University of Nevada, Reno

Evaluation of Pressure-Driven and Novel Membrane Processes for Treatment of Cooling Tower Water

A thesis submitted in partial fulfillment of the requirements for the degree of Masters of Science in Environmental Engineering

by

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May 2009

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We recommend that the thesis
prepared under our supervision by

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entitled

Evaluation Of Pressure-Driven And Novel Membrane Processes For Treatment Of Cooling Tower Water

be accepted in partial fulfillment of the
requirements for the degree of

MASTER OF SCIENCE

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May, 2009

Abstract

Cooling towers are an essential part of power generating facilities and require large quantities of water for continuous operation. To meet the high water demands, cooling towers use makeup water from various sources. Low quality makeup water leads to poor efficiency throughout the plant and large amounts of blowdown. This investigation evaluates pressure-driven membrane processes such as nanofiltration (NF) and reverse osmosis (RO) along with novel membrane technologies such as direct contact membrane distillation (DCMD) and forward osmosis (FO) as possible treatment technologies for cooling tower water. Results from bench-scale testing showed that all four technologies were able to effectively treat the cooling tower water. Membrane cleaning strategies were also explored for the NF/RO and DCMD systems. Results showed that, in most cases, the fluxes were recovered and the membranes maintained high rejection after cleaning. Pilot studies were conducted for approximately 300 hr using an NF system with ultrafiltration (UF) pretreatment; performance was evaluated at 30 and 60% water recovery. Results for pilot testing indicate that high quality water can be produced while operating at high fluxes using pressure-driven processes. Finally, a cost estimate was conducted to evaluate the requirements for a full-scale NF system to treat cooling tower waters. It was determined that membrane systems could possibly save $170,000 per year for a 100-MW facility depending on the cost of an alternative water source.

ACKNOWLEDGMENT

I would like to express my deepest gratitude to my advisor, Professor Amy Childress for her guidance and patience. She has truly prepared me for the obstacles to come. I would like to sincerely thank Professor Tzahi Cath for his support, mentoring, and constant persistence throughout my thesis research. Without him, this would not have been possible.

I would like to thank my committee members, Professor Edward Kolodziej for his support in this project and his trust in me, and Professor Glenn Miller for his enthusiasm for my research and his valued opinion.

Thank you, Sarah McCormick, Andrea Achilli, Kerri Hickenbottom, Eric Orset, and Jeff Rushowitz for your assistance and interest in this research. I hope our relationship as friends and colleagues continues through the years.

I would like to thank the National Renewable Energy Laboratory for providing the support for this research project. I would also like to thank Ormat, Inc., especially Eyal Hen and Dwayne Gaylord, in Reno, NV for allowing me to conduct pilot testing at their facility and helping me to install the membrane pilot system.

And thank you to my family – my parents, Kim and Larry Willis, Klint and Mary Hutton, my brothers, Jayson Johnson and Trevor Hutton, and my sister Bobbie Jo Medling - your love, encouragement, and support were a constant source of strength that helped me through this process. I would also like to thank all of my friends for their constant encouragement and support. Finally, I would like to thank my boyfriend, Chris Lockie (and my pooch, Tripp) for keeping sane throughout numerous nights of insanity.

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Introduction

Water and energy are intertwined. Energy is required to extract, clean, and distribute drinking water and to clean wastewater before discharging it to the environment. On the other hand, water is required to produce energy, both to generate the steam that spins turbines and as a cooling medium for the processes. This interdependency, coupled with the increasing population and increasing standard of living in the US, is creating an ever increasing demand on the nation's water and energy resources. In 2000, thermoelectric energy production accounted for 39% (~135 billion gal/day) of all freshwater withdraws in the US. In addition to cooling power plants, large quantities of fresh water are also used for cooling of buildings and industrial processes [1].

Geothermal power facilities harness energy from superheated water and use it to generate electricity. Geothermal energy is generated by converting hot water or steam from deep beneath the earth's surface into electricity [2]. Geothermal energy is a promising source for renewable energy because it is reliable, clean, and sustainable [3]. The US is currently the largest producer of geothermal power throughout the world and California is the largest geothermal power producer in the US with a total generation capacity of 2,605 MW. In 2007, geothermal power production accounted for 4.5% of the total power generated in California [4]. Nevada has the electrical generation capacity to be the second largest geothermal power producer in the US, but is not currently operating at those outputs. As of 2008, Nevada has 18 geothermal

facilities that at any given time should produce 330 megawatts (MW) of power and Nevada is currently developing more projects than any other state [4].

Typically there are three types of geothermal power facilities: dry steam power plants, flash steam power plants, and binary power plants. Dry steam power plants use steam that is piped directly from the geothermal reservoir to spin a turbine and generate electricity. Flash steam power plants use a flash tank which generates steam from the superheated geothermal brine and the steam is used to spin a turbine and generate electricity. Binary power plants transfer heat from the superheated geothermal brine to another liquid that is typically an organic solvent with a low boiling point that is used to generate steam and spin a turbine to generate electricity [4, 5].

Geothermal power plants can also be a combination of a flash steam and binary power plant (Figure 1). In this scenario, hot geothermal brine is pumped from the geothermal reservoir to a flash tank. The water that is not flashed to steam is then sent through heat exchangers to boil the organic solvent. The steam produced by the organic solvent is used to spin another turbine and generate additional electricity. After the steam passes through the turbine, it flows through a condenser and the liquid is reused. The hot water is returned to the geothermal reservoir after passing through the heat exchanger.

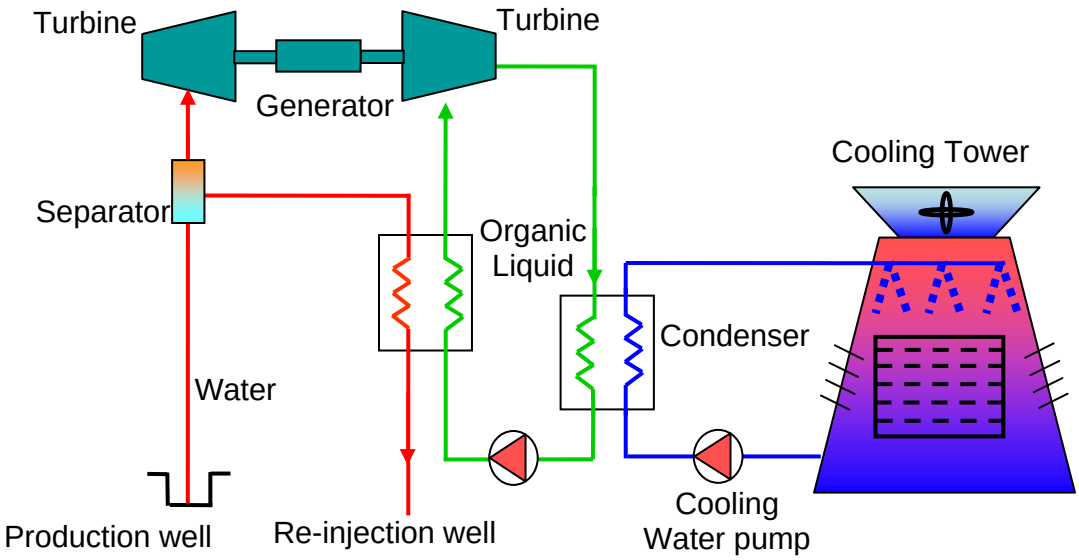

Figure 1. Schematic drawing of a combined flash steam-binary cycle geothermal facility. Electricity is produced in turbines by both super heated steam and organic vapors.

Cooling is an essential part of power generation; it facilitates and dictates the intensity of steam flow through the turbine and therefore the efficiency of the power plant. Typically there are three types of cooling systems: surface water (once-through systems), direct contact, and recirculating non-contact cooling towers [6]. In 2000, 195 Mgal/day were required for cooling systems throughout the United States [7]. Of the cooling systems used, recirculating non-contact cooling towers are the most widely used cooling system throughout the United States [1] with water requirements of 18,300 Mgal/day [7]. Recirculating non-contact cooling towers are the main focus of this investigation (Figure 2). After water has been passed through the condenser it is sent to the cooling tower where heat is removed. The hot water is sprayed onto a highly porous

media to increase the air/water contact and facilitate heat transfer by increasing the surface area of the water. The water is allowed to percolate down the media while a fan is used to pull air through the cooling tower to cool the water by the latent heat of evaporation. Water is also cooled by sensible heat transfer with the surrounding air [6, 8] As the water continues to be cycled through the condenser and the cooling tower, pure water evaporates and the remaining water gets concentrated with salts and other solids. Drift losses due to air flow also cause a loss of volume in the cooling tower [9]. In order to maintain a necessary volume of water, an external water source is used. The external water source is referred to as makeup water for the cooling tower. The makeup water can be from a variety of sources such as the sea, lakes, rivers, irrigation ditches, and groundwater that in some cases is trucked to the geothermal facility [10, 11]. Reclaimed and recycled water have also been shown to be used as a makeup water source [12-14]. Depending on the quality of the makeup water, the severity of problems (e.g., corrosion, scaling, and biological fouling) throughout the cooling tower can vary [6, 15, 16].

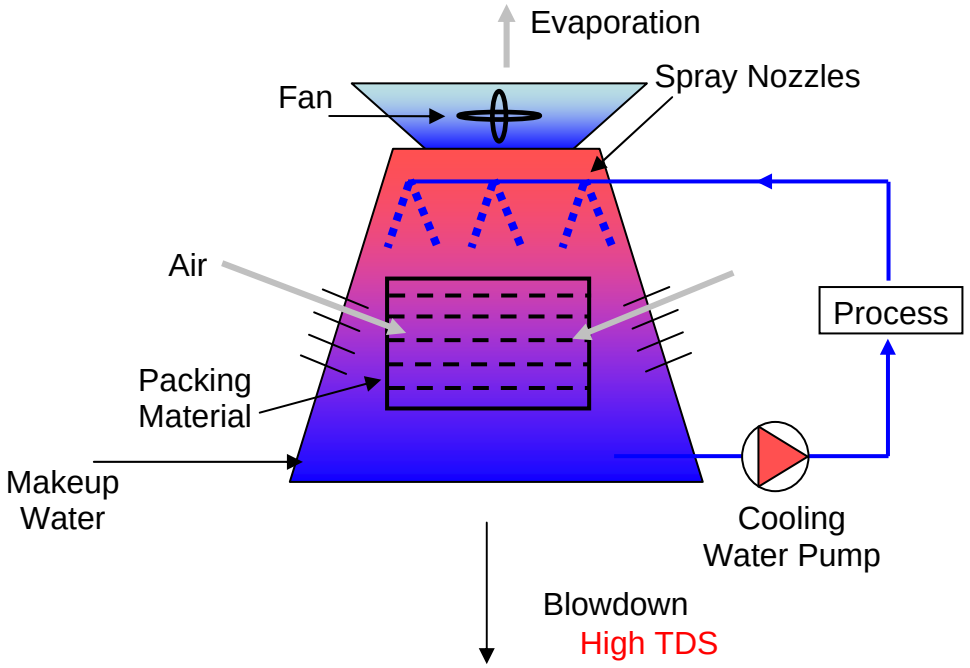

Figure 2. Flow diagram of recirculating, non-contact cooling loop with evaporative tower.

Scaling occurs when salt precipitates out of supersaturated solutions formed during concentration of cooling water. Scale layers can form on any surface in the cooling system and are often deposited on heat exchangers; reducing heat transfer efficiency, and therefore, the overall efficiency of the cooling/energy production process. Corrosion occurs when an electrochemical potential develops between metal surfaces, which drives a redox reaction between the metal ions and the ions present in the surrounding water. Presence of total dissolved solids (TDS) and dissolved oxygen can accelerate this process because there are more ions in the water to react with the metal.

Corrosion usually results in metal oxides or metal salts being formed. These reactions can weaken the metals and reduce performance. The interior of wet cooling towers is cool, moist, and saturated with oxygen; creating an ideal environment for biological growth. Biological growth can be accelerated when makeup water contains organic carbon and nutrients such as phosphorous and nitrogen [16].

When any combination of scaling, corrosion, or biological growth occurs, it is generally referred to as fouling [16, 17]. To help avoid fouling in cooling towers, a portion of the cooling water, the blowdown water, is constantly removed from the system and replaced with makeup water. This reduces contaminant concentrations, and therefore, fouling potential, but subsequently creates inefficient water use. Because cooling water needs to be continually replenished due to evaporative loss, drift, and blowdown, cooling towers require large volumes of water for operation. In order to mitigate the effects that this water demand has, it is necessary to increase the efficiency of water use through development and utilization of new, innovative water treatment and reuse technologies.

Currently, advanced membrane separation technologies are one of the more promising technologies for treatment of cooling tower water. Pressure-driven membrane processes (i.e., reverse osmosis (RO) and nanofiltration (NF)) have already been shown to be an effective method for producing high quality water for reuse [17, 18]. RO and NF separation technologies remove dissolved impurities from water through the use of a semi-permeable membrane. NF separation technologies primarily remove divalent ions (e.g., Ca^{2+}, Mg^{2+}) whereas RO removal will reject divalent ions along with monovalent ions (e.g., Na^{+}). RO removal can be a high energy process because the applied pressure

for removal of ions must be greater than the osmotic pressure created by the high salinity feed solution on the active side of the membrane and the high purity (permeate) stream on the support side of the membrane. Because of the potential for high energy requirements, NF membranes are also being investigated. NF is a much less energy intensive process that will still produce high quality water.

One of the major problems associated with operation of RO and NF is membrane fouling and scaling. Membrane fouling is the formation of deposits on the membrane surface that can cause loss of flux, membrane damage, and increased capital and operating costs [19-22]. To reduce the occurrence or severity of fouling, the amount of contaminant loading to NF/RO systems needs to be reduced. This is typically accomplished by pretreatment of the source water. Various methods of pretreatment exist, such as coagulation coupled with ultrafiltration (UF) and microfiltration (MF), along with UF and MF without coagulation. UF and MF remove particles based on size, the membrane acts as a sieve to remove particles that are approximately less than 0.01 micron (for UF membranes) and approximately less than 1-5 micron (for MF membranes) [23]. Many studies have shown that UF and MF are a reliable pretreatment for RO/NF systems [19, 22, 24-27]. Studies have also shown that coagulation enhance removal for MF and UF [28]. When coagulation is coupled with MF or UF, turbidities have been reduced by 55% for impaired water (approximately 24 NTU) and to less than 0.05 NTU when using water with approximately 15 NTU [20, 28].

Another option for treatment of cooling tower water is direct contact membrane distillation (DCMD). DCMD is a thermally-driven separation. Warmer feed water is in

contact with the active side of the membrane and a cooler water stream is in direct contact with the support side. The driving force for mass transfer in DCMD is the vapor pressure difference across the membrane induced by the temperature difference across the membrane. Because the partial vapor pressure of water is only minimally affected by increased concentrations of dissolved salts, DCMD has the potential to be an ideal treatment method for highly saline feeds [29].

Another potential method for treatment of cooling tower brines is forward osmosis (FO). FO is an advantageous method because it operates with little or no hydraulic pressure and has the ability to reject a wide range of contaminants. FO is an osmotically-driven membrane separation process involving the diffusion of water through a semipermeable membrane. The driving force for mass transport is the difference in osmotic pressure between the feed solution and a draw solution (DS); water diffuses from the feed stream of lower osmotic pressure to the DS of higher osmotic pressure. As water diffuses through the membrane, the feed solution becomes concentrated and the DS is diluted and must be reconcentrated in order to maintain the osmotic driving force.

Previous Investigations of Cooling Tower Water Treatment

A study conducted in Taiwan evaluated the use of treated wastewater effluent for use as makeup water for five industrial plants. The water was treated with multiple combinations of various pretreatment steps (e.g., coagulation, chemical oxidation, and biological oxidation) followed by UF and RO membrane systems. The water that was produced was able to meet drinking water standards, so it was easily used as a makeup

water source. The concentrate that was produced met effluent standards and was able to be discharged to the environment [12].

Another study used membrane technology to treat the blowdown from the cooling tower. The study evaluated the effectiveness of MF and UF as pretreatment for RO. Both MF and UF successfully removed suspended solids (SS) and colloids in the wastewater prior to treatment by RO. The RO system was able to achieve >99% rejection for most constituents, so the water was acceptable for reuse in the cooling towers [14].

No studies were found on testing of DCMD or FO for treatment of cooling tower water.

Objectives and Scope of Work

The overall goal of this project was to evaluate membrane processes, such as pressure-driven processes (e.g., RO and NF) and novel membrane processes (e.g., MD and FO), for treatment of cooling tower water. The sub-objectives identified to carry out this goal included: 1) testing all three processes at bench-scale to compare the performance of novel membrane processes with conventional, pressure-driven membrane processes, 2) testing of pressure-driven processes at pilot-scale to evaluate system performance (flux, fouling, and rejection), and 3) modeling of pressure-driven processes to evaluate the economics of full-scale. Specifically it was desired to address the following questions: 1) would NF be a more cost effective way of achieving the desired TDS reduction than RO, 2) what pretreatment and cleaning requirements would be

required using actual cooling tower water in the pilot system, and 3) what is the current cost savings and what would be the future cost savings.

Materials and Methods

Feed Water for Bench-Scale Experiments

Bench-scale testing was performed in the University of Nevada, Reno (UNR) Laboratory using water from the Ormesa 1 geothermal power facility in Imperial Valley, CA. Water samples were collected in 5-gal containers from locations at Ormesa 1. Raw water (designated O-RAW) was collected directly from the irrigation canal and blowdown water (designated O-BD) was collected from the bottom of the cooling tower, just before it is mixed back into the pond. The three feed waters were analyzed in the Aqwatec Environmental Engineering Laboratory at CSM. Cations were analyzed using inductively coupled plasma (ICP) (Perkin Elmer, Norwalk, CT), anions were analyzed using ion chromatograph (IC) (Agilent, Santa Clara, CA), total organic carbon (TOC) was measured using a TOC analyzer (GE Analytical Instruments, Boulder, CO), turbidity was measured using a turbidimeter (Hach, Loveland, CO), and silt density index (SDI) was measured according to ASTM D 4189-82 standard method.

Feed Water for Pilot-Scale Experiments

Pilot-scale testing was performed at the Steamboat Hills (SBH) geothermal power facility in Reno, NV. Water from the cooling tower basin was used for the pilot testing.

The water collected in the cooling tower basin is water that has been used in heat exchangers throughout the geothermal facility to remove excess heat from the geothermal brine. The water is then sent to the cooling tower where the heat is removed by allowing the water to percolate down various media while exposed to the atmosphere. Throughout this process, evaporation causes a reduction in water quality and quantity. Makeup water is often brought into the plant from varying sources to replenish losses due to evaporation and blowdown. At SBH, the makeup water is primarily steam condensate, however, alternate makeup water sources are groundwater that is trucked to the facility or effluent from the chemical abatement tower (CAT). The CAT removes hydrogen sulfide (H_2S) from the air being discharged to the environment, which creates an acidic solution that must be neutralized prior to leaving the CAT. The solution is neutralized with a caustic solution (NaOH); this neutralized solution is used as makeup water. Thus, the quality of the water in the cooling tower basin can be highly variable.

The water was collected in a 5-gal container from the cooling tower basin at SBH (SBH-BD). The water was analyzed in the Environmental Engineering Laboratory at University of Nevada, Reno (UNR). Cations were analyzed using an ICP (Perkin Elmer, Norwalk, CT), anions were analyzed using an IC (Dionex, Sunnyvale, CA), total organic carbon (TOC) was measured using a TOC analyzer (Shimadzu, Columbia, MD), and turbidity was measured using a turbidimeter (Hach, Loveland, CO).

Reverse Osmosis/Nanofiltration

Membranes

Three DOW Filmtec (Midland, MI) membranes were tested: the NF90, NF4040, and XLE (RO) membranes. Two KOCH (Wilmington, MA) membranes were tested: the ULP (RO) and TFC-S (NF) membranes. One TriSep (Goleta, CA) membrane was tested: the XN45 (NF) membrane. These membranes were chosen because of their ability to operate at low pressures and their reported high salt rejection (Table 1). All membranes were supplied as flat-sheet membranes and stored at 5 ºC.

Table 1. Membrane Properties

Membrane	Type	Typical Operating Pressure (psi)	Average Rejection (%)	Compaction pressure (psi)	Operating Pressure (psi)
NF90	NF	70	>97.0 $MgSO_4$	40-50	48
4040	NF	130	98.0 $MgSO_4$	40-50	44
TFC-S	NF	75-125	99.0 $MgSO_4$	70	70
XN45	NF	40-200	95.0 $MgSO_4$	60	35
XLE	RO	100	99.0 NaCl	40	50
ULP	RO	75-175	98.5 NaCl	70	66

Bench-Scale Testing

The objective of the bench-scale testing was to obtain performance data that would be used to select a membrane for pilot testing. A bench-scale apparatus (Figure 3) was used in evaluating the selected membranes. Each membrane was hydrated for 24 hrs prior to the experiment. After being hydrated, the membranes were compacted for 15-20

hrs with doubly deionized water (DDW) at the compaction pressures listed in Table 1. The compaction pressures were determined based on typical operating pressures and membrane specification sheets.

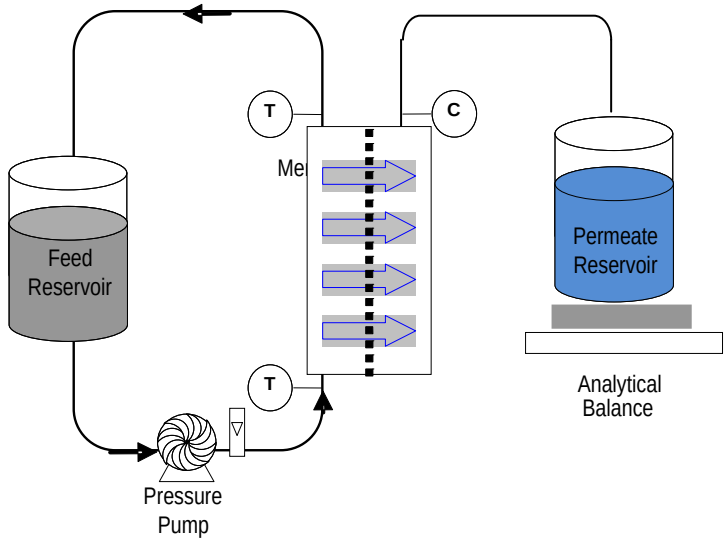

Figure 3. Schematic of NF and RO experimental system configuration.

Filtered O-BD was used as the feed solution for these experiments. O-BD was filtered through a 5-µm filter followed by a 0.5-µm cartridge filter and was then stored in a feed water reservoir. The filtered O-BD was pumped from the reservoir using a high pressure piston pump (Hydra-Cell, Minneapolis, MN) and through a SEPA-CF membrane cell (Osmonics, Minnetonka, MN) with a feed channel that was 95 mm wide and 1 mm high. The membrane cell was fitted with feed and permeate spacers in an attempt to simulate the hydrodynamics of a spiral-wound membrane element. The membrane surface area was 0.0139 m^2. Pressure was adjusted and maintained using a

pressure transducer coupled with an electronic control valve; pressure was constantly monitored using a SCADA system and LabVIEW Software (National Instruments Corp., Austin, TX). The operating pressures were adjusted to achieve an initial permeate flux between 17-24 L/m^2-hr (according to DOW's *Membrane System Design Guidelines*). The operating pressures used for this investigation are summarized in Table 1. The permeate was collected in a reservoir on an analytical balance that was linked to a computer and the weight of the permeate reservoir was recorded. The concentrate was returned to the feed water reservoir. The feed water temperature was maintained at 23 ± 2 ºC using a chiller (Fisher Scientific, Pittsburgh, PA). The flow rate was kept constant at 1.5 L/min and was monitored using a rotameter (Key Instruments, Trevose, PA).

The experiments were terminated when the flux dropped below 5 L/m^2-hr. At this point, the membranes were cleaned using an ethylenediaminetetraacetic acid (EDTA) solution. Sodium hydroxide (NaOH) was added to the EDTA solution until the pH was 11 [30]. This solution was then circulated through the system at a slow flow of 0.5 L/min for 30 min. Next, the system soaked for 30 min and after soaking, the EDTA solution was flushed out of the system at a high flow rate. The system was then rinsed with DDW. To evaluate the effectiveness of the cleaning procedure, flux experiments were performed again using 2 L of fresh filtered O-BD.

Consecutive Cleaning Experiments

Two of the six membranes were selected for consecutive cleaning experiments. Consecutive cleaning experiments were conducted using the same systems and

procedures for the initial bench-scale testing experiment; however, three cleaning cycles and a fourth restart were conducted to evaluate flux recovery over a series of fouling cleaning cycles.

Pilot-Scale Testing

A pilot-scale RO/NF membrane system was used in this investigation (Figure 4a). The system used a diaphragm pump (Hydra-Cell, Minneapolis, MN) to circulate the water through two membranes in series. The system was designed for two recovery rates (30 and 60%). For each recovery, a desired average flux was chosen. Using the desired flux along with the membrane area, the permeate flowrates were calculated. The permeate flowrates were held constant to ensure the desired flux and recoveries were achieved. To maintain the desired permeate flowrates, the applied pressure was increased, which was later used to determine the specific flux. The influent feed flowrate was held constant to support similar feed flow velocity and to ensure that the flow velocity in the tail element would still be above the minimum required. Recycle flows were calculated using ROSA membrane software. The concentrate and recycle flows were determined based on the desired recoveries (Table 2). For each recovery, membrane performance was tested for 100 hr in order to evaluate the membrane's ability to operate continuously with varying degrees of fouling and evaluate the membrane's recovery after cleaning.

Table 2. Desired flowrates depending on recovery rates

Recovery (%)	Total Flow (L/min)	Flux (L/m^2-d)	Permeate Flow (L/min)	Recycle Flow (L/min)
30	23	488	5.3	5.7
60	19	488	5.3	8.7

The feed water to the pilot system was pretreated using a pilot-scale UF membrane system (Figure 4b). Water was pumped from the cooling tower basin using a centrifugal pump (TEEL, Grainger) to the UF membrane (Hydranautics, Oceanside, CA). A coagulant (alum) was continuously fed to the feed water to increase turbidity removal by the UF system. Sodium metabisulfite was also continuously added to eliminate chlorine from the feed water. The UF permeate flow rate was monitored using a rotameter (specifications unknown). The permeate water was collected in a reservoir that was used for storage of the feed water for the RO/NF membrane system. The RO/NF permeate was collected in a reservoir that was attached to the UF system and was used for UF membrane backwashing. Figure 5 is a schematic illustrating the flow of water through both the UF pretreatment system and the RO/NF pilot system.

Figure 4. (a) RO/NF pilot system and (b) UF pretreatment system.

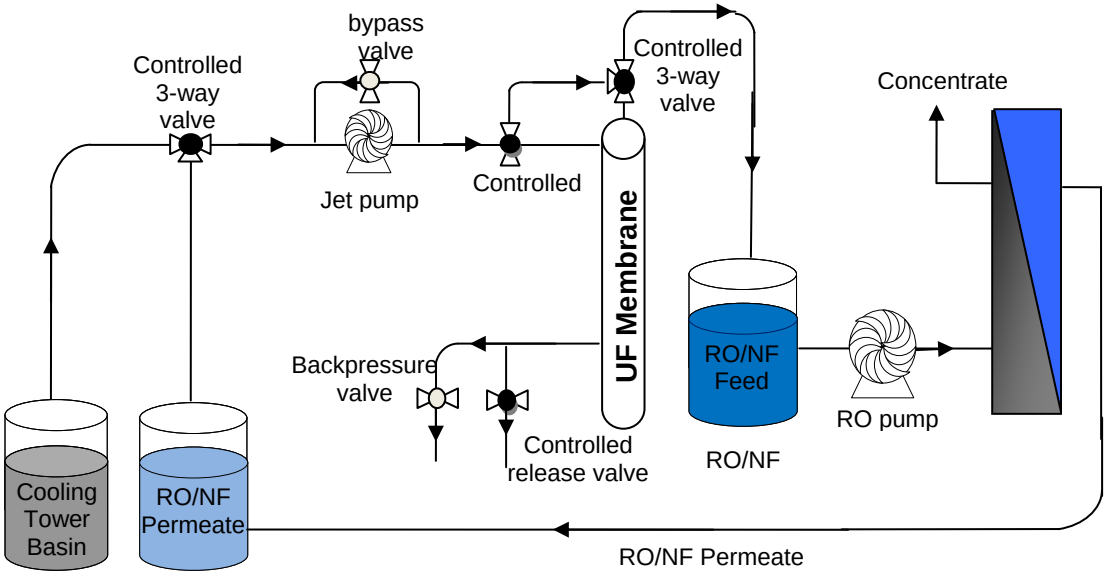

Figure 5. Schematic of flow of water through UF pretreatment system and the RO/NF pilot system.

Membrane Distillation

Membranes

Two microporous flat-sheet membranes were used for the bench-scale DCMD experiments; the membranes were acquired from GE Osmonics (Minnetonka, MN). One membrane is a composite membrane made of a thin polytetrafluoroethylene (PTFE) active layer and a woven mesh polypropylene (PP) support layer. The effective pore size of the PTFE membrane is 0.22 μm. The GE PP membrane is a symmetric, isotropic membrane made of PP. The effective pore size of the PP membrane is 0.2 μm.

Bench-Scale Testing

The membranes were tested in a modified SEPA-CF cell which has symmetric channels on both the active and support sides of the membrane to allow for tangential flow on both sides of the membrane. The active membrane surface area was 0.0139 m^2.

The performance of the DCMD process was evaluated using a bench-scale membrane test system (Figure 6). 4 L of feed solution were added to the feed reservoir and heated to 40 ºC using a hot bath (Precision Scientific, Winchester, VA). 2 L of deionized (DI) water were added to the permeate reservoir and cooled to 20 °C using a heat exchanger connected to a chiller (Thermo-Electron, Newington, NH). Feed water and permeate temperatures were monitored using dual channel digital thermometers (Cole-Parmer, Vernon Hills, IL) with thermocouples located at the inlets and outlets of the modified SEPA cell. The feed solution was recirculated on the active side of the membrane at 1.5 L/min and the DI water was recirculated countercurrently on the support

side of the membrane at 1.5 L/min. As water evaporated through the membrane, the concentration of the feed stream slowly increased and excess water from the permeate reservoir was allowed to overflow and collect in a beaker. The overflow water was continuously weighed on an analytical balance (Mettler Toledo, Inc., Columbus, OH) and the data was used to calculate water flux and batch recovery. A conductivity meter (Jenway Ltd., U.K.) was used in the permeate reservoir to ensure rejection.

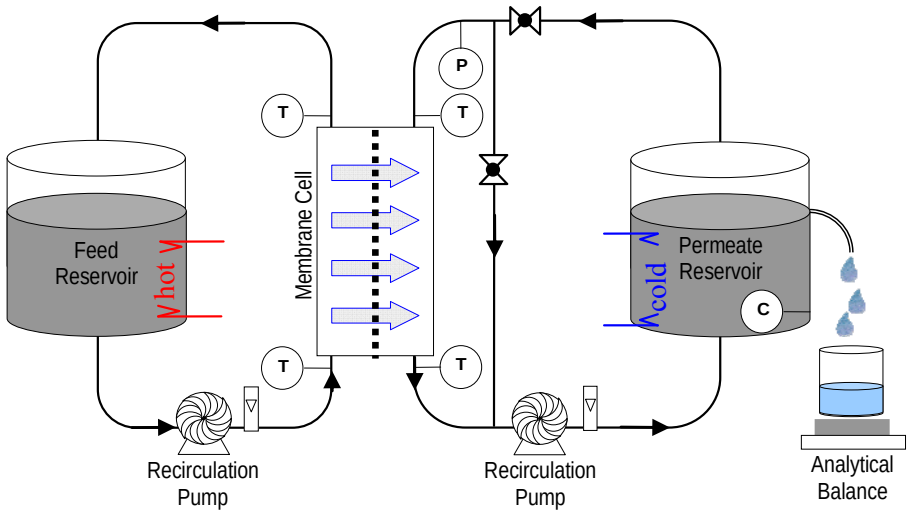

Figure 6. Schematic of bench-scale DCMD test system. Adapted from [29].

Forward Osmosis

Membranes

A flat-sheet cellulose triacetate (CTA) FO membrane was used for the FO experiments. The CTA membrane was specifically developed for FO applications and was acquired from Hydration Technologies Inc. (Albany, OR).

Bench-Scale Testing

A schematic drawing of the FO bench-scale system is illustrated in Figure 7. The CTA membrane was installed in a modified SEPA cell that has symmetric channels on both sides of the membrane similar to the cell used for the DCMD experiments. This allowed for both the feed and the draw solution (50±2 g/L NaCl) to flow tangential to the membrane. The effective membrane surface area was 0.0139 m^2. Mesh spacers placed in the feed and draw solution channels supported the membrane and enhanced mixing. Two variable-speed gear pumps (Cole-Palmer, Vernon Hills, IL) were used to recirculate the feed and draw solutions on opposite sides of the membrane at 1.5 L/min. The temperatures of the feed and draw solutions were monitored with thermometers (Cole-Parmer, Vernon Hills, IL) installed at the inlets of the SEPA cell. The temperature was held constant at 25 °C. The feed solution was contained in a 5.4-L constant-level reservoir. Both constant feed concentration experiments and increasing feed concentration experiments were performed. For constant feed concentration experiments, the feed volume was kept constant by continuous replenishment of the water that crossed the membrane by the DDW reservoir. The DDW reservoir was placed on an analytical balance linked to a computer. Flux through the membrane was calculated based on the change of weight of the DDW water reservoir. The draw solution was contained in a 10-L reservoir. The draw solution concentration was held constant by continuous addition of concentrated draw solution from a concentrated draw solution reservoir. pH and conductivity of the feed and draw solutions were monitored with meters placed in the feed and draw solution reservoir. For the increasing feed concentration experiments, as

water diffused through the membrane, the water level in the feed reservoir declined and the ion concentration of the feed stream increased. The feed reservoir was weighed using an analytical balance (Mettler Toledo, Inc., Columbus, OH) and the data was used to calculate water flux and batch recovery.

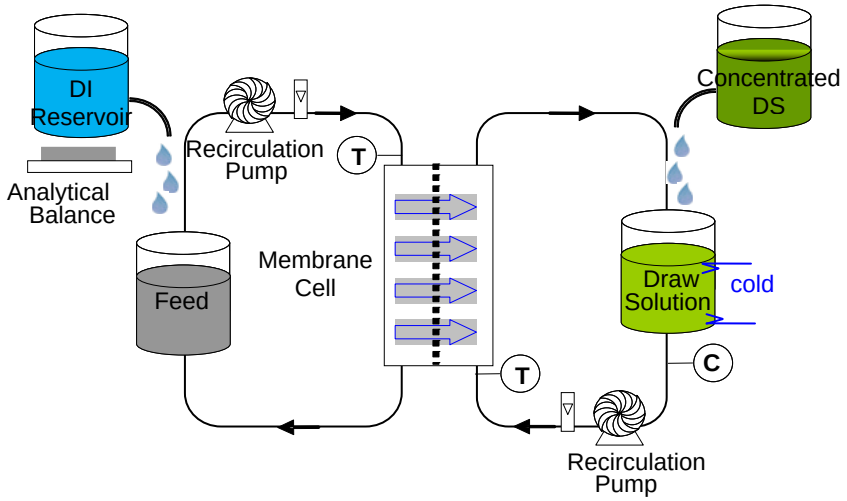

Figure 7. Schematic of bench-scale FO system.

Cost Modeling

Two NF systems were modeled using RO System Analysis (ROSA) membrane design software. Both systems consisted of two stages of NF90-400 membrane elements. A schematic of the modeled NF system is illustrated in Figure 8. The first system was designed to optimize both recovery and membrane area. Both stages had a variable number of pressure vessels and a variable number of elements in each pressure vessel. The first system was evaluated for the treatment of O-BD and SBH-BD feed water (Table

3) for 55-80% water recoveries. Model output (i.e., number of pressure vessels and elements per pressure vessel) at 80% recovery was used to design the second system.

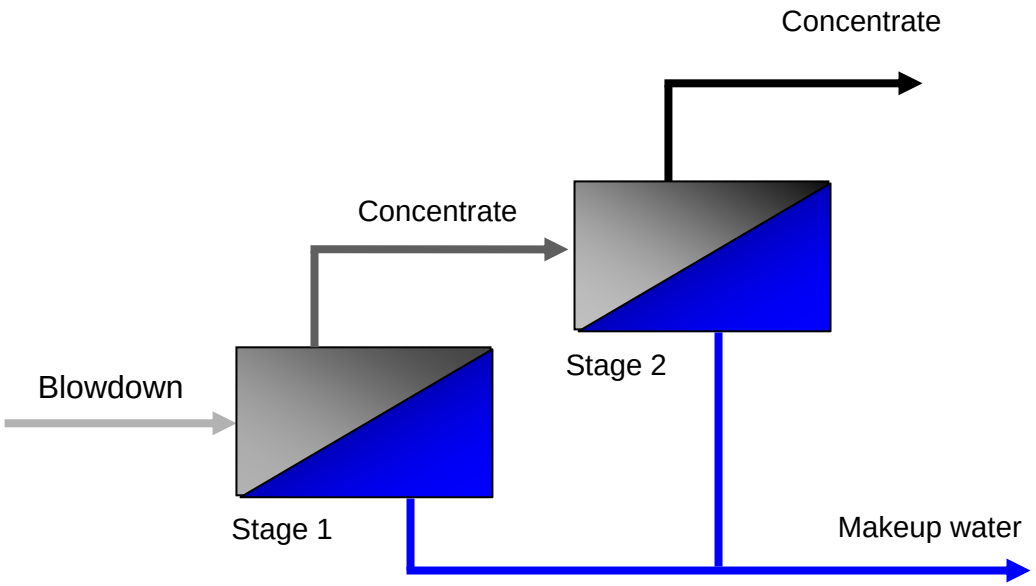

Figure 8. Schematic drawing of the modeled NF system. Stages have a variable number of pressure vessels and each pressure vessel has a variable number of membrane elements for the initial model. The second model has a constant number of pressure vessels and elements in each stage.

The second system was designed to evaluate the recovery that could be achieved using a system with a set number of pressure vessels and elements per pressure vessel. The second system was evaluated for treatment of O-BD feed water for 55-80% water recoveries.

To determine the flowrate of blowdown water that will be generated and available for treatment and water recovery/reuse, a simplified equation that relates operating cycles

to cooling tower flows was used:

$$\#\text{ of cycles in the cooling tower} = Q_{makeup} / Q_{blowdown\ volume}$$

where Q_{makeup} is the flowrate of makeup water and $Q_{blowdown}$ is the flowrate of blowdown water from the cooling tower. This equation assumes that drifting losses are negligible. Both systems were modeled assuming a 10-MW facility requiring 3450 L/min (900 gal/min) of makeup water for cooling tower operation [31]. It was assumed that the number of cycles in the cooling tower was five. Based on these values, the blowdown flowrate is 690 L/min (177 gal/min), which is used as the feed water flowrate for both modeled systems.

Results and Discussion

Water Quality

The water qualities of the two process waters from the Ormesa 1 geothermal facility are shown in Table 3 . The calcium concentration coupled with the sulfate concentration in O-BD is cause for concern because calcium sulfate is a known scalant in membrane processes. It has a low solubility and will precipitate out of solution causing a decrease in flux or increase in applied pressure. The TOC and turbidity concentrations in O-BD was approximately 3-4 times greater than those in O-RAW [16]. TOC in water is caused by decaying natural organic matter (NOM) which has been shown to be a significant factor in membrane fouling [32]. SDI values were only obtained for O-RAW. After five min of filtration, the SDI was calculated to be 16.7, however, the filter became

>90% clogged after approximately 10 min of filtration time. SDI values could not be calculated for O-BD [16]. Typically membrane manufacturers will not guarantee membrane performance if a feed water with an SDI >5 is used. The high blockage rates during the SDI testing of these waters indicate high fouling potential. O-BD was chosen for use in the bench-scale testing because it was one of the more challenging waters. Based on the values obtained for TOC, turbidity, and SDI, pre-treatment must be performed if membrane processes such as RO, NF, or MD are to be used.

Table 3. Water quality of two process waters from Ormesa 1 geothermal facility (Imperial Valley, CA) and one process water from Steamboat Hills geothermal facility (Reno, NV)

	Ormesa 1 Raw Water (O-RAW)	**Ormesa 1 Blowdown Water (O-BD)**	**Steamboat Hills Blowdown (SBH-BD)**
Ca^{2+} mg/L	81.56	277.56	5.13
Mg^{2+} mg/L	30.14	107.70	1.27
Na^{+} mg/L	111.54	402.70	587.7
Si^{2+} mg/L	3.90	13.31	21.58
Cl^{-} mg/L	95.37	408.53	77.6
SO_4^{2-} mg/L	261.4	1372.7	968.4
TOC mg/L	3.05	10.00	3.94
Turbidity NTU	1.30	3.00	7.20
SDI	16.7	> 90% blockage	NA
Res Cl mg/L	0.00	0.40	0.50
pH	7.83	7.82	8.6

The water quality of the SBH-BD that was used for pilot testing at the Steamboat Hills geothermal facility in Reno, NV is also summarized in Table 3. The calcium concentration is significantly less than concentrations in O-BD indicating that calcium

sulfate will not likely scale the membrane. The SBH-BD water has higher turbidity, which is nearly double the turbidity of O-BD suggesting that particulate matter will be more of a problem during pilot testing. SDI was not performed on this water. However, the high value of turbidity again supports the need for pretreatment of the SBH-BD.

Reverse Osmosis/Nanofiltration

Bench-Scale Testing using O-BD

Normalized specific flux as a function of time for the six membranes is shown in Figure 9. By using normalized specific flux to report the results, both the applied pressure and difference in initial fluxes are accounted for. The experiments were terminated when all the water had been recovered or the flux was below 5 L/m^2-hr. At this point, the membrane was cleaned and the flux after cleaning was analyzed. Results in Figure 9 illustrate that most of the membranes followed a similar pattern of relatively severe flux decline starting at the beginning of the run and continuing over the next 14-20 hr. After cleaning, the flux for most membranes was recovered back to the initial flux or close to it. The XN45 membrane did not follow this pattern. The initial flux decline less severe over the first 4 hr and then became more dramatic for approximately 2 or 3 hr before beginning a rapid decline until a flux of 15 L/m^2-hr was reached and all of the water had been processed. At this point, the membrane was cleaned and after cleaning, did not show the same pattern as the other membranes. The flux recovered to 25 L/m^2-hr, which was greater than the initial operating flux of 21 L/m^2-hr. Along with the XN45 membrane, the 4040 membrane also had a higher flux after cleaning. The higher flux could be due to

residual cleaning agent or NOM on the membrane that made the membrane more hydrophilic and allowed water to pass more easily through the pores

For each membrane, the higher flux was not likely due to membrane damage caused during cleaning because the conductivity before cleaning was much higher than the conductivity after cleaning (Table 4).

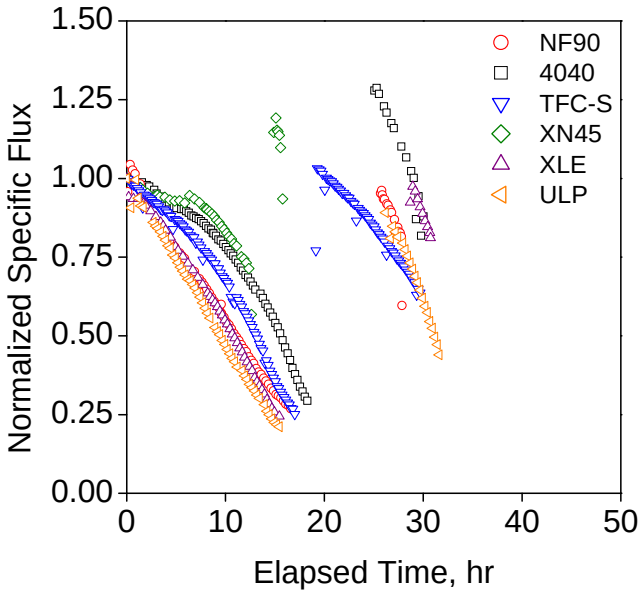

Figure 9. Normalized specific flux as function of time for all NF and RO membranes tested at bench-scale. Experiments were conducted with O-BD feed water with a temperature of 23±2 °C. The operating pressure for each membrane is shown in Table 4.

Table 4. Summary of operating conditions and membrane performance for bench-scale membrane testing

	Operating Pressure (psi)	Initial Flux (L/m^2-hr)	Flux Decline	Flux After Cleaning (L/m^2-hr)	Permeate Conductivity (µS/cm)		
					Initial	Before cleaning	After cleaning
NF90	48	19	Med	17	148	503	237
4040	44	17	Med	22	1374	1884	1913
TFC-S	70	20	Med	21	150	1270	178
XN45	35	20	Low	24	2690	3533	2980
XLE	50	21	High	20	206	3370	1180
ULP	66	20	High	20	178	321	NA

In addition to permeate conductivity, Table 4 also summarizes the operating pressures and fluxes of the membranes tested. Flux decline was assessed over the first 15 hr of operation and is indicated by a high (> 75%), medium (25-75%), or low (< 25%) flux decline. The flux after cleaning was used to determine the efficiency of the cleaning procedure.

Consecutive Cleaning Experiments

Results from the bench-scale testing were used to select two membranes for further study. When choosing a membrane for further testing, it was desirable that the membrane had a high rejection and high recovery after cleaning. Flux decline and operating pressures were also considered. The two membranes chosen for further testing were the NF90 membrane and the TFC-S membrane; both are NF membranes. The NF90 membrane maintained high rejection (~90%) throughout the duration of testing and the flux after cleaning (17 L/m^2-hr) was 90% of the initial flux; indicating that most of the fouling was reversible. The flux decline was moderate and the operating pressure (48 psi)

was the lowest of the membranes that exhibited high rejection. Compared to the NF90 membrane, the TFC-S membrane had a relatively high permeate conductivity (1270 μS/cm) at the termination of the flux experiment, but after cleaning, the permeate conductivity was reduced to 178 μS/cm, indicating that the membrane was still rejecting most dissolved solids. After cleaning, the TFC-S membrane flux returned to the initial flux, indicating that the fouling was reversible. The flux decline was moderate and the operating pressure (70 psi) was the highest of the membranes evaluated.

Consecutive cleaning experiments were performed on the NF90 and TFC-S membranes to select one of them for pilot-scale testing. For pilot-scale testing, it is desirable that the membrane be able to operate for long periods of time with minimal flux decline, and have the ability to retain flux and rejection after cleaning. Figure 10 illustrates the results for consecutive cleaning experiments, three cycles of concentration and cleaning and a short fourth cycle to measure the initial flux after the third cleaning, using O-BD. The operating pressures were the same as in the single cleaning experiments.

The NF90 membrane exhibited the same trends as seen in the initial testing. The initial flux was approximately 19 L/m^2-hr and steadily dropped over 18 hr. After the first and second cleanings, the initial flux was approximately 17 L/m^2-hr (representing a 90% flux recovery) and the final cycle had an initial flux of approximately 16 L/m^2-hr (representing an 84% flux recovery). Permeate conductivity before the first cleaning was 514 μS/cm and after cleaning was 146 μS/cm, thus high rejection was maintained. This was true for each cleaning cycle.

The TFC-S membrane had more rapid flux decline for the consecutive cleaning experiments than the single cleaning experiments. For the single cleaning experiments, the operation time was approximately 17 hr whereas the consecutive cleaning experiments operation time was approximately 15 hr achieving the same recoveries. Both the single cleaning and consecutive cleaning experiments had a slightly higher flux after cleaning, likely due to a small amount of cleaning solution left on the membrane [32]. The higher initial fluxes are not likely due to membrane damage because the permeate conductivity before the first cleaning was 1550 µS/cm and after cleaning was 122 µS/cm, showing that the membrane rejection was maintained after cleaning. This was true for each cleaning cycle.

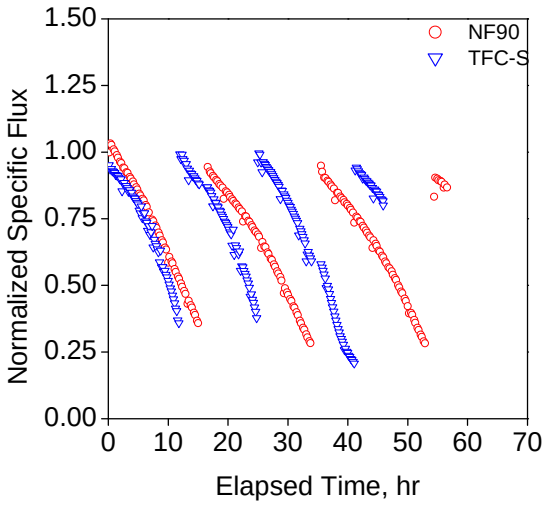

Figure 10. Water flux as function of time for NF90 and TFC-S membranes in consecutive cleaning experiments. Experiments were conducted with O-BD feed water with a temperature of 23±2 °C. The operating pressure for the NF90 membrane was 48 psi and for the TFC-S membrane was 70 psi.

Bench-Scale Testing using SBH-BD

At this point, selection of the most appropriate membrane for pilot testing was not obvious. Therefore, a final set of bench-scale experiments was performed on both membranes using SBH-BD, the feed water that would be used for pilot testing. The results are illustrated in Figure 11 and indicate that under the same operating conditions, the NF90 membrane (48 psi) had similar results as the O-BD experiments. The initial flux was again approximately 19 L/m^2-hr and steadily dropped over the course of 18 hr. Unlike the consecutive cleaning O-BD experiments, the flux after cleaning was not recovered as well. The flux after cleaning was approximately 16 L/m^2-hr (representing an 84% flux recovery). Permeate conductivities indicated high rejection (> 90%) throughout the experiment and after cleaning with the permeate conductivity before cleaning at 109 µS/cm and after cleaning at 26 µS/cm.

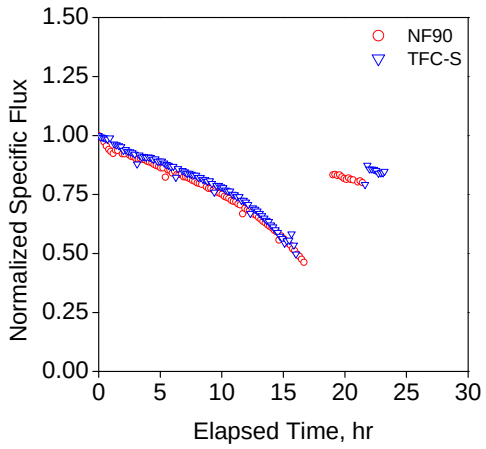

Figure 11. Normalized specific flux as function of time for single cleaning experiments. Experiments were conducted with SBH-BD feed water with a temperature of 23±2 °C. The operating pressure for the NF90 membrane was 48 psi and for the TFC-S membrane was 70 psi.

The TFC-S membrane performed very similarly to the single cleaning O-BD testing and somewhat less flux decline was observed using the SBH-BD than the consecutive O-BD experiments. The system operated for approximately 16 hr showing a much less dramatic flux decline than experiments conducted with O-BD. After cleaning, the membrane did not show the same flux recoveries. Both experiments using the O-BD had higher fluxes after cleaning than the initial flux. However, in SBH-BD experiments, the membrane recovered to nearly the initial flux (18 L/m^2-hr). The different ions in the feed waters could be a possible reason that membrane cleaning performed different for the two feed waters. The O-BD flux decline could be due to calcium sulfate scaling the membrane. However, calcium sulfate will not cause scaling for the SBH-BD. Also, the SBH-BD caused irreversible fouling of the membrane whereas the O-BD water did not. The conductivity trends remained the same with the conductivity before cleaning 2560 µS/cm and after cleaning 27µS/cm.

Based on the results of O-BD and SBH-BD single cleaning experiments and O-BD consecutive cleaning experiments, the NF90 membrane was chosen for pilot-scale testing. As stated earlier, it is desirable to have a membrane that can operate for long periods of time with minimal flux decline and high salt rejection. Both membranes demonstrate the ability to operate well over consecutive cleanings and both had somewhat similar flux decline, however the NF90 membrane showed consistently high rejection throughout experimentation and after cleaning, whereas the TFC-S membrane had consistently low rejection towards the end of the experiments.

Membrane Distillation

Bench-Scale Testing

DCMD bench-scale experiments were performed at both UNR and CSM using the PTFE membrane and O-BD (Figure 12). The initial flux for both systems is near 13 $L/m^{2-}hr$. After equilibration, both systems maintain a steady flux for approximately 7-10 hr before flux decline begins to occur. In the UNR experiment, flux decline continues until the flux is approximately 7 L/m^2-hr (after approximately 16 hr) at which point the flux starts to increase slightly due to an increase in feed water temperature. The volume of water remaining in the feed reservoir was minimal, thus maintaining the desired feed water temperature was difficult. Results from the CSM experiment maintained a relatively constant flux for approximately 9 hr before a steady flux decline was observed. It should be noted that experiments at CSM were terminated earlier; they were only performed until the system reached 60% recovery with a final flux of approximately 11 L/m^2-hr at which point the system was stopped and the membrane was cleaned. However, the results were still similar to those obtained at UNR indicating the reproducibility of the DCMD experiments.

Experiments were also conducted at UNR using the SBH-BD feed water. Results in Figure 12 show an initial flux of approximately 13 L/m^2-hr which slowly increases due to equilibration until the flux reaches approximately 14 L/m^2-hr and continues to slowly increase over 15 hr until no more water is able to be recovered. No flux decline is exhibited using the SBH-BD because unlike the O-BD experiments, there is little presence of calcium sulfate, so membrane scaling and thus flux decline is not observed.

The continuous increases observed throughout experimentation could imply that membrane damage occurred, however, the permeate conductivity was less than 4 µS/cm throughout the entire experiment, indicating high rejection of contaminants, so the increase is likely due to slight fluctuations in permeate and feed water temperatures.

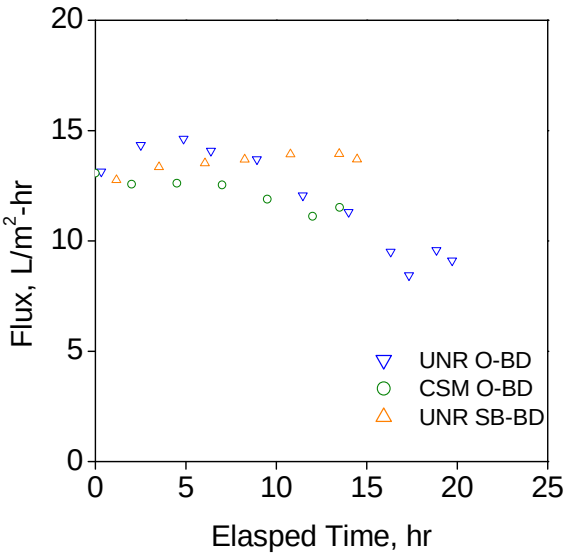

Figure 12. Water flux as function of time for DCMD experiments. Experiments were conducted with either O-BD or SBH-BD feed water with a ΔT of 20 °C (permeate T = 40 ºC and feed water T = 20 ºC).

Longer-term testing on the PTFE membrane was performed at CSM. Results of the long term experiments performed at CSM using the PTFE membrane and PP membrane are shown in Figure 13. After the completion of cycle 1, the membrane was cleaned. The PTFE membrane had the same initial flux of approximately 13 L/m^2-hr after

cleaning and the same trend is seen for each cleaning cycle. No irreversible fouling was detected because the water flux returned to the initial flux after each cleaning cycle.

The initial water flux through the PP membrane (approximately 5 L/m^2-hr) was much lower than the initial water flux for the PTFE membrane (approximately 13L/m^2-hr). The lower flux is due to the fact that the PTFE membrane is a more highly porous composite membrane whereas the PP membrane is much less porous and a symmetric membrane. Since the driving force for MD is vapor pressure, a membrane with less pores will not allow as much vapor to cross the membrane as a highly porous membrane. However, the PP membrane underwent much less flux decline throughout the fouling cycles possibly due to a slower concentration of the calcium sulfate. This is obvious in Figure 14 which illustrates flux as a function of water recovery. Results show that the flux decline for both the PTFE and the PP membrane occur at the same recovery (approximately 34%) indicating at this point that calcium sulfate reaches saturation and begins scaling the membrane. After each cleaning, the initial flux was slightly increased. This could have been caused by pore flooding or membrane damage; however, the conductivity of the distillate stream remained very low (<7 μS/cm), indicating that the probability of membrane damage was low. Fluctuations in the feed water or distillate temperatures or flow rates may have contributed to the slight flux increases.

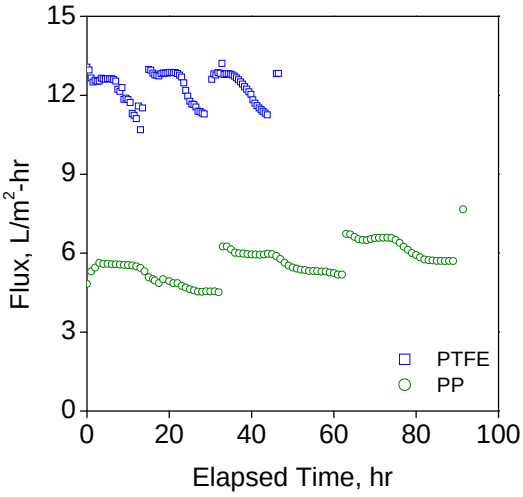

Figure 13. Water flux as function of time for cleaning DCMD experiments conducted at CSM. Experiments were conducted with O-BD feed water with a ΔT of 20 °C (permeate T = 40 ºC and feed water T = 20 ºC).

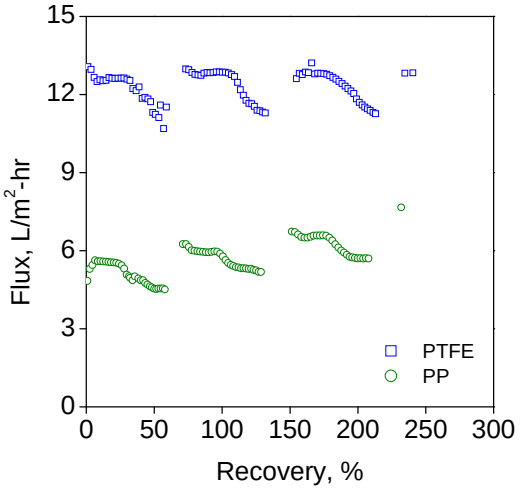

Figure 14. Water flux as function of water recovery for DCMD experiments conducted at CSM. Experiments were conducted with O-BD feed water with a ΔT of 20 °C (permeate T = 40 ºC and feed water T = 20 ºC).

Forward Osmosis

Bench-Scale Testing

During FO bench-scale experiments, the DS was maintained at a constant concentration of 50±2 g/L NaCl. The NaCl DS concentration was high enough to produce initial water fluxes comparable to those achieved using DCMD at a temperature difference of 20 °C and comparable to typical RO operating fluxes for seawater desalination. For the constant feed concentration experiments, the feed solution was maintained at a constant concentration to determine how the water would scale or foul the membrane. Water flux as a function of time for the constant feed concentration FO experiment using both O-BD and SBH-BD is shown in Figure 15. Both feed waters showed very similar results. The small increases in flux throughout the experiments occurred when DDW was added to the DDW reservoir. The general flux trend for both experiments was maintained for 75 hr, but no obvious declines. Because no flux declines were observed, it was determined that the feed waters have minimal scaling and fouling potential. Because the experiments had no flux decline due to fouling or scaling, it was important to operate the system with increasing feed concentration to determine the loss of flux due to a loss in driving force.

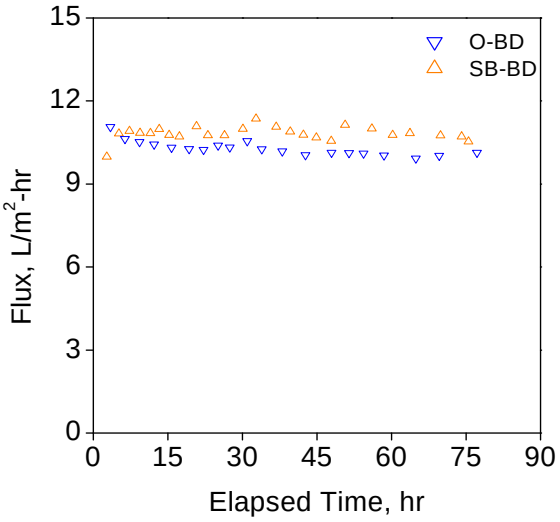

Figure 15. Water flux as function of elapsed time for continuous FO experiments. Experiments were conducted with either O-BD or SBH-BD feed water with a temperature of 25±2 °C with a DS concentration of 50±2 g/L.

Experiments were performed using both O-BD and SBH-BD for the increasing feed concentration experiments. The experiments were terminated when water flux reached approximately 3 L/m^2-hr (Figure 16). As water diffused through the membrane, the concentration of the feed solution, and thus its osmotic pressure, slowly increased. At the same time, the osmotic pressure difference, and thus, the driving force across the membrane decreased. Fluxes for the two feed waters followed almost identical trends with the main difference being the initial fluxes. The initial flux using O-BD was approximately 11 L/m^2-hr whereas the initial flux using SBH-BD was approximately 15 L/m^2-hr. The higher flux in the SB-BD experiment was due to an initially elevated

operating temperature, which was corrected and is observed as immediate flux decline. After approximately 5 hr, both fluxes are approximately 11-12 L/m^2-hr. Moderate flux decline occurs until approximately 15 hr when more rapid flux decline occurs for both experiments until a final flux of approximately 3 L/m^2-hr is reached and the experiment is terminated. Unlike the pressure driven and MD experiments, the flux decline observed for FO is due to the loss in driving force. As operation continues, the feed solution continues to become concentrated so the difference in osmotic pressure continuously decreases, thus the driving force decreases. This is evident by the constant feed concentration experiments (Figure 15) that showed no flux decline due to scaling or fouling.

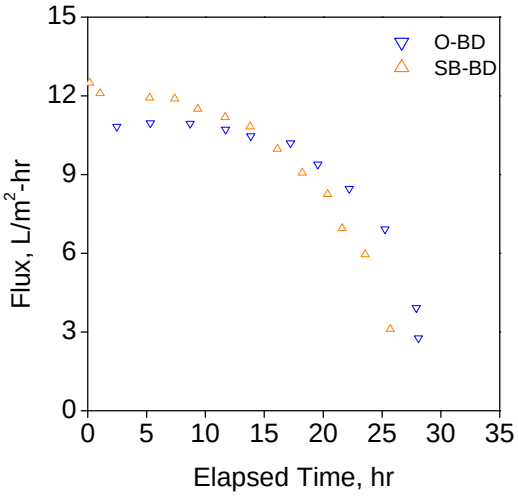

Figure 16. Water flux as function of elapsed time for concentration FO experiments. Experiments were conducted with either O-BD or SBH-BD feed water with a temperature of 25±2 °C with a DS concentration of 50±2 g/L.

Comparison of Results from Bench-Scale Testing

A comparison of results from NF, DCMD, and FO experiments is shown in Figure 17 for the SBH-BD feed water. The initial flux of the NF process is nearly double the initial fluxes of the DCMD and FO processes. The NF membrane flux begins declining immediately which indicates a much higher rate of fouling and scaling than the other membrane processes. This is expected due to the applied pressure. However, it is important to note that the initial flux of the NF process can be adjusted to increase or decrease the flux, which in turn will increase or decrease the rate at which membrane fouling/scaling occurs. This is an important point because if the applied pressure for the NF system is reduced to reduce the flux, the membrane fouling may occur at a slower rate, making it more comparable to DCMD or FO. DCMD does not exhibit flux decline over the 15 hr of the experiment. FO constant feed concentration experiments exhibit no flux decline for approximately 80 hr indicating that the feed water has little to no fouling or scaling potential. The FO increasing feed concentration experiments exhibit flux decline after approximately 17 hr of operation; however, this flux decline is not due to membrane fouling, but due to the concentration of the feed water. As the feed water becomes more concentrated throughout the experiment, the osmotic pressure difference, and hence, the driving force decrease.

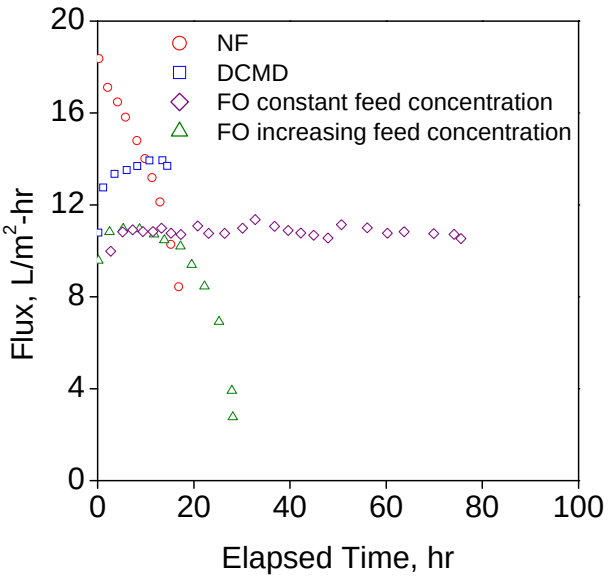

Figure 17. Comparison of results from all bench-scale membrane processes using the SBH-BD feed water.

Membrane rejection was monitored during all of the experiments. During the NF experiment, the permeate conductivity did not exceed 109 µS/cm. During the DCMD experiment, the permeate conductivity did not exceed 4 µS/cm. Because FO does not produce a permeate stream, permeate conductivity was not able to be evaluated. However, high rejection was evident during the FO concentration experiments because precipitates did form in the feed reservoir before the experiment was terminated and the final feed water conductivity was approximately 32 mS/cm.

Based on these results, it is apparent that DCMD and FO provide viable alternatives to pressure driven processes for treatment of cooling tower waters. Both

DCMD and FO maintain high rejection with minimal flux decline. However, selecting the most appropriate system for pilot-scale testing, additional parameters were considered. DCMD is not a logical choice for treatment of cooling tower water because cooling tower water is typically at 25 ºC, and thus does not provide a waste heat. However, it should be pointed out that DCMD may be a suitable process for treatment of geothermal return brines which have temperatures greater than 100 °C. FO was also not chosen because it is not commercially available for pilot-scale testing. Thus, based on the lack of waste heat and commercial availability, NF was chosen for pilot testing.

Nanofiltration Pilot-Scale Testing

Figure 18 shows specific flux as a function of elapsed time for the pilot-scale testing of the NF90 membrane. Specific flux accounts for changes in pressure during operation. For pilot testing, the permeate flow, and thus flux was held constant at all times by increasing the applied pressure, so any increases or decreases observed throughout testing are strictly due to a pressure adjustment. In stage 1 (30% recovery), specific flux generally decreased with time. The sharp initial flux decline over the first 10 hr is most likely due to compaction of the membranes. The membranes were virgin membranes, so operation caused compaction and a slight flux decline. After the first membrane cleaning, the flux recovered to approximately 61% of the original flux. Thus, the membrane began stage 2 with irreversible fouling. In stage 2, system operation continued at 30% recovery. Again, stage 2 follows a general trend of flux decline with slight increases and decreases throughout. It is important to note that when pilot testing

took place, the geothermal facility where testing took place was testing alternate methods for additional makeup water. Because of this, operating conditions and makeup water quality was constantly changing. Many of the increases and decreases throughout experimentation were caused by these changes. Membrane cleaning was performed after the second 100 hr of operation. Cleaning was not effective for the fouling that occurred during stage 2. The flux after cleaning was lower than the final flux for stage 2 (approximately 21% lower). Overall flux decline in stage 2 was 16%; this was less than stage 1 flux decline because stage 1 used a virgin membrane. Operating conditions were adjusted to 60% recovery in stage 3. Stage 3 had the most constant flux decline, which could be attributed to more stable operating conditions throughout the geothermal facility. The flux decline during stage 3 was 19%. Compared to stage 2 (with 16% flux decline), the overall flux decline of both stages was similar. However, membrane cleaning after each stage was much different. The flux in stage 2 was unable to be recovered after cleaning while the flux in stage 3 was recovered to just 8% lower than the initial flux of stage 3. Based on these results, the membranes were able to produce high quality water and maintain performance at high recovery just as well as low recovery, and in some cases the membranes performed better at higher recovery.

Prior to the final cleaning for stage 3, the overall flux decline from the beginning to the end of pilot testing was 58%. These results show that membranes can be a very effective treatment for cooling tower brines. If membrane systems are used to treat cooling tower brines, a higher quality makeup water would lead to lower conductivity

and TDS. This would allow for higher operating fluxes and lower pressures because concentration polarization would be decreased.

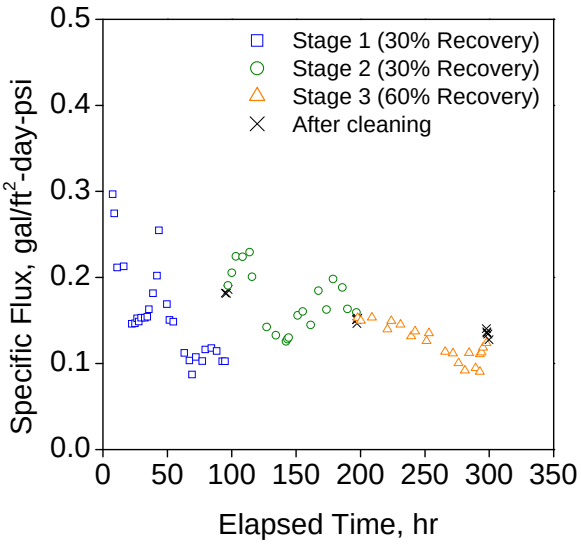

Figure 18. Specific flux as function of elapsed time for pilot-scale testing using the NF90 membrane.

Nanofiltration Process Modeling

Both systems were evaluated assuming a 10-MW facility that operates with 5 cycles in the cooling tower leading to a blowdown flowrate of 690 L/min. This flowrate is used as the feed water flowrate for each modeled system. Specific energy and energy cost as a function of percent water recovery are illustrated in Figure 19 for the first system using SBH-BD and O-BD feed water. Specific energy is the amount of energy (kWh) required to produce a set amount of water (e.g., 1000 L of water) and is the ratio

between power consumption and product water flowrate. Results show the same trends for both feed waters modeled. The highest specific energy is observed at the lowest recovery (55%) and the lowest specific energy is observed at the highest recovery. Once the specific energy requirements are determined, the energy cost per gallon of water produced can also be determined using a typical value of $0.1 per kWh energy [33]. If the specific energy is multiplied by the cost of energy, the cost per gallon of water produced is determined. Results in Figure 19 illustrate that specific cost has a sharp initial decrease from 55-60% recovery and then from 65-80% recovery, maintains a more constant decline. The decrease in specific energy is caused by the decrease in power demand illustrated in Figure 20.

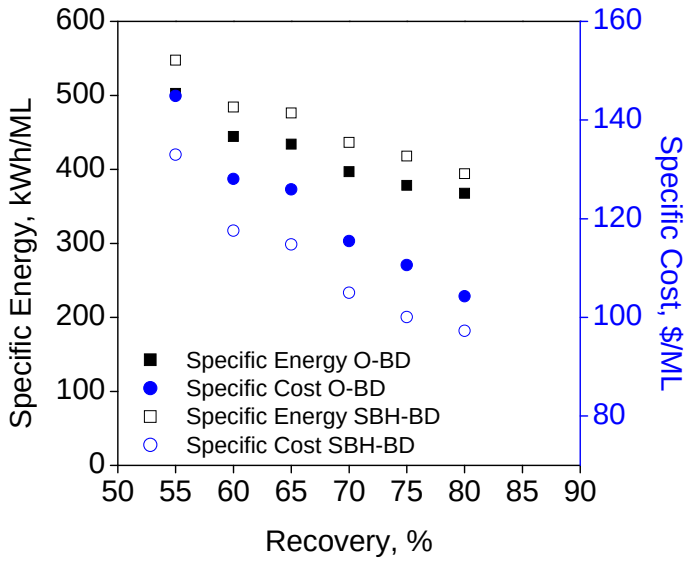

Figure 19. Specific energy as function of recovery for system 1 of the NF model system treating SBH-BD and O-BD feed water. The specific energy is based on the energy needed to treat 690 L/min of cooling tower blowdown at 55, 60, 65, 70, 75, and 80% recoveries.

Power demand and applied pressure as a function of percent water recovery are illustrated in Figure 20 for both SBH-BD and O-BD feed water. Again, both feed waters exhibit nearly identical results. Power demand for the full-scale system decreased from 55 to 60% recovery and then steadily increased from 60% recovery to 80% recovery. The increase in power demand is due primarily to increasing pressure requirements. The dependence of the power demand on the applied pressure is obvious in the figure because the applied pressure follows the exact trend as the power demand. When comparing the graphs for power demand and applied pressure with specific energy, the large decrease in specific energy from 55 to 60% recovery can be explained by the higher recovery, which means higher permeate flowrate along with much lower applied pressure, but only minimal difference in power demand. The graph also shows a non-linear applied pressure and power demand. This non-linearity is caused by the optimization of both membrane area and recovery. At 55% recovery, each membrane element is operating near its maximum recovery, so for 60% recovery, the system could not be operated, so the number of membrane elements was increased. However, because the system is designed with a variable number of pressure vessels with a variable number of membrane elements, a single membrane element cannot be added. And, since more than one membrane element is added, the individual elements are no longer operating near their maximum recovery, so the applied pressure can be reduced at higher recoveries.

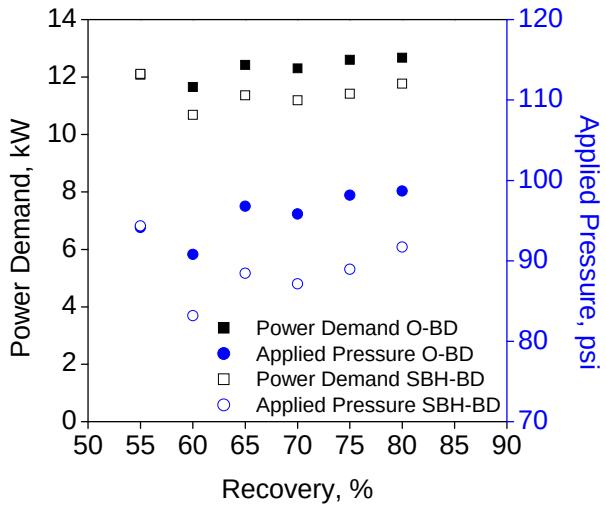

Figure 20. Power demand and applied pressure as function of recovery for system 1 of the modeled NF system treating 690 L/min of SBH-BD and O-BD feed water at 55, 60, 65, 70, 75, and 80% recoveries.

Both the specific energy and power demand for the first system were calculated at a constant feed solution flowrate (690 L/min). In NF, as with other membrane processes, the amount of recovered water depends on the total surface area of membrane that is in contact with the feed stream. This is because water flux through the membrane is limited in order to maintain the integrity of the membrane. For each membrane element used in this model, the maximum recommended recovery is 15%. Therefore, in order to obtain 55, 60, 65, 70, 75, and 80% recoveries and in order to not exceed the manufacturer recommended water flux, the number of membranes in the model was modified by adding or subtracting pressure vessels and/or membrane elements until the desired total recovery was reached. The membrane area and number of membrane elements required to achieve the modeled recoveries are illustrated in Figure 21. The data points were

changed slightly to show that there are four sets of data, but it is important to point out that both membrane area and the number of membrane elements are the same for both feed waters expect at 80% recovery. At lower recoveries (55-65%), less membrane area is required and at higher recoveries (75-80%), significantly more membrane area is required to ensure that the maximum flux requirements are not exceeded. The figure also shows the same membrane area for 60 and 65% recovery (for SBH-BD and O-BD) along with 75 and 80% recovery (for O-BD). This is again due to optimization of both membrane area and recovery. Each recovery was designed to use as little membranes as possible. When more than one membrane element is added, the individual elements are no longer operating near their maximum recovery, and the system can now be operated at both the lower recovery (60 and 75%) and the higher recovery (65 and 80%). This also explains why the O-BD had a higher number of membrane elements at 80% recovery. The O-BD is a more challenging water with more potential for fouling and scaling, so an extra pressure vessel was added to the system to ensure manufacturer recommended operation.

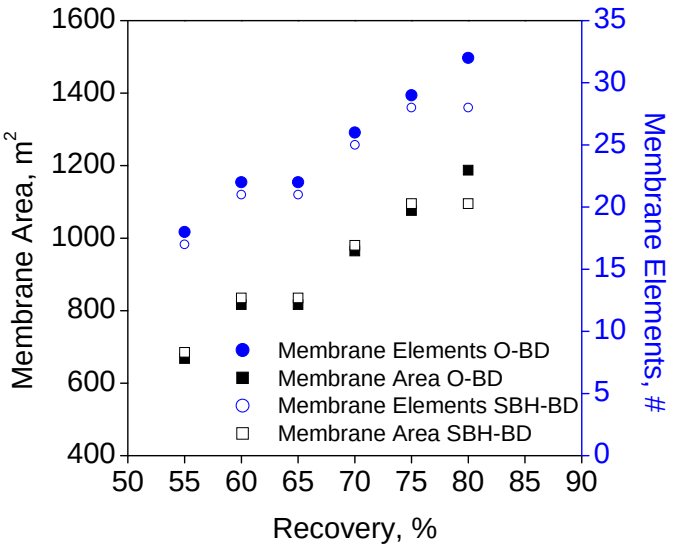

Figure 21. Membrane area and number of membrane elements (DOW Filmtec, NF90-400) as function of recovery for system 1 of the modeled NF system treating SBH-BD and O-BD feed water. The membrane area is based on the area needed to treat 690 L/min of blowdown at 55, 60, 65, 70, 75, and 80% recoveries.

When determining the most appropriate membrane for full-scale application, it is important to choose the system that, in addition to having low specific energy requirements, maintains high salt rejection. TDS rejection for system 1 using both feed waters as a function of water recovery is illustrated in Figure 22. SBH-BD maintains higher rejection for all recoveries with 96% rejection being achieved at 80% recovery. Even though the SBH-BD has higher rejection than O-BD, O-BD maintains rejection greater than 93% for each recovery. These results indicate that even with variable water qualities, system operation is attainable and high quality water is produced.

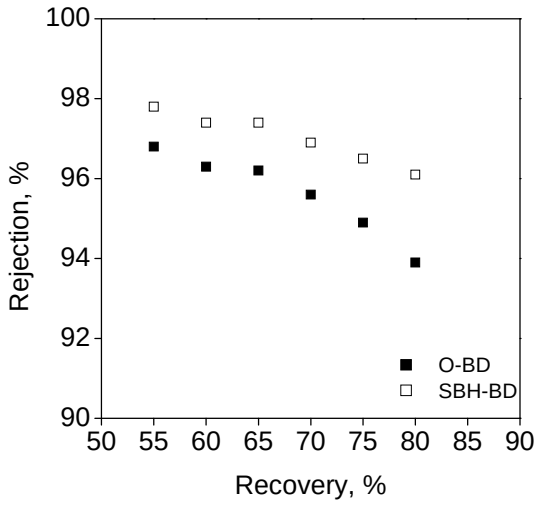

Figure 22. TDS rejection as function of recovery for system 1 of the modeled NF system treating 690 L/min of SBH-BD and O-BD feed water at 55, 60, 65, 70, 75, and 80% recoveries.

System 2 was designed using the 80% recovery design for the O-BD feed water. The system consisted of two stages. Stage 1 had five pressure vessels with four elements in each vessel. Stage 2 had three pressure vessels with four elements in each vessel. Because the system design did not change, the model membrane area is approximately 1200 m^2 (32 elements). Again the specific energy was calculated along with the cost of energy per amount of water produced. The results in Figure 23 are much different than the results observed for both feed waters using system 1. Unlike system 1, the decline in specific energy does not exhibit a rapid decline. This is because the system is not being optimized. However, the lowest specific energy observed using system 1 was

approximately 0.4 Wh/L and using system 2, that value (0.4 Wh/L) is the approximate value for each recovery.

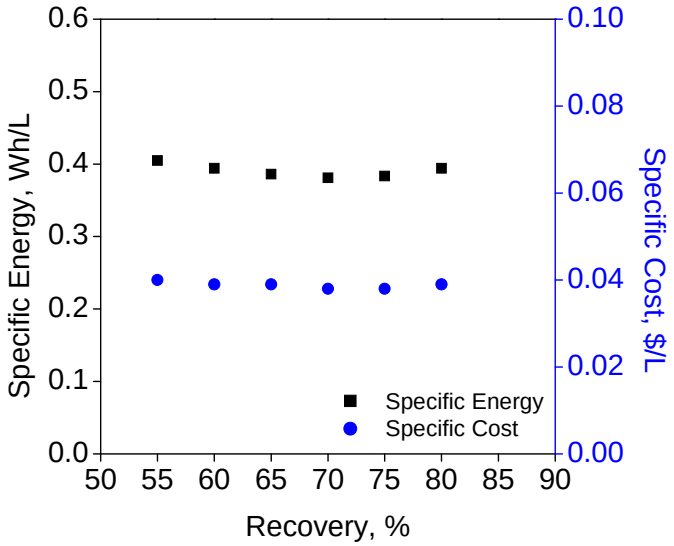

Figure 23. Specific energy and energy cost as function of recovery for system 2 of the modeled NF system treating O-BD feed water. The specific energy is based on the energy needed to treat 690 L/min of geothermal blowdown at 55, 60, 65, 70, 75, and 80% recoveries using the system designed for 80% recovery.

The power demand was also evaluated along with the applied pressure. Again, results in Figure 24 shows much different results that those observed using system 1. The power demand along with the applied pressure increase linearly for each recovery, but again, the initial power demands along with the applied pressure are much lower than those required for system 1. For system 2, the lowest operating pressure is approximately

70 psi which is much less than the approximately 90 psi observed for system 1. The power demands are also lower throughout operation.

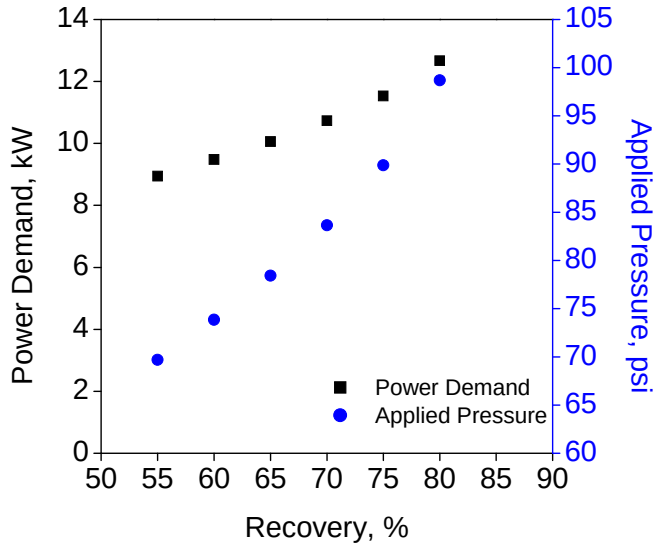

Figure 24. Power demand and applied pressure as a function of recovery for system 2 of the modeled NF system treating 690 L/min of O-BD feed water at 55, 60, 65, 70, 75, and 80% recoveries using the system designed for 80% recovery.

Finally, rejection was calculated. Results in Figure 25 illustrate system 2 does not exhibit the high rejection achieved for system 1, which were each greater than 96% at 55% recovery for each feed water. Here the highest rejection observed is approximately 95%. This loss of rejection observed for system 2 is caused by a decrease in flux per membrane. System 1 was designed so that as recovery changed, the membrane requirements changed. Because of this, nearly the same flux was achieved for each

membrane at each recovery. This model however, has a decrease in flux for each membrane as the recoveries decrease which leads to lower rejection.

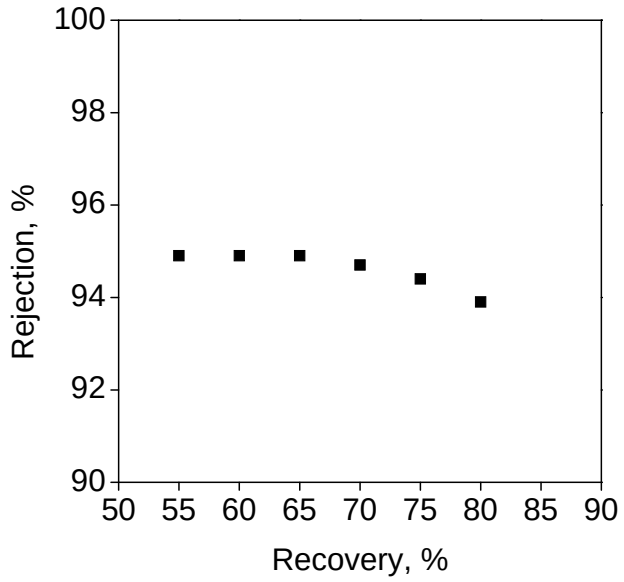

Figure 25. TDS rejection as function of water recovery for system 2 of the modeled NF system treating 690 L/min using O-BD feed water at 55, 60, 65, 70, 75, and 80% recoveries using the system designed for 80% recovery.

Both systems operated at the same recoveries, so the amount of water produced per day was the same for both models and is illustrated as a function of recovery in Figure 26.

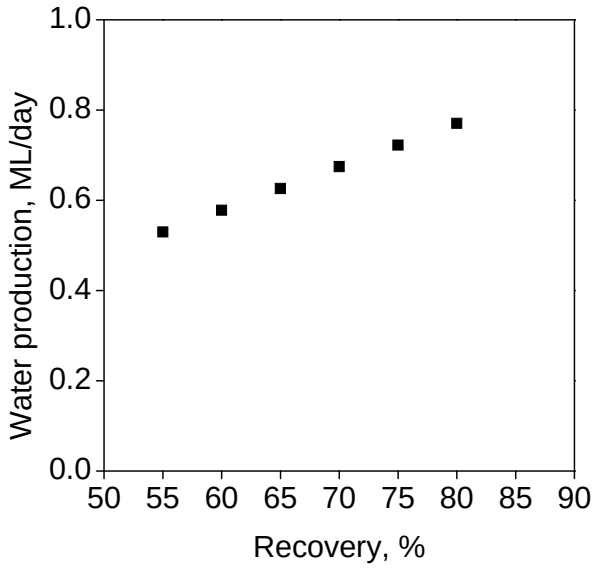

Figure 26. Makeup water produced (ML/day) as function of percent recovery using the modeled NF system at 55, 60, 65, 70, 75, and 80% recovery.

As expected, the amount of water produced per day increases linearly with increasing recovery. Depending on the desired recovery, the cost of an alternate makeup water source for geothermal facilities could be substantially reduced. For example, using the 10-MW facility described earlier, the amount of makeup water required is 3450 L/min (4.95 ML/day) and 690 L/min (1 ML/day) of blowdown is produced. If a membrane system was used to treat the blowdown water with a system that operated at 70% recovery, approximately 482 L/min (0.7 ML/day) is produced. This 482 L/min would be recycled back to the system as makeup water reducing the need for alternate makeup water from 3450 L/min to approximately 3000 L/min. This is close to 14% reduction in

water purchase from the source. Figure 27 is a schematic of the cooling tower with the membrane system for treatment of the geothermal blowdown brine. The permeate produced is being reused as makeup water in the cooling tower. This is very important economically. Currently, geothermal power plants in Southern California are paying approximately $85/1.2 ML of water. Thus, the cost of water for a 10-MW power facility that uses makeup water at a rate of 3450 L/min (4.95 ML/day) is approximately $10,000 per month. If the amount of water required is reduced by 14%, the saving amounts to $1,430 per month. For a 100-MW power facility, the savings will amount to more than $170,000 per year. However, system operating costs make this system economically inefficient. Operating at 70% recovery, the system costs would be approximately $2,500 per month, making system operating costs more expensive than the cost to produce the water. However, recently water prices are rising as water becomes more scarce and thus, a more valuable resource. From 2002 to 2007, water prices in the United States on average rose 27% [34]. As of July 1 2008, an annual study showed that water prices in the United States had increased on average 7.3% [35]. If water prices continue to increase, and the cost of water for the power facility is raised to $200/ML, the saving in water purchasing for a 100-MW power facility will amount to more than $400,000 per year!

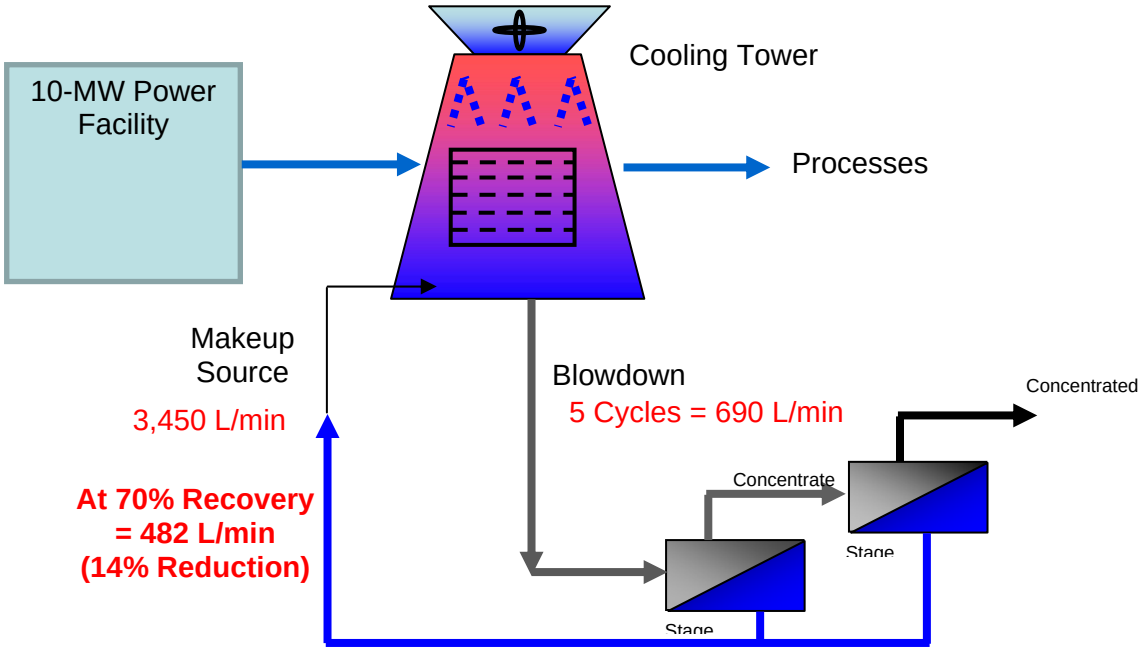

Figure 27. Schematic of cooling tower with the membrane system for treatment of blowdown brine. The permeate water is being reused as makeup water in the cooling tower.

Conclusions

In this study, it was found that NF, DCMD, and FO are effective treatments for geothermal brines. Because of the high turbidity and SDI of the blowdown water, pretreatment was determined to be a necessary step. For bench-scale testing, MF was used to reduce the amount of suspended matter being fed to the membrane system. All systems were able to operate at relatively high fluxes (<10 L/m^2-hr). Both NF and DCMD produced water with very low conductivity indicating high rejection throughout the duration of the experiment. Chemical cleaning was an effective method for

recovering fluxes for both NF and DCMD. DCMD was able to operate with limited flux decline. However implementation at full- or pilot-scale would require waste heat in order to be economically feasible. It is believed that DCMD would be a logical and effective choice for treating superheated geothermal brines. FO experiments indicated that both feed waters had limited fouling or scaling potential when operating at a constant feed concentration. The system was operated for approximately 75 hr with no noticeable flux decline. However, when increasing feed concentration experiments were performed, noticeable flux decline occurred as the feed solution became more concentrated. But, unlike other experiments, the loss in flux was not due to fouling or scaling, it was due to a loss in driving force. With more testing, the feasibility for larger-scale operation can be determined. Pilot testing for the NF system showed that high quality water can be produced at high recovery (60%). The system operated for 100 hr with flux decline that was able to be partially recovered after cleaning. Currently, NF is best suited for immediate implementation due to its commercial availability.

Cost modeling also showed the economic advantages to membrane treatment of blowdown brines. Currently, system costs are more expensive than the water savings, however, if water prices continue to rise, the water saving costs will far outweigh the system operation costs.

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