measurement. To account for this discrepancy, correlations between conductivity measurements and TDS concentrations were established for the TMWRF effluent and product waters used in this study.

To develop correlation equations for conductivity and TDS, samples were collected from the feed and permeate streams for each of the membranes tested. The conductivity of each of the samples was measured and the TDS concentration was analyzed using the standard EPA method 160.2 (EPA 1983). TDS concentrations were plotted against conductivity and linear regressions were developed for the feed stream and three permeate streams as illustrated in Fig. 3. The correlations were then used to calculate TDS rejection using Eq. (1)

$$R_{\mathrm{TDS}} = \left(1 - \frac{C_P}{C_F}\right) * 100 = \left(1 - \frac{\mathrm{EC}_P/a_i}{\mathrm{EC}_F/b}\right) * 100 \qquad (1)$$

where C_P and C_F=permeate and feed TDS concentrations, respectively; EC_P and EC_F=permeate and feed conductivities, respectively; a_i=slope of correlation between a specific permeate stream conductivity and its TDS concentration; and b=slope of correlation between the feed stream conductivity and its TDS concentration. The slope for the feedwater is steeper than those for the three permeate streams because the feedwater has a larger fraction of solids that contribute to TDS measurements but not to conductivity. Membranes A and B are shown over a much shorter range than Membrane H because they had much higher rejections and much lower permeate conductivities.

Results and Discussion

Bench-Scale Studies

Four membranes were chosen for bench-scale investigation based on the results from the short-term stirred-cell experiments. These included two NF membranes (A and B from Table 2) and two RO membranes (H and I). NF Membrane C and RO Membranes E, F, and G were eliminated because they had lower specific fluxes compared to the other NF and RO membranes. RO Membrane J had similar flux to RO Membrane H, but lower rejection. A substantial difference in the initial water flux was observed for the four membranes (Fig. 4). As expected, both NF membranes had higher initial fluxes compared to the RO membranes. A large difference in the initial specific flux was also observed between the two NF membranes, due to the inherent differences in membrane resistance. As time progressed, fouling effects resulted in considerable flux decline for the two NF membranes. The large drop in flux relative to the RO membranes can be attributed to pore plugging and enhanced concentration polarization due to the higher water flux for the NF membranes (Bruggen and Vandecasteele 2001).

Results for TDS rejection as a function of time are shown in Fig. 5 for the four membranes. It was expected that the RO mem-

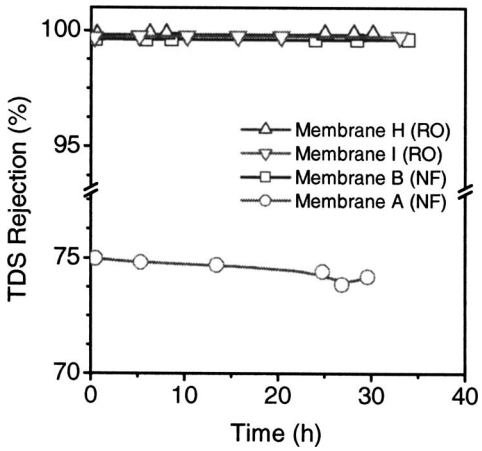

Fig. 5. TDS rejection as function of time for bench-scale experiments. Tests were run for approximately 38 h.

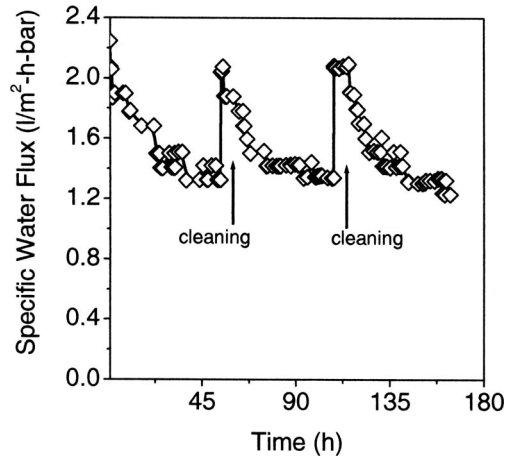

Fig. 6. Water flux as function of time for Osmonics AG4040TF membranes operated with tertiary effluent feedwater. Cleaning was performed after 45 and 100 h.

branes would have a higher rejection than the NF membranes; however, Membrane B, an NF membrane, also had a very high TDS rejection. Membranes B, H, and I had greater than 99.5% TDS rejection; Membrane A rejected less than 80%. Based on this and the high flux observed for Membrane A, it was confirmed that Membrane A is indeed a "looser" NF membrane. This observation was not necessarily anticipated based on the manufacturers' data (Table 2) where the average rejections of Membranes A and B were reported in % $MgSO_4$ rejection and were only different by 1%.

Based on the bench-scale test results, Membranes A, B, and H were selected for pilot-scale testing. Membrane H (RO) was selected because it had a higher initial flux than Membrane I, minimal flux decline, and high TDS rejection. Membrane B (NF) was selected because it had a higher flux than the RO membranes and it also maintained high TDS rejection throughout the 38 h bench-scale test. Membrane A (NF) was selected to represent a looser NF membrane with high flux and low rejection.

Pretreatment Analysis

Because membrane fouling is an obstacle for efficient implementation of membrane processes at wastewater treatment plants, pretreatment needs to be carefully considered. Prior to pilot-scale testing of the three selected membranes, pretreatment experiments were conducted with brackish-water RO membranes (Osmonics, AG4040TF) that were supplied with the pilot system. The specific water flux as a function of time is illustrated in Fig. 6 for the tertiary effluent feed stream. The membranes experienced greater than 30% flux decline after 24 h of operation, indicating that they were severely fouled during this time period; evidently due to the rather high TSS in the filter effluent water (Table 1). This result was not unexpected because SDI testing (AMI 2007) had revealed that the 15 min SDI of the effluent water was too high to measure; however, this basic test of flux decline was still performed in order to assess the effects and reversibility of fouling from untreated effluent on the pressure-driven membrane processes. Essentially, the AG4040TF membranes were sacrificed in order to clearly prove to stakeholders that additional treatment of the effluent was necessary prior to RO or NF treatment. Because membrane manufacturers commonly recommend cleaning when the flux declines below 15–20% of the initial flux, cleaning would have to be performed more frequently than once a day.

Previous studies (Leslie et al. 1996; Chakravorty and Layson 1997; Ebrahim et al. 1997; Ghayeni et al. 1998; Duin et al. 2000; Brehant et al. 2002; Bourgeous and Narayanan 2004) have demonstrated that microporous membrane filtration processes (e.g., MF or UF) can efficiently pretreat effluent water prior to treatment with RO or NF. Run times can be extended and the frequency of membrane cleaning can be reduced. Thus, a pilot-scale UF system was tested as a pretreatment process for the RO/NF pilot system. The specific water flux through the brackish-water RO membranes was measured over time for two different feed streams, untreated feedwater, and coagulation-enhanced UF-treated feedwater. Results in Fig. 7 indicate that flux decline during operation without pretreatment was on average 40%/day, whereas flux decline with pretreatment was approximately 3%/day. TSS and TOC analyses of the UF feed and permeate revealed that suspended solids were completely removed and TOC was reduced by approximately 11% (at 30 mg/L alum dose) during the coagulation-enhanced UF process.

The operating conditions of the UF pilot system, including transmembrane pressure, feed flowrate, and backwash frequency, were optimized to achieve the required steady supply of feed-

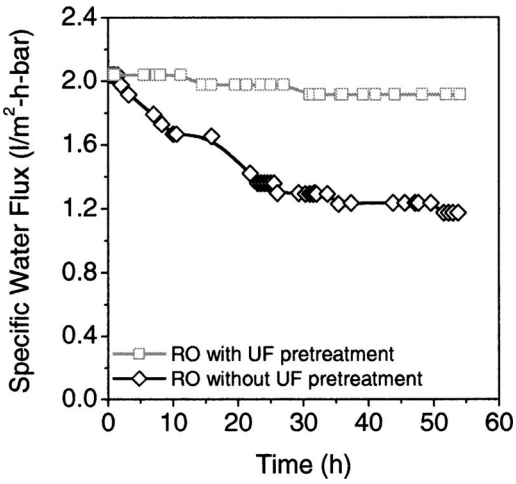

Fig. 7. Specific water flux as function of time for AG4040TF membrane using untreated feedwater and UF-pretreated feedwater

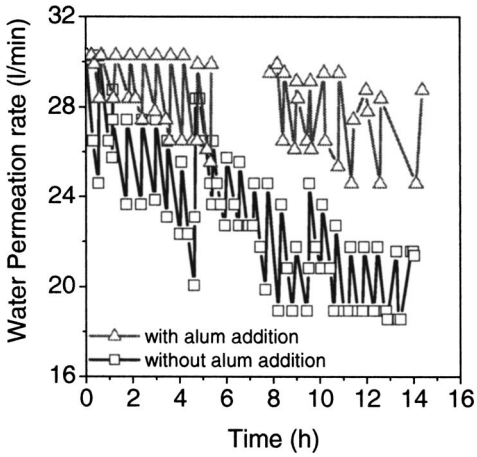

Fig. 8. Water permeation rate as function of time for UF pretreatment of tertiary effluent with and without alum addition. Alum dose was 2 mg/L and operating pressure was 0.7 bar. Backwashing was performed with RO product water every 30 min for 30 s.

water to the RO/NF system and to minimize cleaning frequency. During the tests, the UF system was operated at a transmembrane pressure of approximately 0.7 bars for 30 min intervals separated by 30 s backwash cycles. Subsequently, it was concluded that the UF membrane operated most efficiently with an alum dose of 2 ppm (Fig. 8). Flux decline was reduced when alum was added to the feedwater and the frequency of cleaning cycles was reduced to once every 2 days.

Pilot-Scale Studies

TDS Removal Using RO and NF Processes

Pilot-scale experiments were conducted at the treatment plant for each of the three RO/NF membranes selected using a three-stage high-pressure membrane skid. The feedwater was dechlorinated effluent water at 23–26°C that was pretreated with alum-enhanced UF. The main differences between the pilot- and bench-scale testing were the size of the membrane being tested and the fact that bench-scale tests were conducted as a batch process, whereas pilot-scale experiments were conducted in a continuous flow setup.

Specific water flux as a function of time and transmembrane pressures is illustrated in Fig. 9. All of the membranes experienced decline in specific flux as the transmembrane pressure increased. The higher permeation rates at higher pressures led to enhanced transport of foulants to the membrane, greater compaction of the fouling layer, and increased hydrodynamic resistance (Bruggen and Vandecasteele 2001).

Flux decline during the pilot tests occurred more slowly than flux decline during bench-scale experiments, likely due to the coagulation-enhanced UF performed on the pilot-scale feedwater. Results from the pilot tests (Fig. 9) corresponded well with results from the bench-scale experiments (Fig. 4). Membranes A and B, the NF membranes, experienced gradual flux decline, and Membrane H, the RO membrane, produced a more constant flux. Membrane A had the highest initial flux and fastest flux decline at both bench and pilot scale; however, in the pilot tests the initial sharp flux declines for Membranes A and B were not apparent. It is likely that the large surface area and improved hydrodynamics in the spiral wound elements prevented rapid fouling at the early stages of the pilot tests (Geraldes et al. 2004). Membrane H, the

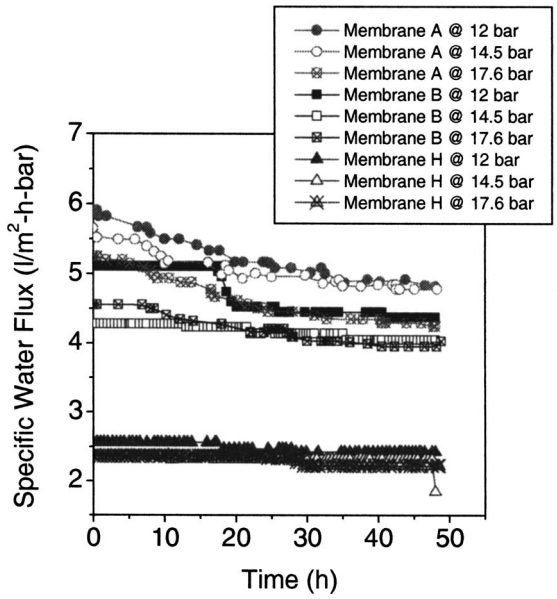

Fig. 9. Specific water flux as function of time and transmembrane pressure for RO and NF membranes tested at pilot scale

RO membrane, had the lowest flux decline during both bench- and pilot-scale testing; this is most likely due to its lower flux, and therefore, lower drag forces toward the membrane (Geraldes et al. 2004). Higher temperature of the feedwater during pilot testing may have contributed to the slightly higher flux of all of the membranes at pilot scale.

TDS rejection as a function of time and transmembrane pressure is shown in Fig. 10. Membrane H provided the highest TDS rejection, as expected for a tight RO membrane, and membrane A, the looser NF membrane, provided the lowest TDS rejection. TDS rejections for both Membranes B and H were generally greater than 99%. Membrane B provided less than 99% TDS rejection at the beginning of the lowest pressure experiment but rejection increased to greater than 99% as time progressed. Rejection in the higher pressure experiments was greater than 99% at all times. This difference in rejection at different transmembrane pressures can be explained by two possible mechanisms. First, when the new Membrane B was introduced to treated wastewater during the low transmembrane pressure experiments,

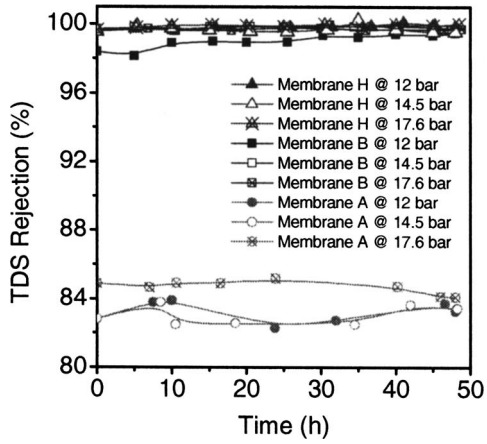

Fig. 10. TDS rejection as function of time and transmembrane pressure for three RO/NF membranes tested at pilot scale

Table 3. Average TOC and Nutrient Removals from TMWRF's Effluent Water during Pilot-Scale Testing of Three RO/NF Membranes

	Average removal (%)		
Membrane	TOC	Nitrogen	Phosphorous
A (NF)	62±5	85±3	85±14
B (NF)	67±1	87±13	91±4
H (RO)	75±1	99±2	82±6

Table 4. Modeled Energy Consumption for Three Membranes Tested at Pilot Scale

Membrane	Average specific power calculated using experimental conditions and results (kWh/m^3)	Average specific power calculated using design programs (kWh/m^3)	Permeate concentration calculated using design programs (ppm TDS)
A (NF)	1.09	0.62	35.6
B (NF)	1.43	0.48	37.6
H (RO)	2.54	0.92	11.6

rapid pore blockage likely occurred over the first 24 h. This would have tightened the membrane and caused increased rejection of smaller contaminants over time. Membrane cleaning between experiments may not have achieved complete flux restoration, and therefore increased rejection would be expected in subsequent experiments. Second, flux was higher at higher transmembrane pressures and dilution effects resulted in increased apparent rejection.

TOC and Nutrient Removal

An additional benefit of implementing RO or NF treatment processes at wastewater treatment plants, is that TOC and nutrients (e.g., nitrogen and phosphorous) are further removed from the effluent. While nitrogen and phosphorous have discharge limits, TOC discharge is not regulated; however, its removal can protect downstream ecosystems. The average rejections of TOC, nitrogen, and phosphorous are summarized in Table 3 for each membrane tested at pilot scale. The average concentrations of TOC, total nitrogen, and total phosphorous in the feed to the membrane system were 7, 2, and 0.3 ppm, respectively. Membrane H, the RO membrane, had the highest rejection of nitrogen and TOC but showed low rejection of phosphorous. Membrane B, the tighter NF membrane, showed higher rejection of these three constituents than Membrane A, the loose NF membrane.

Implementation of RO or NF processes at TMWRF could substantially reduced TDS, TOC, and nutrient discharges. As indicated earlier, approximately 11,500 kg/day of TDS will have to be removed from the effluent after expansion of the treatment plant. To achieve this goal, approximately 30,283 m^3/day (8 mgd) of water will need to be produced with RO or NF and then mixed with the remaining effluent water. Using TMWRF's current TOC, nitrogen, and phosphorous effluent concentrations (7, 2, and 0.3 ppm, respectively) and rejection data from Table 3, the daily amount of TOC and nutrient removal can be calculated. It was determined that an average of 136 kg TOC, 41 kg nitrogen, and 8 kg phosphorous can be removed daily. The TMDLs for nitrogen and phosphorous are 227 and 61 kg/day, respectively, and thus at least 18% of the TMDL for nitrogen and 13% of the TMDL for phosphorous can be removed.

Theoretical Energy Consumption

Performance results from pilot-scale testing were used to calculate the theoretical energy consumption of each membrane tested. Theoretical power consumption was normalized to the permeate flow rate and was reported in terms of specific energy consumption (W h/m^3 water produced).

Average theoretical specific energy consumptions for each of the membranes tested during the pilot-scale investigations are summarized in Table 4. Specific power (SP) was calculated using Eq. (2)

$$\mathrm{SP} = \frac{1}{\text{efficiency}}\frac{\Delta P}{R} = \frac{1}{\text{efficiency}}\frac{\Delta P Q_f}{Q_p} \tag{2}$$

where ΔP=transmembrane pressure and R=water recovery expressed as the ratio between the feed flowrate (Q_f) through the pump and the permeate flowrate (Q_p). The reported efficiency of the pilot test unit is 55%. Under the specific conditions in which the experiments were conducted, the recovery was kept constant, and therefore specific energy was mostly a function of pressure and recycling rate.

Also provided in Table 4 are values of specific power consumption and permeate TDS concentrations calculated using the design programs available from the manufacturers of the membrane tested in the current study. WinFlows (Osmonics 2005) was used for Membrane A, ROSA (Filmtec 2005) for Membrane B, and IMSDesign (Hydranautics 2005) for Membrane H and the UF membrane. For the RO and NF membranes, the data were calculated using three membrane elements similar to those tested in the current study at pilot scale, at the manufacturer tested pressure, and at 63% system recovery [including concentrate recycling of 11.3 L/min (3 gpm)]. For the UF system, the power consumption was calculated based on the feed flowrate required for the RO/NF pilot system (approximately 25 L/min) and was found to be approximately 0.3 kW h/m^3 for either the RO or NF membranes (value not included in Table 4). For both NF membranes, the design programs predicted 40–66% lower specific power consumption and for the RO membrane the design program predicted almost 65% lower specific power consumption than that predicted using experimental conditions and results. As expected, the specific power consumptions for Membranes A and B, the NF membranes, were lower than that of Membrane H, the RO membrane. Although the theoretical power consumption results indicate that NF is favorable to RO, it is important to note that high rejection (in addition to high flux) is desirable in order to reduce the volume of water that requires treatment and in order to reduce capital investment and membrane replacement costs. Based on results from Table 4, approximately 8% (by volume) more water needs to be treated if NF is used instead of low-pressure RO. Thus, in order to more comprehensively compare NF and RO processes for this application, an optimization program that considers more than just energy consumption would need to be used.

Preliminary Cost Estimate

Preliminary estimates for capital and operating costs were made using a model previously published by Sethi and Wiesner (2000). The model considers separate correlations for the major categories of capital costs including membranes, pipes and valves, electrical and instrumentation, tanks and frames, and miscellaneous items (e.g., buildings and power supply). The model is a series of equations that accounts for each of the categories and is based on

Table 5. Design Criteria and Operating Parameters Used in Estimating Cost of Membrane Treatment Plant for Treatment of Tertiary Effluent at TMWRF[a]

Parameter	Spiral wound nanofiltration	Hollow fiber ultrafiltration
Plant design life (years)	20	20
Membrane molecular weight cut off (Da)	250	100,000
Membrane resistance (L/cm)	4.9×10^{11}	1.6×10^{10}
Membrane diameter (mm)	1.52	0.92
Membrane length (mm)	1,000	1,200
Membrane life (years)	5	5
Feed pressure (kPa)	800	150
Recovery (%)	85	Variable
Energy cost ($/kWh)	0.07	0.07
Cost of capital (%)	10	10
Earning interest rate (%)	8	8
Efficiency of pumps (%)	70	70
Temperature (°C)	20	20

[a]Full table and design considerations shown in Sethi and Wiesner (2000).

the active membrane area needed for treatment. Using the feedwater characteristics of TMWRF's effluent, a multistage system was developed and ROSA and IMFDesign were used to obtain an estimate of the membrane area needed to produce 30,283 m^3/day of desalted water. Results from ROSA suggested that approximately 37,500 m^2 of spiral wound NF membrane area will be needed for desalination of the desired volume at 90% recovery, and results from IMFDesign suggested that approximately 130 m^2 of hollow fiber UF membrane area will be needed for pretreatment. Cost estimates for spiral wound NF membranes, hollow fiber UF membranes, membrane design, and operating parameters are summarized in Table 5. Cost formulas for nonmembrane equipment and facilities are expressed as power law expressions correlating the equipment costs to the membrane area requirement, with the generic form

$$\mathrm{COST} = k(A_{\mathrm{mem}})^n \tag{3}$$

where k=constant representing the relative weight of a cost component; A_{mem}=membrane area required to produce the required capacity; and n=exponent representing the economy of scale associated with a cost component (Sethi and Wiesner 2000). The correlations for equipment are

$$C_{\mathrm{Pipes/Valves}} = 5{,}926.13(A_{\mathrm{mem}})^{0.42} \tag{4}$$

$$C_{\mathrm{Instruments/Controls}} = 1{,}445.50(A_{\mathrm{mem}})^{0.66} \tag{5}$$

$$C_{\mathrm{Tanks/Frames}} = 3{,}047.21(A_{\mathrm{mem}})^{0.53} \tag{6}$$

$$C_{\mathrm{Miscellaneous}} = 7{,}865.02(A_{\mathrm{mem}})^{0.57} \tag{7}$$

Using the model and its assumptions, a preliminary estimate of the capital costs for the implementation of a 30,283 m^3/day (8 mgd) UF/NF system at TMWRF is approximately $4.3 million for capital equipment and $3.2 million for miscellaneous works (e.g., new buildings, electrical supply, and other infrastructure). Operational and maintenance costs were estimated to be approximately $0.23/$m^3$ ($0.88/1,000 gal) of water produced. This amounts to approximately $2.5 million/year.

Conclusions

Based on membrane performance during pilot-scale experiments, membrane processes can be implemented at TMWRF for TDS and nutrient removal from the effluent water. The NF membranes tested were shown to be effective in rejecting TDS, TOC, and nutrients while maintaining higher fluxes and lower specific power consumptions than the RO membranes tested. It was shown that the tertiary treated effluent from TMWRF is not an acceptable feed source to the membrane desalination processes; both the RO and NF membranes fouled rapidly when directly exposed to TMWRF's effluent. Coagulation-enhanced UF pretreatment was very effective in conditioning the feedwater for the RO and NF processes. Long-term pilot experiments should be conducted to study the longer-term fouling effects on the performance and economy of the process.

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