





# Continuous Flow Seawater RO System for Recovery of Silica-Saturated RO Concentrate

# WateReuse Research Foundation

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### About the WateReuse Research Foundation

The mission of the WateReuse Research Foundation is to conduct and promote applied research on the reclamation, recycling, reuse, and desalination of water. The Foundation's research advances the science of water reuse and supports communities across the United States and abroad in their efforts to create new sources of high-quality water through reclamation, recycling, reuse, and desalination while protecting public health and the environment.

The Foundation sponsors research on all aspects of water reuse, including emerging chemical contaminants, microbiological agents, treatment technologies, salinity management and desalination, public perception and acceptance, economics, and marketing. The Foundation's research informs the public of the safety of reclaimed water and provides water professionals with the tools and knowledge to meet their commitment of increasing reliability and quality.

The Foundation's funding partners include the Bureau of Reclamation, the California State Water Resources Control Board, the California Energy Commission, and the California Department of Water Resources. Funding is also provided by the Foundation's Subscribers, water and wastewater agencies, and other interested organizations.

# Continuous Flow Seawater RO System for Recovery of Silica-Saturated RO Concentrate

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### Acronyms

AFD	adjustable frequency drive
CHIWAWA	Consortium for High Technology Investigations of Water and Wastewater
EPWU	El Paso Water Utilities
GE	General Electric
gfd	gallons per square foot per day (transmembrane flux)
gpm	gallons per minute (flow)
HERO	high-efficiency reverse osmosis
KBH	Kay Bailey Hutchison
kgal	thousands of gallons (liquid volume)
µS/cm	microsiemens per centimeter (electrical conductivity)
mg/L	milligrams per liter (concentration)
OPUS	optimized pretreatment and unique separation
psi	pounds per square inch
RO	reverse osmosis
SWRO	seawater reverse osmosis
TDS	total dissolved solids
TWDB	Texas Water Development Board
VSEP	vibratory shear enhanced processing

### Foreword

The WateReuse Research Foundation, a nonprofit corporation, sponsors research that advances the science of water reclamation, recycling, reuse, and desalination. The Foundation funds projects that meet the water reuse and desalination research needs of water and wastewater agencies and the public. The goal of the Foundation's research is to ensure that water reuse and desalination projects provide high-quality water, protect public health, and improve the environment.

An Operating Plan guides the Foundation's research program. Under the plan, a research agenda of high-priority topics is maintained. The agenda is developed in cooperation with the water reuse and desalination communities, including water professionals, academics, and Foundation subscribers. The Foundation's research focuses on a broad range of water reuse research topics including

- Definition and addressing of emerging contaminants
- Public perceptions of the benefits and risks of water reuse
- Management practices related to indirect potable reuse
- Groundwater recharge and aquifer storage and recovery
- Evaluation and methods for managing salinity and desalination
- Economics and marketing of water reuse

The Operating Plan outlines the role of the Foundation's Research Advisory Committee (RAC), Project Advisory Committees (PACs), and Foundation staff. The RAC sets priorities, recommends projects for funding, and provides advice and recommendations on the Foundation's research agenda and other related efforts. PACs are convened for each project and provide technical review and oversight. The Foundation's RAC and PACs consist of experts in their fields and provide the Foundation with an independent review, which ensures the credibility of the Foundation's research results. The Foundation's Project Managers facilitate the efforts of the RAC and PACs and provide overall management of projects.

In a prior research study that was funded by the Texas Water Development Board and El Paso Water Utilities, it was shown that a batch-treatment seawater reverse osmosis system (SWRO) can recover more than 85% of the water from the silica-saturated RO concentrate that is generated at the Kay Bailey Hutchison (KBH) desalting plant in El Paso, TX. Final feed water total-dissolved-solids concentrations as high as 75,000 mg/L were routinely achieved. In this project, the pilot plant was converted from batch treatment to continuous flow. The results showed that calcium sulfate precipitated in the feed tank at all recoveries exceeding 35%, fouling the SWRO membrane. At the outset of the project, it was believed that silica fouling would be the most challenging issue to resolve. However, the only membrane fouling that was observed throughout testing was due to calcium sulfate.

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We also thank Dr. Bob Ning of King Lee Technologies for providing scale inhibitors and cleaning solutions and for always being available to discuss new ideas. Frank Leitz of the U.S. Bureau of Reclamation provided valuable technical review to the TWDB.

Finally, we thank the project advisory committee for offering its insightful suggestions about how to overcome the obstacles that we encountered along the way. The members of the committee were as follows:

Patrick Brady, Sandia National Laboratories Juan Gomez, CH2M HILL, Inc. Scott Irvine, Bureau of Reclamation Keith Rutherford, Parkhill, Smith & Cooper, Inc. Bruce Thomson, University of New Mexico In a previous research study that was funded by the Texas Water Development Board (TWDB) and El Paso Water Utilities (EPWU), it was shown that a batch-treatment seawater reverse osmosis (SWRO) system can recover more than 85% of the water from the silicasaturated RO concentrate that is generated at the KBH desalting plant in El Paso, TX. Final feed water total dissolved solids (TDS) concentrations as high as 75,000 mg/L were routinely achieved. In this project, the pilot plant was converted from batch treatment to continuous flow. Concentrate from the KBH plant flows continuously into a feed tank that also receives the recycled concentrate from the SWRO system. A bleed valve at the bottom of the feed tank opens and closes to keep the water in the feed tank at a preset conductivity, which establishes the recovery of the system. A scale inhibitor for sulfate control was added at 5 mg/L, and the pH in the feed tank was maintained at 3.9 through the addition of sulfuric acid. Sulfuric acid was selected because of its lower cost.

The results showed that calcium sulfate precipitated in the feed tank at all recoveries above 35%, fouling the SWRO membrane. By comparison, a 35% recovery corresponds to a feed water TDS concentration of only approximately 15,000 mg/L. Various attempts to prevent precipitation were unsuccessful, including increasing the scale inhibitor concentration, adding a heat exchanger to the feed tank to control the temperature, decreasing the permeate flux, and reducing the volume of water in the feed tank. It appears that the system design was faulty, because the feed tank was maintained at TDS levels such that one or more substances exhibited supersaturated concentrations continuously. The problem of calcium sulfate precipitation might be avoided by designing a single-pass continuous flow system wherein the extreme supersaturated condition exists for only a short period before the concentrate is discharged from the membrane (namely, the treatment time must be shorter than the induction period for calcium sulfate precipitation).

At the outset of the project, it was believed that silica fouling would be the most challenging issue to resolve. However, the only membrane fouling that was observed throughout testing was due to calcium sulfate, and the cleaning solution Diamite CAL was effective in removing it from the fouled SWRO membrane. Silica precipitation was never a problem during this pilot study.

### Chapter 1 Introduction and Background

One of the biggest obstacles to inland desalination projects is related to disposal of the large amount of concentrate that is generated in membrane desalting operations. Although there is a fairly open portfolio of options for concentrate disposal, including mechanical evaporation, enhanced evaporation, concentrate management through wetlands, beneficial reuse of concentrate, deep well injection and traditional evaporation ponds, each of them has limitations of its own. At the present time, economic and other considerations generally limit the concentrate management options for large-capacity, inland desalination plants to injection and evaporation is not reasonably close to the desalting plant and if the injection zone must be situated at extreme depths, such as more than 10,000 ft deep. In the case of El Paso's KBH Desalting Plant, the proper formation is not feasible in areas that receive substantial rainfall, but in the southwestern United States, the arid climate is ideal for evaporation.

Because the land area required for evaporation is a function of the concentrate volume that must be handled, the smaller the volume, the more attractive the evaporation option becomes. Costs for the evaporation pond option are driven by such items as the costs for land and the need for a double liner and a leak detection system, in addition to the area being driven by net evaporation rates in the region in question and the eventual costs associated with removal from the evaporation ponds and disposal of the dry solid residuals in a permanent landfill. Use of ponds can also bring up the issue of potential contamination to wildlife, because water attracts animals. For example, the El Paso area serves as corridors for migrating populations of birds. Furthermore, the extra water that is recovered from the concentrate not only decreases the volume requiring final disposal but also increases the amount of water available for beneficial use, which is generally the reason that the desalination plant was built in the first place. The project described in this report was undertaken for the dual purpose of recovering additional water from the concentrate of a reverse osmosis (RO) desalting plant while coincidentally reducing the volume of concentrate that requires final disposal.

El Paso Water Utilities (EPWU), in partnership with Fort Bliss (a U.S. Army installation), owns and operates the largest inland brackish groundwater RO desalting plant in the United States. The KBH desalting plant produces 27.5 million gal of drinking water per day at full capacity. The plant operates at 80% recovery and disposes of the high-silica concentrate in injection wells situated 22 mi from the plant.

In 2001, EPWU initiated research directed toward removing and/or controlling silica during membrane desalting of brackish groundwater in order to be able to recover additional water from its RO concentrate. With funding from the Bureau of Reclamation in 2002 and 2004, studies were conducted using lime precipitation followed by RO and nanofiltration followed by RO (Tarquin, 2005 and 2006). Recoveries of more than 50% were achieved with the sequential membrane treatment, and much higher recoveries were deemed possible following silica removal with lime (Ning and Tarquin, 2010). Nevertheless, many water utilities avoid using lime treatment because of scaling- and sludge (residual)-handling problems, so alternative concentrate recovery methods were pursued. In 2007, the Texas Water

Development Board (TWDB) funded a project to investigate using vibratory shear enhanced processing (VSEP) and seawater RO (SWRO) batch-treatment systems for recovering water from the KBH concentrate. Both systems were able to achieve recoveries of more than 85%, but the SWRO system was more economically attractive than the VSEP system, so VSEP testing was discontinued.

The initial, or prior, SWRO tests were carried out in a batch-treatment mode using a Crane SWRO unit with a 2.5 in. membrane (Ning, Tarquin, and Balliew, 2010). In 2009, the TWDB amended its contract with EPWU to provide for construction of a fully automated batch-treatment SWRO pilot plant with four 4-in. membrane elements in parallel operating at a constant pressure of 700 psi. The batch-treatment pilot plant was operated for 6 months at concentrate recoveries in the 85% range (with silica concentrations reaching 1000 mg/L) without fouling of the membranes and at a unit water cost comparable to that of the main desalting plant (Tarquin, 2010). When recovery of KBH concentrate was increased to about 90%, precipitation of calcium sulfate occurred, so that mineral became the constituent of concern (rather than silica) for recovering additional water from the KBH concentrate.

A number of techniques have been studied and tested for preventing calcium sulfate scaling in membrane systems. Water softening (via chemical treatment and/or ion exchange) has been used for many years and is an important step in both the high-efficiency reverse osmosis and optimized pretreatment and unique separation technology processes for high product recovery in RO systems. Corbett et al. (2003) evaluated electromagnetic technology for calcium sulfate control and concluded that neither a magnetic device nor a high-voltage capacitance device was effective in preventing calcium sulfate scaling at 91% water recovery. However, the addition of 2 mg of sodium hexametaphosphate/L to the RO feed water was successful in preventing scale at recoveries of 93%. The addition of proprietary scale inhibitors is perhaps the most common method of calcium sulfate control in RO systems. Scale inhibitors interfere with precipitation reactions through threshold inhibition (keeping sparingly soluble salts in solution), crystal modification (interrupting the electric balance that is necessary for crystal growth), or dispersion (imparting anionic charges on crystals to keep them separated) (Avista Technologies, 2008).

In evaluating different types of scale inhibitors, Amjad (1985) concluded that formulated polyelectrolytes were the most effective, with the induction period affected by the scale inhibitor's molecular weight, its concentration, and the nature of the functional groups. Sarig and Mullin (1982) reported that induction periods for the precipitation of CaSO<sub>4</sub>·2 H<sub>2</sub>O were insensitive to calcium sulfate concentration, suggesting that the nucleation process was heterogeneous. Lancia et al. (1999) found that the induction period for homogeneous nucleation of CaSO<sub>4</sub>·2 H<sub>2</sub>O (gypsum) decreased when either temperature or supersaturation increased. Alimi et al. (2003) found that the induction time was strongly dependent on the solution supersaturation and the temperature, with the activation energy decreasing with increasing supersaturation and temperature. Shih et al. (2005) found that gypsum scale development was affected by the formation of crystals on the membrane surface and suggested that research is needed on the impact of surface topology and chemistry on surface crystallization of mineral salts. A number of studies have demonstrated that gypsum scaling of membranes is controlled by both surface/heterogeneous crystallization and by deposition of bulk materials (Mi and Elimelech, 2010). Mi and Elimelech found that gypsum scaling on a polyamide membrane was dominated by heterogeneous/surface crystallization whereas gypsum scaling of a cellulose acetate membrane was dominated by bulk crystallization and subsequent particle deposition. Pomerantz et al. (2006) had success in preventing calcium sulfate scaling by reversing the flow to RO process trains, thereby replacing the

supersaturated brine in the last membrane element with unsaturated feed flow before the induction time was reached. A small-scale unit was operated for 22 h under reverse flow conditions with a calcium saturation index of 5.4 without fouling the last element.

The success of the batch method for SWRO pilot testing naturally leads to the investigation of the continuous flow method, which should theoretically be much simpler to automate and operate. This project involved constructing and evaluating an automated, continuous flow SWRO system for recovering a high percentage of the water from the silica-saturated RO concentrate.

### Chapter 2

### **Concentrate Characterization**

Concentrate from KBH served as the raw water feed in this project. The overall recovery of KBH is 80%. The characteristics of the KBH concentrate are shown in Table 2.1. Notable characteristics are total dissolved solids (TDS) at 12,763 mg/L, silica at 145 mg/L, calcium at 722 mg/L, and sulfates at 1410 mg/L as expressed in the column marked Avg under Data for 2010. As shown in the % Change columns of the average values, all parameters related to salt content in the water have continued to increase during the past 2 years. This increase stemmed from the general degradation of the raw water supply over time caused by brackish water intrusion. In addition, several new source water wells were drilled to replace old wells during construction of KBH. (Several of the existing wells had either collapsed or were producing excessive amounts of sand.) In order to allow for the draw-down required during pumping at the high rates necessary to supply KBH, these new wells were completed at a greater depth than the original wells. This condition in turn resulted in tapping into the higher-TDS waters encountered in the lower portion of the aquifer.

		Data for 2008			Data for 2009		% Change <sup>b</sup>	Ι	Data for 201	.0	% Change <sup>b</sup>
<b>Parameter</b> <sup>a</sup>	Min	Avg	Max	Min	Avg	Max	2008-2009	Min	Avg	Max	2009–2010
Cl <sup>-1</sup>	265	4699	9710	2590	5439	6740	16%	4040	6099	10,100	12%
SO4 <sup>-2</sup>	127	1039	2110	1020	1410	2260	36%	1020	1410	2260	0%
ALK-T	418.8	417	498	400	427	445	3%	130	397	477	-7%
EC	10,400	16,267	22,100	8740	18,712	21,500	15%	10,700	20,565	28,800	10%
Fe-T	0.03	0.13	0.57	0.032	0.121	0.42	-7%	0.06	0.16	0.41	32%
Hard-T	528	2089	3030	1200	2328	3050	11%	540	2490	4330	7%
Mn <sup>+2</sup>	0.1	0.16	0.21	0.08	0.18	0.22	11%	0.14	0.21	0.34	17%
PO <sub>4</sub> <sup>-3</sup>	0.1	0.16	1.26	0.1	0.16	0.49	1%	0.1	0.17	0.96	6%
pН	7.1	8.0	8.3	7.5	7.8	8.0	-2%	7.6	8.0	8.0	3%
Ca <sup>+2</sup>	376	589	793	281	640	937	9%	176	722	1220	13%
K <sup>+1</sup>	4.9	76	99.7	38.9	77	114	18%	61	78	89	-12%
$Mg^{+2}$	0.9	153	208	85.7	171	250	12%	146	191	260	12%
Na <sup>+1</sup>	172	2674	4200	1490	3097	4140	16%	2710	3412	4440	10%
SiO <sub>2</sub>	28.7	148	228	23.4	128	173	-14%	36.7	145	187	13%
TDS	6740	10,412	13,200	5730	11,520	13,600	11%	9750	12,763	17,900	11%
Тетр	18.9	22.0	26.0	22.6	25.0	26.3	4%	22.1	26.0	26.6	4%

Table 2.1. Characteristics of KBH Concentrate

<sup>*a*</sup>All values are mg/L except EC (µS/cm), pH (pH units), and Temp (°C).

<sup>b</sup>% change is based on average values. ALK-T refers to Total Alkalinity; Fe-T to total iron concentration; hard-T to total water hardness.

During the previous pilot testing described in Chapter 1, seven scale inhibitors were tested for sulfate control during treatment of KBH concentrate in the SWRO unit. Of these seven products, the proprietary scale inhibitor Pretreat Plus 0400 from King Lee Technologies provided the best results. It should be pointed out that Pretreat Plus Silica is added to the feed water to the KBH plant at a dosage of 4 mg/L, which is prescribed by the manufacturer. Therefore, the concentration of that scale inhibitor in the KBH concentrate should be approximately 20 mg/L on the basis of an average KBH recovery of 80%.

### Chapter 3

### **Experimental Setup and Research Methods**

A simplified schematic diagram of the experimental setup is shown in Figure 3.1. Concentrate from the KBH desalting plant continuously flows into the 200-gal feed tank through a float valve. A scale inhibitor for controlling sulfates (Pretreat Plus 400) is added at a preset concentration based on the rate of flow into the feed tank (namely, on a flow-paced basis). At the start of the project, the concentration was set at 4 mg/L.



Figure 3.1. Schematic diagram of pilot plant setup.

Sulfuric acid addition was automatically controlled by feeding a 25% sulfuric acid solution to maintain the pH at a preset value, 3.9 in this study, to ensure that there would be no fouling due to carbonates. In addition to the pH probe, the feed tank was equipped with a conductivity probe that enabled the feed tank to be maintained at any preset value by opening and closing the bleed valve as necessary. The pressure vessel housed a single GE-Osmonics SWRO thin-film membrane Model AD 4040FM. The active area of the membrane was 86 sq ft (sf) with an average NaCl rejection capacity of 99.6% (under standard conditions as tested by the manufacturer). The operating parameter of maximum pressure drop over a single element is 12 psi. A schematic diagram of the membrane is shown in Figure 3.2, with A, membrane length, at 40 in., B, diameter of the permeate tube, at 0.75 in., and C, outside diameter of the membrane, 3.9 in.



**Figure 3.2. Schematic of SWRO membrane.** Source: Fact Sheet, AD Series, Seawater RO High Rejection www.gewater.com/products/consumables/pure\_water\_elements/index.jsp

The high-pressure feed pump was a Cat Pumps Model 820 triplex positive displacement pump that is driven by a 7.5 hp motor. A flow meter in the permeate discharge line allowed for the system to be operated in a constant-permeate-flow mode (namely, the speed of the pump was automatically adjusted to maintain the preset permeate flow rate). At the beginning of the project, the permeate flow rate was set to 0.5 gal per min (gpm), which translated to a membrane flux of 9 gal per sq ft per day (gfd). The treatment unit has continuous data-logging capability. Appendix A of this report exhibits the screen shots (examples) of the system control panel. Data are collected at 1-min intervals whenever the system is running in the automatic mode (namely, during all test runs). The data collected include various flow rates, conductivities, pressures, temperature, and the pH of the feed water. Part of the data sheet from June 24, 2010, is shown in Table 3.2.

All of the analytical results included in this report were obtained at the International Water Quality Laboratory (IWQL) of EPWU. After samples were collected at the research site (namely, the KBH laboratory), a chain-of-custody form was filled out and the samples were transported to the IWQL within 2 h. The quality assurance/quality control procedures associated with the test results are contained in section 23 of its Quality Manual (EPWU, 2011). Additional samples were collected for analysis in the laboratories of the Civil Engineering Department at the University of Texas at El Paso. Those samples were collected and analyzed (using wet chemical techniques of the HACH Chemical Co.) primarily to obtain "instant feedback" about the test run from which they were taken, but they also provided the precipitated solids that were analyzed via X-ray diffraction or electron microscopy to determine the composition of the precipitate. It should be pointed out that the HACH procedure for silica determination measures only reactive silica. Thus, after silica begins to polymerize, the HACH colorimetric procedure will measure only the monomeric silica. Total silica could be determined by using inductively coupled plasma or by digesting the samples prior to analysis, but neither was done in this project, so some mass balances for silica show a silica deficiency in the product waters.

Time Stamp	Feed Flow	Drain Flow	Permeate Flow	Concentrate Flow	Tank Cond	Feed Cond	Permeate Cond	Concentrate Cond	Feed Temp	Feed pH	Feed Pressure	Concentrate Pressure
12:00	0.7	0	0.49	6.84	33339	32429	1333	34530	29.7	3.7	377	375
12:01	0.7	0	0.51	6.79	33211	32820	1321	34188	30.3	3.7	387	364
12:02	0.7	0	0.51	6.79	33571	32478	1368	34676	29.5	3.8	383	365
12:03	0.7	0	0.5	6.87	33468	32845	1358	34334	30.2	3.8	371	371
12:04	0.7	0	0.49	6.86	33417	32429	1331	34383	30.5	3.8	381	372
12:05	0.7	0	0.49	6.84	33339	32747	1333	34554	30.1	3.8	384	369
12:06	0.7	0	0.5	6.86	33622	32503	1338	33968	31.4	3.8	392	369
12:07	0.7	0	0.51	6.86	33185	32600	1363	34407	30.5	3.7	386	366
12:08	0.6	0	0.5	6.83	33725	32723	1363	34383	31.3	3.8	378	370
12:09	0.7	0	0.49	6.87	34033	32869	1365	34432	30.6	3.7	392	369
12:10	0.7	0	0.49	6.83	33417	32723	1358	34554	30.6	3.8	391	371
12:11	0.7	0	0.49	6.87	33211	33235	1355	34994	29.8	3.8	387	371
12:12	0.6	0	0.49	6.89	33931	33284	1350	34652	30.7	3.8	391	378
12:13	0.7	0	0.5	6.83	33828	33113	1387	35189	30.1	3.8	395	375
12:14	0.7	0	0.49	6.84	34085	33577	1375	35042	30.2	3.8	389	377
12:15	0.7	0	0.48	6.89	33648	34017	1350	35531	29.8	3.8	382	377
12:16	0.6	0	0.49	6.88	33699	33284	1350	35628	29.9	3.8	388	374
12:17	0.7	0	0.49	6.9	33828	33675	1358	34627	31.5	3.8	395	379
12:18	0.7	0	0.49	6.94	34085	33333	1331	35067	30.7	3.8	393	382
12:19	0.7	0	0.5	6.88	34291	33650	1355	36166	29	3.7	392	381
12:20	0.6	0	0.49	6.93	34008	33186	1370	35311	30.6	3.8	405	382

 Table 3.2. Sample of Data Sheet (20 Min on June 24, 2010)

Because the primary task of the project was to operate the SWRO system in a continuous flow mode at recoveries beginning at 70% and increasing in 5% increments until membrane fouling occurred, the pilot plant was initially operated at what were believed to be low recoveries (namely, less than 50%) in order to gain familiarity with the vagaries of the system and "work the bugs out." As it turned out, it was not possible to obtain sustainable recoveries anywhere close to what was expected, so most of the testing was directed toward trying to get the recoveries up to at least 50%. Hence, different strategies were tested, such as increasing the scale inhibitor dosage, but none was shown to be successful as discussed later in this report.

In order to clean the membrane after it was fouled, two different cleaning solutions were acquired from King Lee Technologies: High Flux A, which is specific for silica, and Diamite CAL, which is specific for calcium sulfate. The procedure recommended by the supplier was followed in carrying out the cleaning. The first time the membrane was fouled, High Flux A was used, followed by a clean-water test to see if the permeate flow rate was restored to near its original value. If the permeate flow rate after cleaning with High Flux A was deemed to be too low, the membrane was cleaned with Diamite CAL and then retested with clean water. The cleaning solution that was most effective was used thereafter.

### **Equipment Problems Encountered**

#### 4.1 Error in the Feed Readings

A number of problems were encountered in starting up the system, but this was expected because of the complexity of the pilot plant with its sensors, control systems, and datalogging functions. See Appendix A for screen shots of the control panel. For example, the conductivity sensor in the feed tank (George Fisher) was giving highly erratic readings or not functioning at all. After we replaced it with a completely new unit and obtained the same result, we determined that the adjustable frequency drive for the feed pump was emitting electromagnetic radiation that created a voltage and current flow in the feed tank, thereby interfering with the operation of the conductivity meter. Several attempts at grounding various components of the system were unsuccessful, so a meter with a different type of conductivity sensor (toroidal by Cole Parmer) was installed and that problem was eliminated.

### 4.2 Error in Water Level Readings

Next, a problem with the water-level switch was discovered: it indicated that the water level was above the sensor even when the tank was empty. This situation would allow the high-pressure pump to run even when there was no water in the tank, thereby causing it to fail.

The switch was replaced with one specifically intended for use in salt water, but it too failed. Finally, the contractor replaced the switch with a float-type switch, which worked perfectly through the end of the project.

### 4.3 Issues with the Positive Displacement Pump

In operation of the unit intermittently during the shake-down period, an unusual noise was noticed in the positive-displacement pump whenever the pump was stopped at the end of a test run. One of the piston sleeves was subsequently replaced by the vendor, but the problem did not seem to be completely resolved. In the weeks that followed, it became obvious that there was still a problem with the pump, so the manufacturer's representative replaced the other two sleeves, acknowledging that the pump still did not sound right. Another pump was ordered and installed by the supplier, and it worked well through the end of the project.

### 4.4 Problems with Process Control and Data-Logging Systems

At the same time that the pump problems were being resolved, the process control and datalogging systems were not functioning smoothly. After the contractor recalibrated the sensors and made various modifications to the process control software, those components began to function very well. Similarly, software modifications were made that enabled the system to be remotely monitored and controlled, a condition that proved to be extremely valuable after routine operation of the system began. The same is true for a remotely controlled camera that was installed at the KBH lab site, which is situated 13 mi from the University of Texas at El Paso.

#### 4.5 **Problems with Accessories**

Approximately 1 month after the system was initially started, problems were encountered with one of the pH sensors, the high-pressure relief valve, and the low-pressure feed pump. The pH sensor that failed was situated on the suction side of the pump, and it measures the pH of feed water to the membrane. It is also connected to the acid feed pump and is used to control the pH of the feed water. Therefore, it is an important part of the control system and had to be functional at all times. A representative of the manufacturer (George Fisher) looked at the probe and concluded that the electrode had a cracked glass sensor, which was probably defective from the outset, even though there was some vibration where the sensor was situated (because of the pump problems previously discussed) that may have contributed to the sensor failure. After the electrode was replaced, the system functioned normally.

#### 4.6 Problems with the High-Pressure Relief Valve

The high-pressure relief valve is situated on the discharge side of the positive displacement pump and protects the unit from overpressurization. The problem with the high-pressure relief valve was detected by comparing the data-logged inlet flow rate to the sum of the permeate and bleed flow rates. The data showed that the volume of water entering the system was greater than the volume leaving. Inspection of the drain lines revealed that there was flow in the high-pressure relief line, even though the system pressure was nowhere near the pressure that should have activated the valve. When attempts to adjust the relief pressure failed, it was determined that the adjustable spring that controls the relief pressure was defective. Replacement of the relief valve solved the problem.

#### 4.7 Problems with the Feed Pump

The low-pressure feed pump that provides water to the feed tank was a residential water pressure booster pump. Although it was not made to handle salty water, the contractor thought it would last for at least the duration of the project. It did not, and neither did a subsequent replacement. The third attempt involved using a centrifugal pump that was available because it had been used in a previous project at the KBH site. A hydraulic pressure switch was used temporarily to control the on-off functioning of the low-pressure pump until a pressure tank was acquired and installed, after which there were no other problems with the raw water feed pump.

Whereas most of the problems encountered in conducting this research were specific to this project, this research project does illustrate that, as pilot plants increase in complexity by incorporation of automatic controls and datum logging, process interruptions are likely to be more frequent than in a system that is manually operated and manually controlled.

### Chapter 5

### Data Collection, Data Analysis, and Findings

At the outset of the project, it was assumed that concentrate recoveries up to about 70% would be readily achievable because of the 85% recovery that is still routinely achieved in the batch-treatment seawater RO system. At 70% recovery of the concentrate from the KBH plant, the overall recovery would increase from 80% (which is the current recovery at the KBH plant) to 94%. The plan was to start at a relatively low recovery and then increase it in 5% increments until fouling of the membranes occurred. Therefore, the initial test runs were conducted for only 1 day because it appeared that equilibrium conditions had been attained and that the system was stable. It was later discovered that membrane fouling at recoveries below 50% usually did not begin until sometime during the second day of operation.

Figure 5.1 is a plot of transmembrane pressure versus time for the first 15 h after startup at a recovery of 48%. The pressure was essentially constant after the feed tank conductivity reached the preset value of 30,000  $\mu$ S/cm (corresponds to a TDS concentration of approximately 20,000 mg/L), approximately 4 h after the system was started. For the purposes of this research project, because the feed tank and discharge point are both operated under atmospheric conditions, the terms "transmembrane" pressure and "feed" pressure are essentially the same.



Figure 5.1. Transmembrane pressure for first 15 h after startup at 48% recovery, with a preset value of 30,000 µS/cm, in terms of feed tank concentration.

Figure 5.2 is a plot of transmembrane pressure versus time after the first 15 h of run time. The pressure was already starting to increase as evidenced by the slight upward slope (namely, 0.2023) of the best fit line. The pilot membrane system was designed to automatically shut down operation at a prescribed, or preset, pressure. Before the end of the second day, the pressure reached 800 psi and the system automatically shut down. Inspection of the feed tank revealed a significant amount of precipitate on the sides and bottom of the tank. Wet chemical analysis and X-ray diffraction indicated that the precipitate was calcium sulfate. Previous electron microscope analyses of precipitate in the batch-treatment concentrate revealed the same results plus a small amount of silicon. It is not known whether the precipitation began in the membrane or in the tank, but once precipitate was present in either

place, the calcium sulfate crystals would likely have served as a seed, resulting in enhanced precipitation. It is possible that use of a cartridge filter before the membrane could have reduced this problem, if precipitation occurred only in the feed tank.

Table 5.1 shows the concentration of several parameters in the KBH concentrate (namely, the raw feed water to the SWRO unit) and in the permeate and concentrate from the SWRO pilot plant. The rejection for all parameters was very high, averaging more than 98%, yielding a very high-quality permeate having a TDS concentration of 268 mg/L.



Figure 5.2. Transmembrane pressure after first 15 h at 48% recovery, with preset conductivity value of 30,000 µS/cm, in terms of feed tank concentration.

Parameter	SWRO Feed	SWRO Permeate	SWRO Concentrate	% Rejection
1 41 4110001	51110100	5 (110 I trintente		nejeenen
Hardness	2610	5	5250	99.8%
$Ca^{+2}$	570	1	1060	99.8%
$Mg^{+2}$	186	1	364	99.5%
Na <sup>+1</sup>	2950	66	5780	97.8%
$K^{+1}$	86	2.7	164	96.9%
Cl <sup>-1</sup>	6050	167	11,200	97.2%
$SO_4^{-2}$	1800	23	3320	98.7%
Silica	155	2	266	98.7%
Conductivity	19,875	541	36,600	97.3%
TDS	12,970	268	24,100	97.9%

Table 5.1	. Results	at 48%	Recovery
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*Note.* All values in milligrams per liter except conductivity ( $\mu$ S/cm); hardness is expressed as milligrams of CaCO<sub>3</sub> per liter.

The mass balance for the data in Table 5.1 is shown in Table 5.2. The largest discrepancy was -9% for silica. The large difference occurred probably because the samples were analyzed at the EPWU IWQL, where the turnaround is normally 2 weeks. The supersaturated silica in the SWRO concentrate polymerizes with time, and the test that measures silica measures only the monomeric form. Therefore, even if the silica does not precipitate, the polymerized silica would not be measured in the colorimetric test the laboratory uses. All of the other parameters were reasonably close to their expected values.

Parameter	Mass In	Permeate Mass	Concentrate Mass	Permeate + Concentrate	Diff	% Diff
Hardness	10,620	10	10,586	10,595	-25	0%
Ca <sup>+2</sup>	2319	2	2267	2269	-50	-2%
Mg <sup>+2</sup>	757	2	778	780	23	3%
Na <sup>+1</sup>	12,003	128	12,361	12,489	486	4%
$\mathbf{K}^{+1}$	350	5	351	356	6	2%
Cl <sup>-1</sup>	24,617	322	23,951	24,274	-343	-1%
$SO_4^{-2}$	7324	45	7100	7145	-179	-2%
Silica	631	4	569	573	-58	-9%
Conductivity	80,869	1044	78,270	79,314	-1555	-2%
TDS	52,773	517	51,538	52,056	-717	-1%

Table 5.2. Mass Balance at 48% Recovery

Note. Mass values are milligrams per minute, except conductivity (microsiemens-liter/centimeter-minute).

Figure 5.3 is a plot of transmembrane pressure and conductivity versus time for a feed tank conductivity setting of 55,000  $\mu$ S/cm (corresponds to a TDS of approximately 35,000 mg/L). It took approximately 12 h for the feed tank conductivity to reach the preset value of 55,000  $\mu$ S/cm as shown on the graph. Thereafter, the transmembrane pressure remained relatively constant (at about 530 psi) for only 3 h, after which it steadily increased. During the last 30 min of operation, the transmembrane pressure increased very rapidly, going from 600 to 800 psi, causing the system to automatically shut down because of high pressure.



Figure 5.3. Transmembrane pressure and conductivity at 55,000-µS/cm recovery.

The chemical analyses of the SWRO permeate and concentrate are shown in Table 5.3. The rejection of the divalent ions remained high at more than 95%, and rejection of all of the other substances was at 90%. These rejection rates are lower than they would be for a new membrane, but at the time these data were collected, the membrane had been chemically cleaned four times. Whether the fouling and subsequent chemical cleaning had anything to do with the lower-than-anticipated rejection rates is not known.

Parameter	KBH Concentrate	SWRO Permeate	SWRO Concentrate	% Rejection
Hardness	2010	99	7260	95.1
$Ca^{+2}$	538	26.4	1960	95.1
$Mg^{+2}$	142	5.9	505	95.8
$Na^{+1}$	2560	236	9100	90.8
$K^{+1}$	57.6	7.8	223	86.5
Cl <sup>-1</sup>	4510	465	15,800	89.7
$SO_4^{-2}$	1050	50.3	4980	95.2
Silica	107	6.1	342	94.3
TDS	9600	850	34,200	91.1
			Avg	92.6

 Table 5.3. Results at 74% Recovery

Note. All concentrations in milligrams per liter; hardness is expressed as milligrams of CaCO<sub>3</sub> per liter.

The results were the same for all recoveries of more than 48% (namely, conductivities above 30,000  $\mu$ S/cm), so the conductivity set point was reduced to 27,500  $\mu$ S/cm, which represents a recovery of about 35%. The system operated at this recovery almost continuously over 7 days, except for brief interruptions when a loose connection from the pH probe produced the default pH reading of 14, shutting the system down. This recovery was the highest for which no fouling was observed. This disappointingly low recovery led to modifications of the operating conditions as discussed in the next chapter.

### **Membrane Cleaning**

When the membrane was fouled the first time, it was not known whether the foulant was silica or calcium sulfate, so both types of cleaning solutions were acquired. The first attempt was made using High Flux A (1:40 dilution) that is specific for silica. When the pilot plant was restarted, the membrane pressure was still high, so the Diamite CAL cleaning solution at a 1:40 dilution was used. The solution was circulated through the membrane for about 1 h and then was allowed to soak overnight. Recirculation was resumed the next morning for about 10 min (the solution had turned yellow-orange by that time), and then the membrane was flushed with RO permeate. When the system was restarted, the pressures were down to their prefouled values. The results are shown in Figure 6.1.

Figure 6.1 shows the transmembrane pressures for five different events: (a) prior to any membrane fouling events (namely, April 26), (b) while the membrane was fouled (namely, May 10), and (c) after the membrane had been cleaned following three different fouling events (namely, May 17, June 24, and July 8). In all cases, the transmembrane pressures were about the same after chemical cleaning as they were before the membrane was ever fouled, indicating that the foulant was indeed primarily calcium sulfate. The cleaning method was always the same for each and every date represented on Figure 6.1. The purpose of this figure is to portray system test conditions (pressure and feed conductivity) prior to and during membrane fouling events, as well as after chemical cleaning of the membrane. Review of the figure indicates that cleaning of the membrane returns the membrane to a condition that approximates the condition existing prior to fouling, at least in terms of system pressure and feed conductivity.

It should be pointed out that, once an element is fouled with CaSO<sub>4</sub>, all the nucleation sites may not be removed during cleaning. Thus, the membrane may subsequently foul more quickly under supersaturation conditions because of the presence of these nucleation sites, resulting in fouling of the cleaned membrane surface occurring more quickly than fouling of a new one. Whether this factor was significant in this project is not known.



Figure 6.1. Membrane pressures before and after chemical cleaning. April 26 data points represent conditions prior to membrane fouling. May 10 data points represent a fouled membrane condition. May 17, June 24, and July 8 data points represent conditions after membrane cleaning events.

### Chapter 7

### **Operating Modifications**

In an attempt to obtain recoveries in the range of at least 50%, several modifications were made to the operating conditions of the pilot plant. Because calcium sulfate was apparently the foulant that was limiting recovery in the system, two steps were taken to address this problem. The first was to add a heat exchanger to the feed tank to lower the temperature of the feed solution because temperature readings as high as 39 °C were recorded in the feed tank during test runs at recoveries above 50%. Because the solubility of some calcium sulfate phases decreases with increased temperature, a heat exchanger functioned properly, keeping the temperature below 26 °C, but precipitation occurred again in Day 2, even at the lower conductivity setting of 30,000  $\mu$ S/cm. In the unlikely event that the stainless steel tubing may have initiated the precipitation, the stainless steel tubing was replaced with an all-plastic heat exchange unit, but precipitation still occurred on the second day.

After chemical cleaning of the membrane with Diamite CAL, the second step in dealing with calcium sulfate precipitation involved increasing the scale inhibitor feed rate so that its concentration would increase from 4 to 20 mg/L (the scale inhibitor is specifically intended for inhibiting calcium sulfate precipitation). This high dosage of scale inhibitor was applied as an experiment only to determine whether it would in fact prevent precipitation. As a practical matter, such a high dose of scale inhibitor (15 to 20 mg/L) would most likely prove to be excessive and cost prohibitive for a production-scale facility. The increased concentration did not work, as precipitation occurred again within 2 days.

The third modification involved reducing the membrane flux from 9 to 5.4 gfd, even though the flux of 9 gfd was well within the manufacturer's specification for that membrane element. The reduced flux did not solve the problem, as the membrane fouled again.

The final modification involved reducing the volume of water in the feed tank from 200 to 15 gal. Researchers hoped that the shorter time of retention in the feed tank would decrease the tendency for calcium sulfate to precipitate, but it did not. The adjustment to pH was made using only sulfuric acid; hydrochloric acid was not applied. Thus, none of the changes in operating conditions beneficially affected system performance from the standpoint of reduced membrane fouling. In terms of comparing the final membrane performance using sulfuric acid with that using hydrochloric acid, it turned out that the actual impact of selecting sulfuric acid was minimal, compared to using hydrochloric acid, in terms of any additional sulfate scaling potential on the membranes.

The main problem appears to be associated with the design of the treatment system itself. That is, the feed tank is *maintained* at a solid concentration that exceeds the saturation value of one or more compounds. As soon as something triggers one of the compounds to precipitate, the process continues very rapidly in both the feed tank and in the pilot membrane, thus fouling the membrane. In the batch-treatment system that was the forerunner to this project, the supersaturated condition lasts for only a relatively short period (namely, less than the induction time for calcium sulfate precipitation) before the superconcentrate is dumped, and thus the precipitation problem is avoided. It would seem that a possible solution to this problem is a continuous flow system that has a single-pass design with no RO concentrate recirculation back to the feed tank.

### Chapter 8

### **Economic Considerations and Evaluation of System Performance**

In considering the economics of the continuous flow SWRO concentrate recovery process, the values used in the calculations are reflective of the costs at KBH in El Paso. It is likely that researchers and engineers investigating potential projects at locations exhibiting conditions similar to those in El Paso could utilize the results of this study to derive cost estimates for their particular project site. For example, prudent adjustments could be made in terms of scale, regional wages, and other cost components in order to derive estimated costs for water produced at other locations using the continuous flow SWRO process. The values associated with the parameters used in deriving the costs for a full-scale project are shown in Table 8.1. Costs shown in this table are primarily unit costs and are derived from various sources including general industry standards, bids values received by EPWU for various projects, commonly accepted values within the water industry, and so forth. Table 8.2 contains total, site-specific costs for this project based on the unit costs from Table 8.1.

The \$2.00/1000-gal (kgal) selling price for potable water used in Table 8.1 is equivalent to a selling price to the customer who uses water very frugally and therefore purchases his water at the lower rates under EPWU's progressive rate structure. Comparatively, the \$2.00/kgal price is also representative of the total, current cost of potable water to EPWU from the KBH Plant. It is an equivalent amortized cost that includes both capital and long-term operations and maintenance costs. The KBH plant is EPWU's most recently constructed facility for the treatment and production of potable water provided to the customer.

Therefore, this \$2.00/kgal value represents a low estimate of income to the water utility, because EPWU charges a progressive rate for water sales. For example, a substantial amount of water is sold at higher prices, especially during the summer, when demand and consumption are highest and when many customers purchase water at the higher unit rates. Cost-related components that are not included in this project and that may have to be included in projects considered elsewhere are land costs, pipeline costs, and permanent solid disposal costs in a landfill. These components are not included herein because of circumstances that are unique to this project.

Equipment Design Criteria						
Item	Value					
Initial RO conc volume (gpd)	3,600,000					
Evaporation rate (in./year)	50					
Liner life (years)	20					
Flow storage (months)	6					
Equipment life (years)	20					
Blending well depth (ft.)	300					
Blending ratio (blend vol./total vol.)	33%					
SWRO membrane life (years)	5					
Pump and motor efficiency (%)	75%					
SWRO flux (gpd/sq. ft.)	9					
SWRO operating pressure (psi)	300					
H <sub>2</sub> SO <sub>4</sub> feed rate (mL/gallon)	1.0					
Area per membrane element (sq. ft.)	400					
Capital Cost Assumptions						
Item	Value					
Interest rate (%)	6%					
Interest rate (%) Liner cost (\$/sq. ft.)	6% 0.73					
Interest rate (%) Liner cost (\$/sq. ft.) Excavation cost (\$/cu. yard)	6% 0.73 \$3.00					
Interest rate (%) Liner cost (\$/sq. ft.) Excavation cost (\$/cu. yard) Fence cost (\$/linear ft.)	6% 0.73 \$3.00 \$10.00					
Interest rate (%) Liner cost (\$/sq. ft.) Excavation cost (\$/cu. yard) Fence cost (\$/linear ft.) Excavation, fence amortization basis (years)	6% 0.73 \$3.00 \$10.00 20					
Interest rate (%) Liner cost (\$/sq. ft.) Excavation cost (\$/cu. yard) Fence cost (\$/linear ft.) Excavation, fence amortization basis (years) Water selling price (\$/kgal)	6% 0.73 \$3.00 \$10.00 20 \$2.00					
Interest rate (%) Liner cost (\$/sq. ft.) Excavation cost (\$/cu. yard) Fence cost (\$/linear ft.) Excavation, fence amortization basis (years) Water selling price (\$/kgal) SWRO capital cost (\$/mgd)	6% 0.73 \$3.00 \$10.00 20 \$2.00 \$1,000,000					
Interest rate (%) Liner cost (\$/sq. ft.) Excavation cost (\$/cu. yard) Fence cost (\$/linear ft.) Excavation, fence amortization basis (years) Water selling price (\$/kgal) SWRO capital cost (\$/mgd) Building cost (\$)	6% 0.73 \$3.00 \$10.00 20 \$2.00 \$1,000,000 \$320,000					
Interest rate (%) Liner cost (\$/sq. ft.) Excavation cost (\$/cu. yard) Fence cost (\$/linear ft.) Excavation, fence amortization basis (years) Water selling price (\$/kgal) SWRO capital cost (\$/mgd) Building cost (\$) Cost per 8-in. membrane	6% 0.73 \$3.00 \$10.00 20 \$2.00 \$1,000,000 \$320,000 \$800.00					
Interest rate (%) Liner cost (\$/sq. ft.) Excavation cost (\$/cu. yard) Fence cost (\$/linear ft.) Excavation, fence amortization basis (years) Water selling price (\$/kgal) SWRO capital cost (\$/kgal) Building cost (\$) Cost per 8-in. membrane SWRO membrane cost (\$/sq.ft.)	6% 0.73 \$3.00 \$10.00 20 \$2.00 \$1,000,000 \$320,000 \$800.00 \$2.00					
Interest rate (%) Liner cost (\$/sq. ft.) Excavation cost (\$/cu. yard) Fence cost (\$/linear ft.) Excavation, fence amortization basis (years) Water selling price (\$/kgal) SWRO capital cost (\$/kgal) Building cost (\$) Cost per 8-in. membrane SWRO membrane cost (\$/sq.ft.)	6% 0.73 \$3.00 \$10.00 20 \$2.00 \$1,000,000 \$320,000 \$800.00 \$2.00					
Interest rate (%) Liner cost (\$/sq. ft.) Excavation cost (\$/cu. yard) Fence cost (\$/linear ft.) Excavation, fence amortization basis (years) Water selling price (\$/kgal) SWRO capital cost (\$/mgd) Building cost (\$) Cost per 8-in. membrane SWRO membrane cost (\$/sq.ft.) O&M Cost Assumptions	6% 0.73 \$3.00 \$10.00 20 \$2.00 \$1,000,000 \$320,000 \$800.00 \$2.00					
Interest rate (%) Liner cost (\$/sq. ft.) Excavation cost (\$/cu. yard) Fence cost (\$/linear ft.) Excavation, fence amortization basis (years) Water selling price (\$/kgal) SWRO capital cost (\$/mgd) Building cost (\$) Cost per 8-in. membrane SWRO membrane cost (\$/sq.ft.) O&M Cost Assumptions Item	6% 0.73 \$3.00 \$10.00 20 \$2.00 \$1,000,000 \$320,000 \$320,000 \$2.00 <b>Value</b>					
Interest rate (%) Liner cost (\$/sq. ft.) Excavation cost (\$/cu. yard) Fence cost (\$/linear ft.) Excavation, fence amortization basis (years) Water selling price (\$/kgal) SWRO capital cost (\$/mgd) Building cost (\$) Cost per 8-in. membrane SWRO membrane cost (\$/sq.ft.) O&M Cost Assumptions Item Power costs (\$/kwh)	6% 0.73 \$3.00 \$10.00 20 \$2.00 \$1,000,000 \$320,000 \$800.00 \$2.00 <b>Value</b> \$0.08					
Interest rate (%) Liner cost (\$/sq. ft.) Excavation cost (\$/cu. yard) Fence cost (\$/linear ft.) Excavation, fence amortization basis (years) Water selling price (\$/kgal) SWRO capital cost (\$/mgd) Building cost (\$) Cost per 8-in. membrane SWRO membrane cost (\$/sq.ft.) O&M Cost Assumptions Item Power costs (\$/kwh) Scale inhibitor dosage (ppm)	6% 0.73 \$3.00 \$10.00 20 \$2.00 \$1,000,000 \$320,000 \$320,000 \$2.00 \$2.00					
Interest rate (%) Liner cost (\$/sq. ft.) Excavation cost (\$/cu. yard) Fence cost (\$/linear ft.) Excavation, fence amortization basis (years) Water selling price (\$/kgal) SWRO capital cost (\$/mgd) Building cost (\$) Cost per 8-in. membrane SWRO membrane cost (\$/sq.ft.) <b>O&amp;M Cost Assumptions</b> <b>Item</b> Power costs (\$/kwh) Scale inhibitor dosage (ppm) Scale inhibitor cost (\$/gal.)	6% 0.73 \$3.00 \$10.00 20 \$2.00 \$1,000,000 \$320,000 \$800.00 \$2.00 <b>Value</b> \$0.08 4 \$11.00					
Interest rate (%) Liner cost (\$/sq. ft.) Excavation cost (\$/cu. yard) Fence cost (\$/linear ft.) Excavation, fence amortization basis (years) Water selling price (\$/kgal) SWRO capital cost (\$/mgd) Building cost (\$) Cost per 8-in. membrane SWRO membrane cost (\$/sq.ft.) <b>O&amp;M Cost Assumptions</b> Item Power costs (\$/kwh) Scale inhibitor dosage (ppm) Scale inhibitor cost (\$/gal.) $H_2SO_4 cost ($/gal.)$	6% 0.73 \$3.00 \$10.00 20 \$2.00 \$1,000,000 \$320,000 \$800.00 \$2.00 <b>Value</b> \$0.08 4 \$11.00 \$2.53					
Interest rate (%) Liner cost (\$/sq. ft.) Excavation cost (\$/cu. yard) Fence cost (\$/linear ft.) Excavation, fence amortization basis (years) Water selling price (\$/kgal) SWRO capital cost (\$/mgd) Building cost (\$) Cost per 8-in. membrane SWRO membrane cost (\$/sq.ft.) <b>O&amp;M Cost Assumptions</b> <b>Item</b> Power costs (\$/kwh) Scale inhibitor dosage (ppm) Scale inhibitor cost (\$/gal.) $H_2SO_4 cost ($/gal.)$ Personnel cost (\$/year, 3@\$35,000 each)	6% 0.73 \$3.00 \$10.00 20 \$2.00 \$320,000 \$320,000 \$2.00 \$2.00 <b>Value</b> \$0.08 4 \$11.00 \$2.53 \$105,000					

Table 8.1. Unit Values for Calculating Cost of Continuous FlowSWRO System at 35% Recovery

Item				Value		
Membran	140,000					
SWRO eo Amortizeo	SWRO equipment capital cost (\$) Amortized SWRO equipment capital cost (\$ per					
year)				\$109,853		
Membran	e cost (\$)			\$280,000		
Amortized	d membran	e cost (\$ p	er year)	\$66,471		
Flow rate Horsepov	S.)	5.54				
(hp)				579		
Power co Disposal	st (\$) volume (ga	I. per		\$304,327		
day)				2,340,000		
Pond area	a (sq. ft.) ume (cu.			27,262,872		
ft.)				56,953,260		
Liner cost	Liner cost (\$)					
Excavatio	on cost (\$)			\$6,328,410		
Fencing o	cost (\$)			\$185,093		

Table 8.2. Total Values for Calculating Cost of Continuous Flow SWROSystem at 35% Recovery

The calculations using the values from Table 8.1 at 35% recovery reveal that the cost of the product water would be \$5.80/kgal, which represents a net cost of about \$3.08 million per year, in terms of cost of amortized capital plus operations and maintenance to the utility (see Appendix B for sample calculations). This cost would have to be compared to the cost of alternative concentrate management options in evaluating the feasibility of the continuous flow process tested here, but it is not likely to be attractive in very many locales. If overall process recoveries, significantly exceeding 35%, were proven to be achievable, then more potable water would be available to sell to the public, thus decreasing the overall cost of the SWRO process. The break-even point is at the recovery that would have a total cost that would be exactly offset by the revenue generated by the sale of the recovered water.

Break-even for this process (when adjusted for an assumed transmembrane operating pressure of 600 psi and no blending) would occur at a recovery of about 95%. A recovery rate of 95% appears to be unreasonably high for a continuous flow SWRO process, as exhibited during this pilot program. Recoveries between 35% and 95% would have costs between \$2.00/1000 gal and \$5.80/1000 gal, but as discussed previously in this report, recoveries higher than 35% were not achievable with the system design as tested here. The capital cost portion of the total cost is a fixed item, because the capital costs do not vary after the system is built. At higher recoveries, more water is available to sell, generating more income. Furthermore, the cost of final disposal of the superconcentrate (through evaporation) decreases proportionately with increased recovery. Thus, higher recoveries are more economically attractive. Inclusion of items that were not considered here such as land cost would add about \$0.05 per kgal of product water for each \$1 million of capital cost.

### **Conclusions and Recommendations**

On the basis of this investigation, the following conclusions can be made with reasonable certainty:

- 1. Water recoveries exceeding 35% of KBH concentrate were not achievable with the design of the pilot plant used in this project. The continuously supersaturated condition of the concentrate in the feed tank led to precipitation of calcium sulfate, which consequently fouled the SWRO membrane.
- 2. Various modifications to the pilot test regimen were not successful in preventing calcium sulfate precipitation. They included installing a heat exchanger in the feed tank, raising the scale inhibitor dosage from 4 to 20 mg/L, reducing the membrane flux from 9 to 5.4 gfd, and decreasing the feed tank volume from 200 to 15 gal.
- 3. The cleaning solution Diamite CAL from King Lee Technologies was effective in removing calcium sulfate (and other compounds that might have also been there) from the fouled SWRO membrane.
- 4. The original presumption at the beginning of this research project was that the continuous flow arrangement would be effective and economical. This presumption was proven incorrect for the particular design of the pilot treatment system implemented. The plumbing and design of this pilot system resulted in a supersaturated liquid condition prevailing within the feed tank for many minerals and compounds for an extended period ranging from hours to days. Because one or more chemicals were in a supersaturated state in the final concentrate, precipitation was inevitable sooner or later, resulting in a fouled membrane. This result stems from exceeding the induction phenomenon time previously described in detail in Chapter 7.
- 5. A better design for a continuous flow system would include a once-through, singlepass flow scheme with no recirculation back to the feed tank. In other words, the concentrate would be discarded to waste just before the end of the induction period for precipitation of calcium sulfate. This arrangement would be consistent with a fullscale system and would minimize the time that the concentrate with supersaturated salts would remain in the SWRO membrane and the treatment system.
- 6. At 35% recovery, the cost of the produced water would be \$5.80/kgal in a full-scale continuous flow system that is similar to the one tested in this project. The net cost per year would be about \$3.08 million. For break-even, the recovery of KBH concentrate would have to be about 95%, which would represent an overall desalting plant recovery of 99%. Concentrate recoveries that high could not be achieved in any SWRO system for several reasons, including excessive osmotic pressures.

The previous success obtained in treating KBH concentrate with a batch-treatment SWRO system suggests that a different design of the experimental continuous flow system may possibly be successful in blocking the exceeding of the previously described induction time for calcium sulfate. A single-pass SWRO system should be designed and tested to prove the concept and then be evaluated from the standpoint of economic feasibility. A single-pass design would theoretically exhibit many of the advantages of the batch-system design, such as maintaining the ability to restrain the TDS in the feed tank at a solid concentration near the "supersaturated" condition for only a short period before the concentrate is discharged to waste.

### References

- Alimi, F.; Elfil, H.; Gadri, A. Kinetics of the precipitation of calcium sulfate dihydrate in a desalination unit. *Desalination* **2003**, *158*, 9–16.
- Amjad, Z. Applications of antiscalants to control calcium sulfate scaling in reverse osmosis systems. *Desalination* 1985, 54, 263–276.
- Avista Technologies. Antiscalant Technical Data Sheet, 2008. http://www.avistatech.com/Antiscalants/Antiscalant%20Technical%20Data%20Shee t.htm (accessed July 2012).
- Corbett, B. E.; Moody, C. D.; Morris, M. D. Evaluation of Reverse Osmosis Scaling Prevention Devices at High Recovery, March 2003. http://www.usbr.gov/pmts/water/publications/reportpdfs/report091.pdf (accessed July 2012).
- EPWU. *Quality Manual;* International Water Quality Laboratory, EPWU: El Paso, TX, 2011; 157–173.
- Lancia, A.; Musmarra, D.; Prisciandaro, M. Measuring induction period for calcium sulfate dihydrate precipitation. AIChE J. 1999, 45, 390–397. doi: 10.1002/aic.690450218
- Mi, B.; Elimelech, M. Gypsum scaling and cleaning in forward osmosis: measurements and mechanisms. *Environ. Sci. Technol.* **2010**, *44*, 2022–2028.
- Ning, R.; Tarquin, A. Crystallization of salts from super-concentrate produced by tandem RO process. *Desalin. Water Treat.* 2010, 16, 238–242.
- Ning, R.; Tarquin, A.; Balliew, J. Seawater RO treatment of RO concentrate to extreme silica concentrations. *Desalin. Water Treat.* **2010**, *22*, 286–291.
- Pomerantz, N.; Ladizhansky, Y.; Korin, E.; Waisman, M.; Daltrophe, N.; Gilron, J. Prevention of scaling of reverse osmosis membranes by "zeroing" the elapsed nucleation time. Part 1. Calcium sulfate. *Ind. Eng. Chem. Res.* 2006, 45, 2008–2016.
- Sarig, S.; Mullin, J. W. Effect of trace impurities on calcium sulfate precipitation. J. Chem. Technol. Biotechnol. 1982, 32, 525–531.
- Shih, W-Y.; Rahardianto, A.; Lee, R-W.; Cohen, Y. Morphometric characterization of calcium sulfate dihydrate (gypsum) scale on reverse osmosis membranes. J. Membr. Sci. 2005, 252, 253–263.
- Tarquin, A. J. Volume Reduction of High-Silica RO Concentrate Using Membranes and Lime Treatment, Feb. 2005. http://www.usbr.gov/research/AWT/reportpdfs/report108.pdf (accessed July 2012).
- Tarquin, A. J. Cost-Effective Volume Reduction of Silica-Saturated RO Concentrate, March 2006. http://www.usbr.gov/research/AWT/reportpdfs/report125.pdf (accessed July 2012).
- Tarquin, A. *High Tech Methods to Reduce Concentrate Volume Prior to Disposal*, March 2010.

http://www.twdb.texas.gov/RWPG/rpgm\_rpts/0704830769\_ConcentrateVolume.pdf (accessed July 2012).

## Appendix A

## **Control Panel Screen Shots**

Industrial Water Servic	10:50:33 AM 16-APR-10	
Se: De:	a Water Re∨erse Osmo Pilot Unit signed by: Anthony Tarq	sis uin
Information	Navigation	Screen Saver

Figure A.1. Main screen.

Antiscalant Injection Control		Acid Injection Control		
Gallons per Pulse	Press to place	Press to place	Gallons per Pulse	
(% Str*1.3 mL/Str)	In AUTO	in AUTO	(% Str*1.3 mL/Str)	
0.00 Gal	Press to Pulse Pump	Press to Pulse Pump	0.00 Gal	
Antiscalant Injection		Acid Injection		
Stop		Stop		
l	Press to Prime	Press to Prime	Minimum pH Setpoint	
	Pump	Pump	0.0	
Information	Navigation		Screen Saver	

Figure A.2. Chemical injection screen.



Figure A.3. Navigation screen.

	System Sta	art System Stop				
	Startup Flush 9	Coff	Periodic Flush			Permeate Flush
Startup Flush Time O Sec	Shutdown Flu Time O Sec	sh Pe	riodic Flush Time O Sec	Time Be Periodic 0.0	etween s Flush Hrs	Permeate Flush Length Omins
		Res	et Flow Averag Totalizers	ing		
Information Nav		Navigation			Screen Saver	

Figure A.4. System start screen.

Bleed Valve Status (Press to OPEN/CLOSE Manually)	Bleed Flow 0.00 GPM	Feed Cond O uS
Cond C Off	Control On O uS	
Information	Navigation	Screen Saver

Figure A.5. Tank PID control screen.



Figure A.6. System performance screen.



Figure A.7. Permeate PID flow control screen.

Alarm Histo	ry Tot	al of 0 Alarms
Entry No Alarm No	Message	Confirm
Information	Navigation	Screen Saver

Figure A.8. Alarm screen.



Figure A.9. Process screen.



Figure A.10. Information screen.



Figure A.11. Emergency stop screen.

### Appendix B Sample Calculations

These calculations are used to generate the unit cost of water produced.

Membrane area = permeate volume/flux = 1,260,000/9.0 $= 140.000 \text{ ft}^2$ Number of membrane elements = total membrane area/element area = 140,000/400= 350 membrane elements RO equipment capital cost = concentrate volume\*capital cost/mgd = 1.26 \* 1,000,000= \$1,260,000 Amortized RO equipment cost = RO equipment cost\*(A/P, 6%, 20)= 1,260,000\*(A/P, 6%, 20)= \$109,853 per year Membrane cost = (permeate volume/flux)\*membrane cost/ft<sup>2</sup>= (1,260,000/9)\*2.00 (membrane cost can be estimated on the basis of unit cost per membrane element. A membrane manufacturer can provide a budget quote for SWRO membrane elements, namely, \$550-600 per membrane element) = \$280,000 Amortized membrane cost = membrane cost\*(A/P, 6%, 5) = 280,000\*(A/P, 6%, 5)= \$66,471 per year Rate of flow to RO in cfs = RO water volume\*conversion to cfs from gpd = 3,600,000\*0.133/(24\*60\*60)= 5.54 cfsHorsepower for RO unit =  $flow^*$  \*head/(550\*Eff) = 5.54\*62.4\*300\*2.303/(550\*0.75)= 579 hp Power cost = horsepower\*0.75 kw/hp\*h/day\*\$/kWh\*days/year = 579\*0.75\*24\*0.08\*365= \$304,327 per year

Disposal volume = raw concentrate volume - raw concentrate volume\*recovery = 3,600,000 - 3,600,000(0.35)= 2,340,000 gal per day Pond area ( $ft^2$ ) = disposal volume\*0.133\*365/(evaporation rate/12) =2.340,000\*0.133\*365/(50/12) $= 27,262,872 \text{ ft}^2$ Pond area (acres) = disposal volume\*0.133\*365/(evaporation rate/12)/43,560 = 2,340,000\*0.133\*365/(50/12)/43,560 = 626 acresLiner cost = pond area\*liner cost/  $ft^2$ = 27,262,872\*0.73 = \$19,901,897 (may be higher for a double liner, per some regulations) Pond volume ( $ft^3$ ) = disposal volume\*0.133\*30.5\*months of storage = 2,340,000\*0.133\*30.5\*6  $= 56,953,260 \text{ ft}^3$ Excavation cost = (pond volume/27)\*excavation cost= (56,953,260/27)\*\$3/CY = \$6,328,140 Fencing cost = pond circumference\*fence cost/ft  $=([27,262,872*4]/\pi)^{0.5*\pi*}10/LF$ = \$185,093

#### Explanatory Note:

Unit costs shown in Appendix B are derived from various sources including general industry standards, bid values received by EPWU for various projects, commonly accepted values within the water industry, and so forth.

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