



## Membrane treatment of side-stream cooling tower water for reduction of water usage

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### ABSTRACT

A pilot study was conducted to determine whether membrane treatment on a side stream of recirculating cooling-tower water could reduce overall water usage and discharge. The treated permeate was returned to the cooling tower while the concentrate was discharged to the sanitary sewer. Flow rates, pressures and water chemistry were monitored. The pilot demonstrated potential substantial water savings. Maximum make-up water and discharge reduction were 16% and 49%, respectively. As high as possible permeate recovery is needed to maximize water conservation. Silica scaling on the membranes limited water savings in this pilot. Development of membranes with a solute-rejection capacity less than the 92% average of the membranes used in the pilot would assist in optimizing water savings. Decreased water outlays compensated for the additional energy used by membrane treatment. Scaling control is critical for economic operation.

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### 1. Introduction

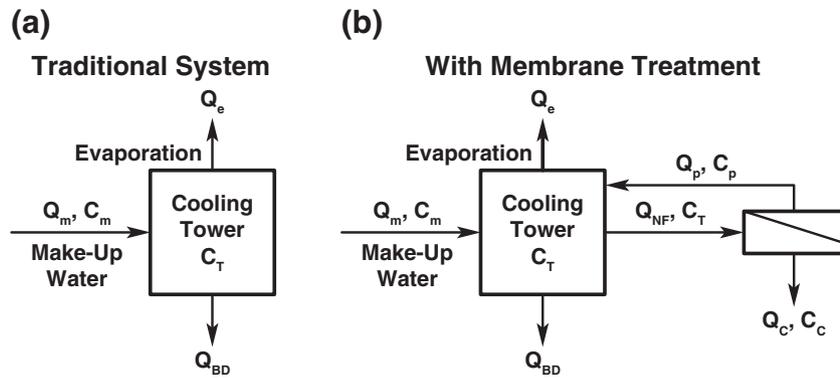
Large volumes of water are needed to support thermoelectric power generation. The majority of this water is used for cooling. Thermoelectric generation was responsible for 49% of freshwater withdrawals in the United States in 2005 [1] and 3% of freshwater consumption [2], where water consumption represents the amount of water withdrawal that is not returned to the source. This translates to approximately 10 million m<sup>3</sup> per day (3 billion gallons per day) freshwater consumption by thermoelectric power generation [3]. As freshwater supplies become scarcer there will be increased competition and cost for freshwater usage. In addition, there can be costs associated with water disposal. It is to the power generation industries' benefit to develop methods to minimize water consumption and maximize water usage efficiency. Maximizing water usage efficiency may allow power plants to expand operations while minimizing new water supply and water disposal costs.

Thermoelectric power plants will benefit from optimization of water usage by cooling water towers. Recirculating (closed loop) wet cooling towers are used in 41.9% of existing thermoelectric power plants [3]. The recirculating cooling tower first withdraws water from a freshwater source and recirculates this water through the cooling tower.

Salt concentrations increase as water is recirculated and the cooling water evaporates until dissolved salt concentrations reach a threshold value (the set-point). To keep the concentrations of dissolved salt at or below the threshold value, the concentrated water is discharged as blowdown and water from the fresh-water source (make-up water) is added to the tower. Schematically, the fluxes of water and salt in the system are represented by Fig. 1a. Blowdown and make-up water are required to control the higher salt loads that can lead to scale formation of, for example, calcite and silica.

Side-stream treatment can be applied to mitigate scale formation and reduce water usage. Examples of in-use side-stream treatment include lime-soda softening [4] and brine concentrators [5]. Lime-soda softening precipitates calcium, magnesium (as calcium carbonate and magnesium hydroxide floc), and silica (by adsorption onto the magnesium hydroxide floc) by adding lime (CaO and/or Ca(OH)<sub>2</sub>) and soda ash (Na<sub>2</sub>CO<sub>3</sub>) to the blowdown. The treated water is clarified, filtered, pH adjusted, and added as make-up. Sodium, potassium, sulfate and chloride are not removed in this process. Matson and Harris [4] give two examples of industrial plants achieving zero liquid discharge by lime-soda softening. They report the major problems facing side-stream softening are drift losses and sludge disposal. DiFilippo [5] describes the San Juan Generating Station where blowdown is treated by brine concentrators. The brine concentrator is reported to recover over 98% of the inflow. Thermal evaporative systems have the advantage of creating

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**Fig. 1.** Conceptual model of the flow and mass balances for a cooling-tower water system without (a) and with (b) nanofiltration system.  $Q$  = flow,  $C$  = concentration,  $m$  = make-up water,  $BD$  = blowdown, subscript  $C$  = concentrate,  $NF$  = NF feed,  $p$  = permeate,  $e$  = evaporation,  $t$  = cooling tower.

high quality distillate and low volumes of discharge. However, they are large power consumers (26–83 kWh/m<sup>3</sup>[6]). Membrane systems consume less energy than thermal distillation systems (2–20 kWh/m<sup>3</sup>[6]) and can still produce relatively pure water. Jones [7] showed that using nanofiltration combined with ionization can lower water discharge by 80% and eliminate the need for antiscalent chemicals.

Membrane technologies are widely used for desalination and are described in [8]. Pressure driven membranes include microfiltration (MF), ultrafiltration (UF), nanofiltration (NF), and reverse osmosis (RO). These membranes are listed in order from lower to higher of removal capacity and energy requirements, with RO being the most energy intensive, but able to remove monovalent ions. In general, NF reject multi-valent ions well. Both RO and NF can be thin film composite (TFC) membranes that consist of a porous substrate coated with a dense thin polymer film such as a polyamide. They are considered non-porous in that they do not have a well defined pore structure and can exclude low molar mass species. In reality, there is overlap between how the membranes perform and the specifications of individual membranes need to be carefully examined when choosing the proper membrane for a specific use. Typically, 50% to 90% and 35% to 60% recovery can be expected for brackish water and seawater, respectively [8].

We tested the use of membrane treatment to decrease both the amount of make-up water and blowdown needed to run a cooling tower. As the cooling-tower water is recirculated, a side stream of the water flows through a membrane system to remove some of the dissolved salts. The permeate is returned to the cooling tower and the concentrate is discharged. Addition of permeate to the cooling tower allows the volume of blowdown to decrease because the dissolved salt concentration should not increase as rapidly or reach an equilibrium below the set-point. The schematic of the system is presented in Fig. 1b. Predictive calculations [9] indicated that ultrapure water (>90% chemical rejection) was not necessary for optimal water savings in this application. Therefore, membrane selection focused on minimizing energy use with sufficient chemical rejection to dilute the cooling tower water.

This paper describes a pilot study that was run to test whether the membrane treatment of cooling-tower water can lead to overall water savings. The pilot was run for 3 months in the summer and fall of 2010 on a cooling tower for a building containing office and laboratory space. The data collected from the pilot are assessed to determine membrane performance and measured and potential water savings. In addition, an economic analysis of the energy required to run the membrane system is presented.

## 2. Materials and methods

### 2.1. Equipment descriptions

The pilot study was conducted on the cooling tower (Baltimore Aircoil Company (BAC) chiller, model No. 3501A-X) at the Center for

Integrated Nanotechnology (CINT) at Sandia National Laboratories. This tower is used to cool a building with both laboratory and office space. The cooling tower can produce 300 tons of refrigeration and operates at a flow rate of 3.4 m<sup>3</sup>/min (900 gpm). The tower has a groundwater source provided by the Kirtland Air Force Base. Cooling tower water is recirculated from a below-ground sump in the vicinity of the cooling tower. Sump water quality is monitored and scale and corrosion problems are mitigated by maintaining a conductivity below approximately 1200  $\mu$ S/cm. The cooling tower water is normally treated with a scale and corrosion inhibitor (Industrial Water Engineering, Inc., product number CW-8590) and a bromide-based biocide (Safetybrom 6300). Scaling of calcium and silica were controlled by phosphonate (to control the formation of precipitates) and a terpolymer dispersant (to prevent settling and clumping). During the time of the pilot, the oxidizing biocide was replaced with Aquacar 515, a glutaraldehyde-based biocide produced by Dow Chemical. This replacement was done to minimize damage to the polyamide membranes.

A schematic of the pilot design is presented in Fig. 2. A portable utility pump pulled water from the sump and maintained a constant flow into the system, which was located in the pilot trailer. Once in the trailer, the water circulated first through two media filters: one designed to remove particulates, the second, an activated carbon filter for removing residual chlorine and some organic contaminants. From the media filters the water circulated through a 51 cm (20 in.), 50  $\mu$ m cartridge filter and then a 50.8 cm (20 in.), 1  $\mu$ m cartridge filter to remove particulates. Between the cartridge filters and the membrane system a chemical injection pump was used for the addition of an antiscalant/dispersant (SpectraGuard™). The product sheet for SpectraGuard™ states that it controls inorganic scale, metals and colloids in membrane feedwater, including silica, calcium sulfate, calcium phosphate, and calcium carbonate. A high-pressure pump (EBARA Fluid Handling model EVMU3-18F0500T36) with 5 horse-power motor pressurized the water as it entered the elements. The membrane system consisted of two identical parallel six-element arrays. Each array contained six 6.4 cm (2.5 in.) elements in series with each element housed in a single element 316 stainless steel pressure vessel. The elements used were KOCH TFC® softening membranes. These elements were chosen because they are a low-pressure element with a reported >90% chloride and 98.5% total hardness rejection. As mentioned in the Introduction, the aim was to select a membrane that would minimize energy use but still reject sufficient dissolved salts. Permeate flow was returned to the cooling tower sump and concentrate flow discharged to the sanitary sewer system.

Flow rates, pressure, temperature, and conductivity were monitored through analog meters and inline sensors with a programmable logic controller. Pressure was monitored at the feed and concentrate ends of the membrane arrays and halfway through the two arrays (interbank 1 and 2). Flow from the cooling tower sump ( $Q_T$ ), recycle flow ( $Q_R$ ) combined feed ( $Q_T + Q_R$ ), permeate ( $Q_p$ ) and concentrate flow ( $Q_c$ ) were all monitored. In addition, conductivity was measured on the

## Schematic of Pilot

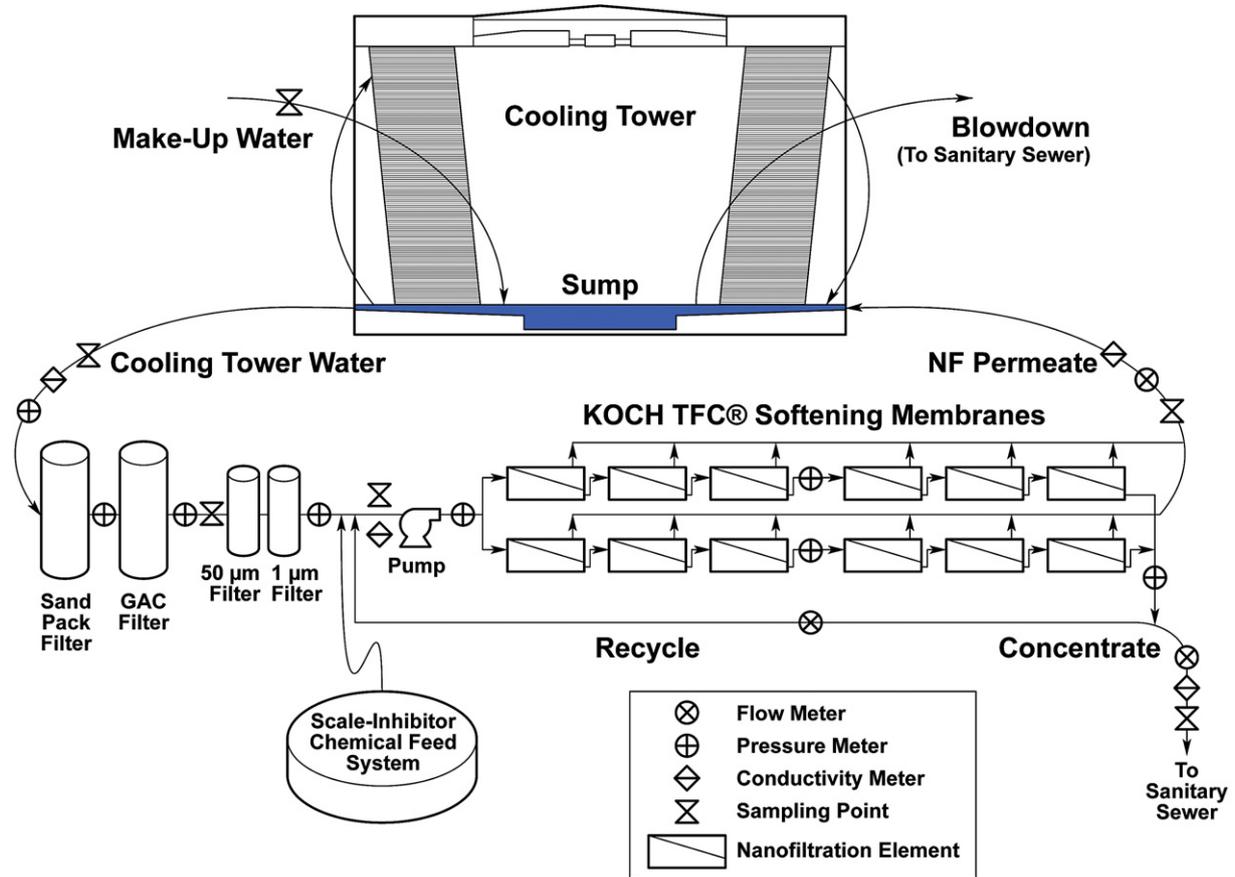


Fig. 2. Nanofiltration pilot flow diagram.

tower water ( $C_T$ ), combined feed, permeate ( $C_P$ ) and concentrate ( $C_C$ ) flow streams.

### 2.2. Pilot operations

The pilot ran from August 4th through October 29th, 2010. The system was considered to be running stably by August 11th, 2010. Water samples were collected for chemical analyses periodically throughout the pilot. Sampling was frequent (3 times a day) at the beginning of operations and decreased to 1 or 2 times a week once the system stabilized.

Membranes were cleaned several times during the pilot. Initially, a base washing was performed using a solution of sodium hydroxide and Alconox to attain a pH between 10.5 and 11.0. The solution was circulated through the system for approximately 30 min. An approximately pH 2 HCl acid solution wash followed the base wash. The pH of the acid solution was periodically checked to monitor the neutralization of the acid by the scale on the filters. When the solution reached a pH between 5 and 6, additional HCl was added to reduce the pH back to 2. When the acid solution no longer neutralized, the clean-in-place was deemed to have reached its maximum level of effectiveness and was terminated.

After completion of the pilot, the membrane surface and feed spacer were examined both visually and with Scanning Electron Microscopy-Energy Dispersive X-ray Spectroscopy (SEM-EDS) to determine possible scalants.

### 2.3. Chemical analyses

Major cations (calcium, magnesium, potassium and sodium) were analyzed by a Teledyne Leeman Labs inductively coupled plasma mass

spectrometer (ICP-MS). Major anions (bromide, chloride, fluoride, nitrate, orthophosphate, and sulfate) were analyzed by a Dionex ICS2500 ion chromatograph (IC) with a Dionex RFIC IonPac AS18 column. Specific conductivity and pH were measured using a Hach Sension 5 meter and Hach Sension 1 pH meter, respectively. A Hach 2100P turbidimeter was used to measure turbidity. Dissolved silica ( $\text{SiO}_2$ ) was measured using the Hach DR2400 colorimeter using the silicomolybdate method (Hach Method 8185). This method does not detect polymerized silica. Samples were stored and measurements made at room temperature.

Scanning electron microscope (SEM) images of the feed spacer and membrane surface were attained using a Zeiss Supra 55VP. Energy Dispersive Spectroscopy (EDS) provided relative proportions of inorganic elements in the deposits.

### 2.4. Data analysis

Permeate recovery ( $R$ ), the fraction of water passed through the membrane as permeate, was calculated as follows:

$$R = 1 - \frac{Q_C}{Q_T} \quad (3)$$

Where  $Q$  refers to flow rates and the subscripts  $C$  and  $T$  refer to the concentrate and cooling tower flow streams. The flow from the cooling tower was used instead of the combined feed flow into the membrane elements in order to calculate overall recovery.

**Table 1**  
Flow and mass-balance equations.

System	Flow balance	Mass balance
Traditional cooling tower (no membrane system)	$Q_m = Q_e + Q_{BD}$	$Q_m C_m = Q_{BD} C_T$
Cooling tower with membrane system and blowdown	$Q_m = Q_e + Q_{BD} + Q_C$	$Q_m C_m = Q_{BD} C_T + Q_C C_C$
Cooling tower with membrane system and no blowdown	$Q_m = Q_e + Q_C$	$Q_m C_m = Q_C C_C$

Solute rejection ( $w$ ) of the membranes was calculated as:

$$w = \frac{C_C(C_T - C_P)}{C_T(C_C - C_P)} \quad (4)$$

where  $C$  refers to the solute concentration and subscripts  $C$ ,  $T$ , and  $P$  refer to the concentrate, cooling tower, and permeate concentrations, respectively. Eq. (4) was derived from Eqs. (5) and (6) below, defining the mass transfer through the membranes:

$$C_C = \frac{w C_T}{(1-R)} \quad (5)$$

$$C_P = \frac{(1-w)C_T}{R} \quad (6)$$

The specific conductivity measured using a conductivity probe (as opposed to the in-line meters) were used to calculate the bulk  $w$  (as opposed to  $w$  for specific solutes).

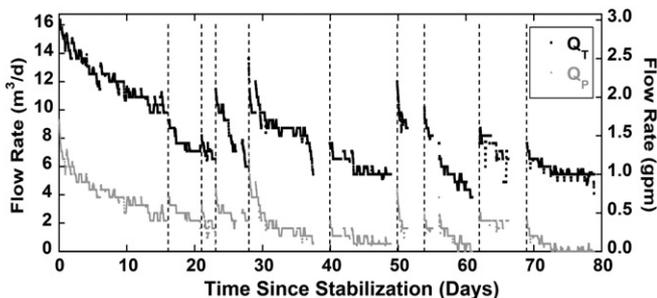
Flow and mass-balance equations were used to calculate theoretical water savings (Table 1). Cooling-tower and pilot flow-rate data were used for  $Q_M$  and  $Q_C$  values, respectively. Drift was assumed to be negligible.  $C_T$  was set to 1200  $\mu\text{S}/\text{cm}$ , the set-point for the cooling tower water. Measurements made from the grab samples during the pilot were used for  $C_C$ . A constant value of 400  $\mu\text{S}/\text{cm}$  was used for  $C_M$ , a rounded value of the average of periodic measurements of 375  $\mu\text{S}/\text{cm}$  recorded from the make-up water samples. Evaporation rate ( $Q_e$ ) and blowdown flow rate ( $Q_{BD}$ ) were calculated from the mass and flow-balance equations.

Cycles of concentration (CoC) for a cooling tower without membrane treatment was calculated as:

$$\text{CoC} = \frac{Q_M}{Q_{BD}} \quad (7)$$

and, for the system with membrane treatment:

$$\text{CoC} = \frac{Q_M}{Q_{BD} + Q_C} \quad (8)$$



**Fig. 3.** Flow rates from the cooling tower sump into the membrane system ( $Q_T$ ) and from the permeate flow stream ( $Q_P$ ) over the duration of the pilot. Dashed vertical lines indicate times of membrane cleanings.

**Table 2**  
Summary of flow data ( $\text{m}^3/\text{d}$  (gpm)).

	Cooling tower	Recycle	Feed	Permeate	Concentrate
Average flow	8.3 (1.5)	19.2 (3.5)	27.5 (5.1)	2.1 (0.4)	6.2 (1.1)
Standard deviation	2.8 (0.5)	6.7 (1.2)	8.0 (1.5)	1.6 (0.3)	1.4 (0.3)
Minimum	3.8 (0.7)	8.7 (1.6)	12.5 (2.3)	0.0 (0.0)	2.7 (0.5)
Maximum	18.5 (3.4)	36.0 (6.6)	44.2 (8.1)	9.3 (1.7)	9.7 (1.8)

Water savings were assessed by comparing  $Q_M$  and  $C_T$  data measured during a month the pilot was running (August 2010) to those data collected during the same month of a previous year (August 2008). A comparison was also made for July 2010 (just prior to the pilot running) to July 2008 and 2009 as a control. Average and maximum temperatures for those months were also compared, as the higher the temperature, the higher the expected  $Q_M$ .

An analysis was performed to assess the economic feasibility of using membranes to treat cooling tower water. This was done by calculating the necessary savings in clean water and disposal cost to pay for the electrical costs of running the membrane system. The cost of make-up water was assumed to range from \$0.79 to \$3.24 per  $\text{m}^3$  (\$3.00–\$12.25 per 1,000 gallons) based on the data presented in Walton [10] and Hightower and Pierce [11]. It was assumed that the cost of water disposal was the same as the cost of new water (based on discussions with people working in Sandia's facilities group). Energy usage was calculated assuming that 50% of pump capacity was being used for the pump in our system (1.87 kWh). Average  $Q_P$  as a function of  $R$  was extracted from the pilot data and used to calculate the energy cost per volume of permeate produced as a function of membrane efficiency. Calculations were made using electricity costs of \$0.10, \$0.20, \$0.30, and \$0.40 per kWh.

### 3. Results

There was clear evidence of membrane scaling or fouling as can be seen by decreases in permeate flow rates (Fig. 3). Permeate flux increased again after membrane cleanings, though subsequently decreased quickly. Membrane performance improved less and less with subsequent cleanings. Decreases in permeate flux led to decreasing flow rates from the cooling tower. The permeate recovery ( $R$ ) ranged from 0 to 64% with an average of 23%. By the end of the pilot there was virtually no passage of water through the membranes.

Flow rate from the cooling tower ranged from 3.8 to 18.5  $\text{m}^3/\text{d}$  (0.7–3.4 gpm) over the duration of the plot (Table 2 and Fig. 3). Permeate and concentrate flow rates are both positively correlated with the cooling tower flow rate. Recycle flow rate is negatively correlated with  $Q_T$  maintaining a relatively constant feed flow into the membrane system.

Pressure remained relatively stable except when the capacity of the pump was adjusted (Table 3). The interbank pressure was monitored to determine whether pressure buildup may be due to scaling (in which case the pressure would be expected to increase downstream) or biofouling (in which case the pressure would be expected to increase upstream). The pressure drop was greater between the feed and interbank than between the interbank and the discharge. However, the differences were not great enough to provide clear insight on where in the element arrays the membranes were scaling or biofouling.

**Table 3**  
Summary of pressure data (bar (psi)).

	Feed	Interbank 1	Interbank 2	Concentrate
Average pressure	8.6 (124)	7.1 (103)	7.4 (107)	6.6 (96)
Standard deviation	1.7 (25)	1.9 (27)	1.8 (26)	2.3 (34)
Minimum	6.1 (88)	4.6 (66)	4.8 (70)	3.7 (54)
Maximum	11.8 (171)	10.8 (157)	11.0 (159)	11.3 (164)

**Table 4**

Summary of water chemistry data (mg/L) (average values with standard deviations in parentheses).

Constituent	Make-up	Cooling tower	Post-GAC	Feed	Permeate	Concentrate	w (%)
Calcium	37 (2.2)	115 (20)	115 (20)	148 (19)	10 (5.2)	172 (19)	97
Magnesium	6.6 (0.4)	22 (1.0)	22 (0.9)	28 (1.0)	2.0 (0.7)	33 (4.2)	97
Potassium	3.6 (0.3)	20 (6)	20 (5.7)	22 (6.0)	5.7 (1.8)	26 (6.7)	97
Sodium	27 (0.9)	103 (9.1)	104 (11)	126 (14)	32 (6.4)	139 (16)	89
Chloride	22 (2.6)	90 (11)	89 (7.0)	105 (11)	30 (6.1)	118 (19)	89
Fluoride	0.6 (0.03)	2.0 (0.2)	2.0 (0.1)	2.5 (0.2)	0.5 (0.1)	2.9 (0.4)	92
Sulfate	36 (2.9)	121 (7.2)	123 (7.0)	157 (15)	5.2 (1.1)	179 (24)	99
Bicarbonate	113 (7.6)	401 (44)	408 (40)	506 (62)	72 (22)	562 (56)	94
Silica	27 (4.8)	116 (12)	116 (18)	136 (19)	38 (8.8)	151 (23)	90

Average solute rejection of the membranes ranged from 89% (sodium and chloride) to 99% sulfate (Table 4). Nanofiltration and reverse osmosis membranes reject species based on size and charge. Therefore, it is not surprising that the membrane was more effective at removing divalent ions than monovalent ions. As atomic radius of potassium is one of the largest monovalent ion, it is also not surprising that the rejection rate was as high as 97%. 90% of the silica was removed. Calculated from the conductivity data, *w* ranged from 77% to 100% with an average of 92%.

It was estimated that the cooling tower discharge (blowdown) could be reduced by as much as 49% with an average calculated decrease of 17% over the duration of the pilot (Fig. 4). Savings in make-up water usage was calculated to be as high as 16%, with an average savings of 6%. Likewise, the CoC can be increased from a baseline of 3 to as much as 4.9, with an average CoC of 3.4 over the duration of the pilot (Fig. 5). Decreases in discharge and make-up water and increases in CoC are all positively correlated with *R*. After membrane cleaning the water savings and CoC increased significantly, but quickly decreased again with the decrease in *R*.

The average make-up flow rate in August 2010 after the start of the pilot was 29% less than that of August 2008—before the pilot (55.3 m<sup>3</sup>/d (14,600 gpd) versus 71.2 m<sup>3</sup>/d (18,800 gpd)). July 2010 *Q<sub>M</sub>* was also less than that for July 2008 and July 2009, but only by 9% and 7% respectively. Because average temperatures for August 2008 and August 2010 were within 3 °C for 25 days of the month, pre- and post-pilot initiation water usage levels are comparable. The days where average temperature differed by more than 3 °C after August 4 were all hotter in 2010 than 2008. Total precipitation in August 2008 and August 2010 were 2.64 and 2.46 cm (1.04 and 0.97 in.), respectively.

In August 2008 the conductivity in the tower water remained close to the set point of 1200 µS/cm (Fig. 6). In contrast, much of the time in August 2010 after commencement of the pilot, the conductivity was below the set-point. The conductivity fluctuated between the set point to below 1000 µS/cm. However, there is not a clear correlation between the conductivity of the tower and other factors such as cleanings of the membrane system and precipitation.

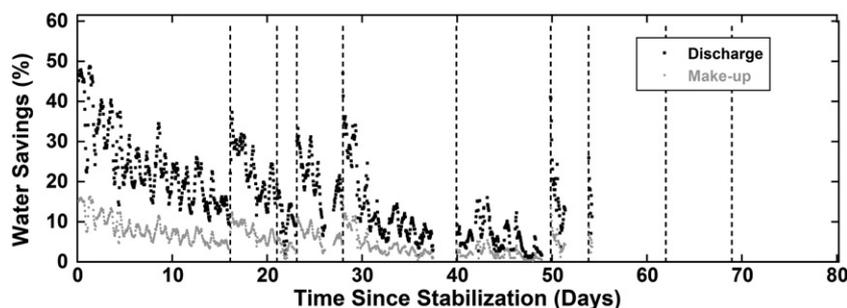
EDS determined that, excluding oxygen, the major constituent on the membrane and spacer surface was silicon at 97%. Calcium

comprised 2% of the constituents with phosphorus, magnesium, and aluminum comprising the remainder. Iron was not detected. SEM images showed visible deposits on the spacer and some mineralization on the downstream membranes (Fig. 7). It is assumed that the deposits are silica based on the EDS results. Neither macroscopic mineralization nor the clearly visible crystals by SEM as presented in [12] and [13] were observed.

#### 4. Discussion

Evidence that side-stream treatment for cooling towers decreases water usage includes: 1) calculated savings in discharge and make-up water usage based on mass and flow balances, 2) calculated increases in CoC, based on mass and flow balances, 3) considerably less measured make-up water usage by the cooling tower in month when the pilot was running compared to the same month in a previous years despite similar temperatures and precipitation conditions, and, 4) the measured conductivity in the cooling tower below the set-point for much of the month of August 2010, when the pilot was running, in contrast to other months.

As the amount of water savings is almost directly tied to the permeate recovery (*R*), scaling must be better controlled for an optimized system. Based on the EDS analyses, silica appears to be the dominant scalant. The cooling tower, feed, and concentrate waters are all supersaturated with respect to amorphous silica and calcite, not surprising given the use of antiscalants. Scale inhibitors are generally more effective on calcium than silica. One study demonstrated that silica scale inhibitors were only able to increase the silica supersaturation ratio from 1.7–1.9 to 2.1–2.3 [14]. Other studies have shown the effectiveness of scale inhibitors on calcium compounds [13,15]. The acid wash should have been effective at cleaning any calcium. But, because silica is more soluble at pH > 10 but relatively insensitive to acid [16], the base wash could be expected to dissolve silica scale while the acid wash should have had less effect. A previous pilot that was run using the same water but without antiscalants pointed to calcium salts as the primary scalant [9]. The difference in results between these two pilots supports the efficacy of the antiscalant on calcium salts.



**Fig. 4.** Calculated percent decrease in discharge and make-up water over the duration of the pilot. Dashed vertical lines indicate times of membrane cleanings. Approximately 54.5 days after stabilization, data from the cooling tower were no longer available, thus calculations could not be made.

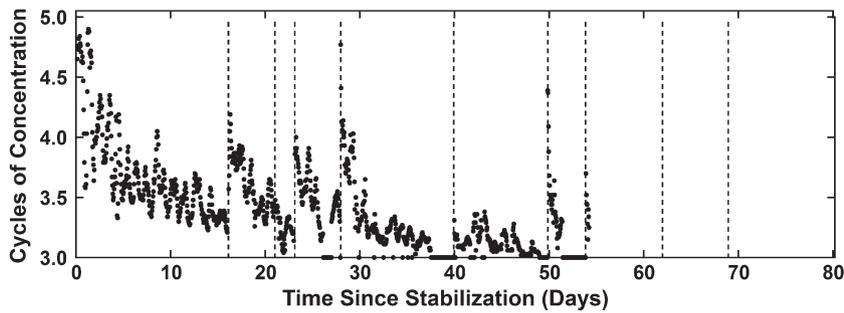


Fig. 5. Calculated cycles of concentration over the duration of the pilot. Note that without the membrane system, CoC would be 3. Dashed vertical lines indicate times of membrane cleanings. Approximately 54.5 days after stabilization, data from the cooling tower were no longer available, thus calculations could not be made.

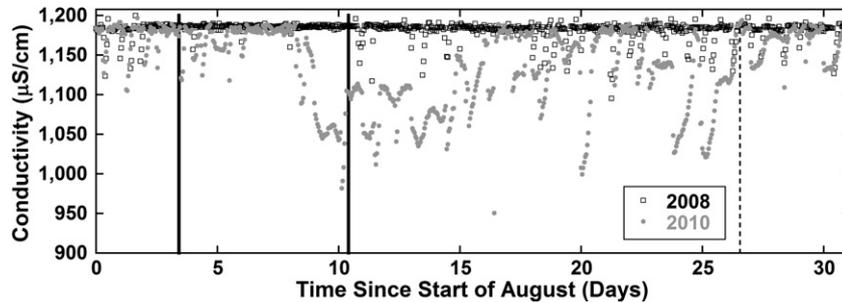


Fig. 6. Comparison of measured conductivity in the cooling tower between August 2008 and August 2010 (the first month of the pilot). Solid vertical lines indication when the pilot started (left) and when the system was considered stable (right). Dashed vertical line indicates time of membrane cleaning.

Membrane-based desalination systems are known to be energy consumers because of the high pressures they require. In general, the higher the salinity of the feed water and the higher the permeate water quality, the more energy needed [8]. Our analyses shows that if the permeate recovery can be maintained high enough (above approximately 0.3), the savings in new water and water disposal costs given the present-day costs should pay for the electrical power at \$0.10/kWh (Fig. 8). At rates of \$0.20/kWh, water savings will pay for the electrical costs if  $R$  is greater than approximately 0.44.

Our analysis is specific to our pilot data. In larger systems, with large pumps, a higher efficiency pump may produce more permeate using less energy. Membrane efficiency may also change as the system is scaled up for a power generation station. Also, the relationship between  $R$  and permeate flow will vary from system to system.

While our economic analysis is promising, a more thorough analysis is needed in order to design a system. For example, seasonal trends in evaporation rates should be accounted for in choosing optimal side-stream flow rates. The analysis needs to take into account water, water disposal, energy and capital costs. The costs of energy and water are intricately entwined [11,17]. Thermal distillation systems have a much greater ratio of treated to waste water (>98% reported by [5]), yet are

also more energy intensive. Membrane systems in general have lower water recovery, but require less energy. The site-specific costs of energy versus water could dictate which system would be best.

Our analyses also indicate that if membranes could be designed with 70–90% chemical rejection rates and higher water recovery, additional water savings could be attained. Many of the commercially available membranes have either chemical rejection rates higher than needed for cooling tower water treatment or much too low (< 40%). Advanced designs using different membrane types and/or staging might allow both scale mitigation and process efficiency needed to optimize the water savings.

## 5. Conclusions

The significant conclusions from this study include:

- Water savings of cooling towers can be attained through the use of side-stream membrane treatment.
- Scaling on the membrane elements needs to be controlled in order for membrane to achieve optimal water savings.

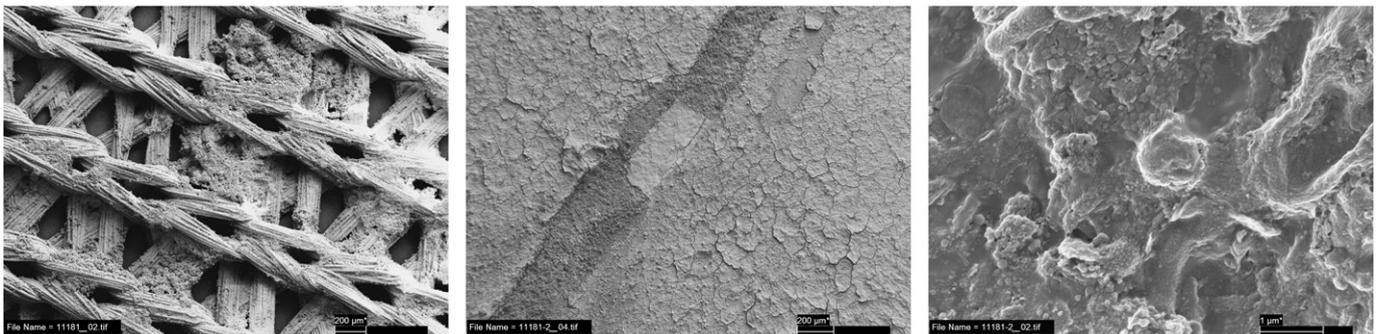
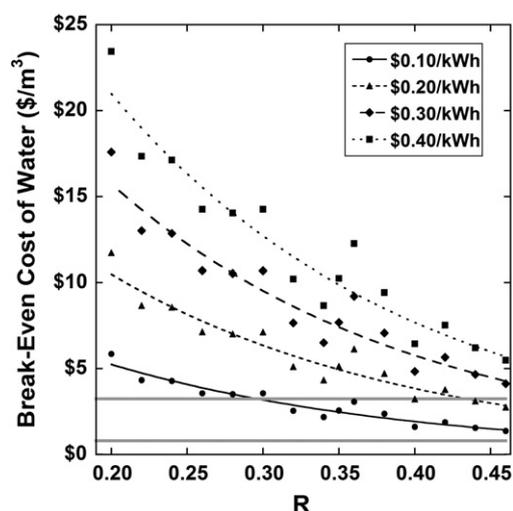


Fig. 7. SEM images of silica scale on the feed spacer left and membrane surface (middle and right).



**Fig. 8.** Relationship between membrane efficiency ( $R$ ) and the necessary cost of water to pay for the energy cost of running a membrane system as a function of the cost of electricity. The horizontal grey lines show the minimum and maximum present cost of water [10,11]. The black lines need to fall below the grey lines for the water savings to offset the additional energy costs.

- The membrane elements removed 89–99% of the dissolved constituents with an average of 92%, higher than is necessary for significant water savings. And,
- It is economically feasible to use the membrane system as long as permeate recovery remains high.

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