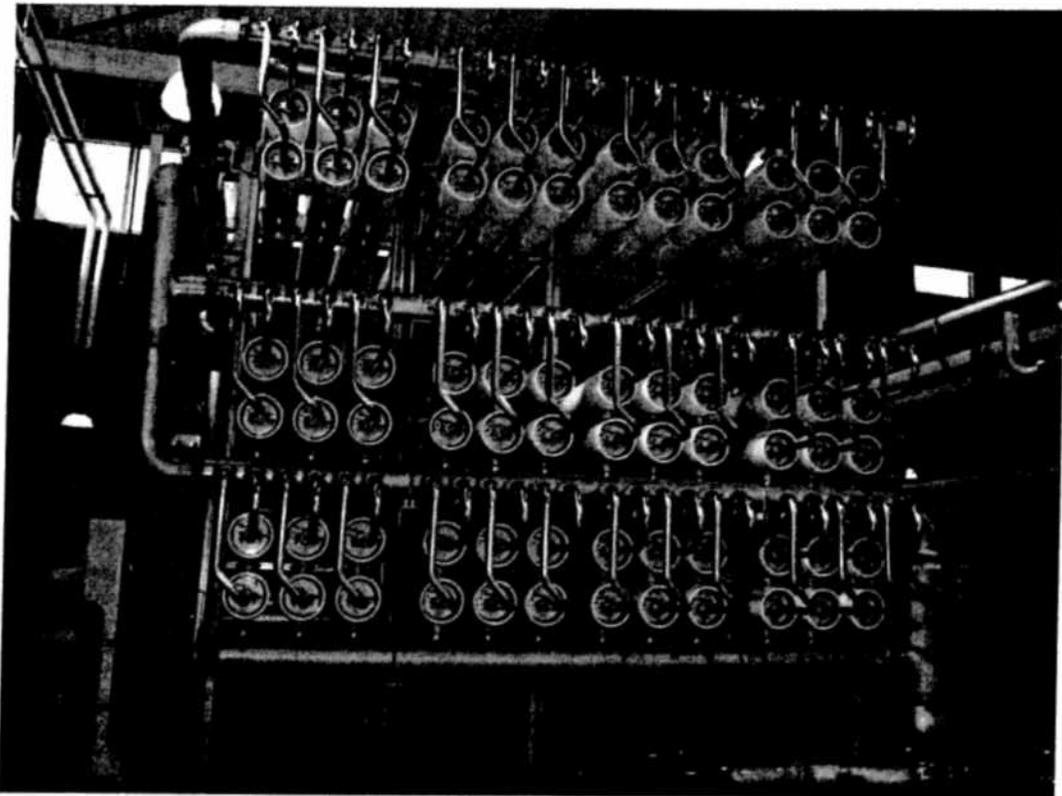


# Continuous Flow Seawater RO System for Recovery of Silica- Saturated RO Concentrate

by  
Anthony J. Tarquin, Ph.D.  
The University of Texas at El Paso



**Texas Water Development Board**  
**Contract #0704830769 (Amendments 3 and 4)**

July 2011

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This project was funded in part by the WaterReuse Research Foundation and by the Texas Water Development Board.

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## List of Abbreviations

EPWU	El Paso Water Utilities
gfd	Gallons per square foot per day
gpd	Gallons per day
gpm	Gallons per minute
KBH	Kay Bailey Hutchison
MGD	Million gallons per day
ppm	Parts per million
psi	Pound per square inch
RO	Reverse osmosis
SWRO	Seawater reverse osmosis
Sq. ft.	Square feet
TDS	Total dissolved solids
VSEP	Vibratory shear enhanced process
$\mu\text{S/cm}$	MicroSiemens per centimeter

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## **Executive summary**

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 over 85% of the water from the silica-saturated reverse osmosis concentrate that is generated at the Kay Bailey Hutchison (KBH) Desalting plant in El Paso, TX. 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 was installed to keep the water in the feed tank at a pre-set conductivity, which establishes the recovery of the system. An antiscalant for sulfate control was added into the system and the pH in the feed tank was maintained at pH 3.9 through the addition of sulfuric acid.

Membrane autopsy revealed that calcium sulfate caused fouling on the SWRO membrane. In addition, calcium sulfate precipitated in the feed tank at all recoveries above 35%, which corresponds to a feed water total dissolved solids (TDS) concentration of approximately 15,000 milligrams per liter. Various attempts to prevent calcium sulfate precipitation were unsuccessful, including increasing the antiscalant 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. 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 of time before the concentrate is discharged from the membrane (i.e. the treatment time must be less than the induction period for calcium sulfate precipitation).

At the outset of the project, it was believed that silica scaling would be the most significant challenge to designing a continuous flow system. However, the only membrane fouling that was observed throughout the study was due to calcium sulfate. Silica precipitation was never a problem during this pilot study.

# 1 Introduction

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. At the present time, economic considerations limit the concentrate management options at inland desalination plants to injection and evaporation, and each of them have limitations of their own. In the case of injection, the cost can be prohibitive if the proper geological formation is not reasonably close to the desalting plant. In the case of El Paso's Kay Bailey Hutchison Desalting Plant, the proper geological formation is 22 miles away and 2,500 feet deep, so injection has been an expensive proposition. Evaporation is not feasible in areas that receive a large amount of rainfall, but in the southwestern U.S., the arid climate is ideal for evaporation. Since 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. 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.

## 1.1 Background

El Paso Water Utilities (EPWU), in partnership with Fort Bliss (a U.S. Army installation), owns and operates the largest inland brackish groundwater reverse osmosis (RO) desalting plant in the United States. The Kay Bailey Hutchison Desalting Plant produces 27.5 million gallons of drinking water per day when operating at full capacity. The plant operates at 80% recovery and disposes of the high-silica concentrate in injection wells located 22 miles from the plant.

In 2001, EPWU initiated research directed toward removing and/or controlling silica during membrane desalination 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 over 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 (residuals) handling problems, so alternative concentrate recovery methods were pursued. In 2007, the TWDB funded a project to investigate using vibratory shear enhanced processing (VSEP) and SWRO batch-treatment systems for recovering water from the KBH concentrate. Both systems were able to achieve recoveries of over 85%, but the SWRO system was more economically attractive than the VSEP system, so VSEP testing was discontinued.

The initial SWRO tests were carried out in a batch-treatment mode using a Crane seawater RO unit with a 2 ½” membrane (Ning et al., 2010). In 2009, the TWDB amended its contract with EPWU that allowed constructing a fully-automated batch-treatment SWRO pilot plant with four 4-inch membrane elements in parallel configuration. The batch-treatment pilot plant was operated at 700 pound per square inch (psi) of operating pressure for 6 months at concentrate recoveries in the 85% range (with silica concentrations reaching 1,000 mg/L in the concentrate) without fouling the membranes, and at a unit water cost comparable to that of the main desalting plant (Tarquin, 2010). When recovery was increased to around 90%, precipitation of calcium sulfate occurred, which became the constituent of concern (rather than silica) for recovering additional water from the KBH concentrate.

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.

## **1.2 Purpose of the study (Scope of Work)**

In its combined form, silicon is the most abundant element next to oxygen in the earth’s crust. Processes for removing silica have been known for many years. Silica is a big problem when present in water that is used to produce high-pressure steam, because it precipitates as a glassy deposit on turbine blades. Silica has also been shown to be a major problem in RO systems, because if it is allowed to reach saturation, it precipitates on RO membranes. There are three general ways to deal with high silica waters in reverse osmosis systems: (1) create an environment within the RO membrane wherein the silica either will not precipitate, or if it does, encourage it to precipitate in a form that will not stick to the surface of the membrane, (2) remove the silica before it reaches the RO membrane, and (3) pre-treat the water in a way that will significantly increase the solubility of the silica in the brine solution. Although the Project Investigator (PI) for this project has pilot plant experience with all three of these methods, only the first one mentioned above (discourage precipitation, especially on the surface of the membrane) was studied during this investigation.

The scope of work for this project included four major tasks. Task 1 required designing and building a continuous-flow seawater reverse osmosis (SWRO) concentrate recovery system in the form of a small, pilot system. Task 2 involved operating the SWRO system and analyzing water samples at various recoveries (beginning at 70%), and raising the recovery in approximately 5% increments until membrane fouling occurred. Task 3 included analyzing the data and evaluating system performance and cost effectiveness. Finally, Task 4 called for preparing quarterly, draft and final reports to the funding agencies.

### **1.3 Organization of the report**

This report is organized in the fashion described in the Table of Contents of the report. Chapter 1 (Introduction) describes the need for the work. Chapter 2 (Summary of literature and technology) includes previous research conducted by other investigators attempting to reduce the volume of concentrate from membrane water treatment processes. This chapter also defines the chemical composition of the RO concentrate at El Paso's KBH Water Plant, which is the site for this research project. Chapter 3 (Experimental set-up and research methods) describes the assembly, arrangement and type of equipment used for the pilot testing. Chapter 3 also defines the Quality Assurance and Quality Control methods employed during the project. Chapter 4 (Operational problems encountered) explains various problems and equipment failures experienced during the research. Chapter 5 (Data collection, data analysis, and findings) includes the real substance of this report because it describes the test results observed at various pilot system water recovery rates, the system's mass balance, and the membrane rejection of various chemical constituents of the raw water source. Chapter 6 (Membrane cleaning) describes the chemicals and techniques used to clean the membranes after they experienced fouling, primarily due to calcium sulfate. Chapter 7 (Operating modifications) discusses the modifications made to the pilot system in an attempt to increase process recoveries and to reduce the amount of treatment down time and equipment malfunctions. Chapter 8 (Economic considerations and evaluation of system performance) lists the unit values for various process components, equipment and chemicals used to estimate overall costs for a full-scale, Continuous Flow SWRO system, and the equivalent unit production cost for the additional potable water available for sale using this method. Finally, Chapter 9 (Conclusions) provides conclusion of this research project. This chapter also provides recommendations for possibly developing a membrane treatment method capable of operating at higher overall water recoveries and with fewer equipment problems.

## **2 Summary of literature and technology**

### **2.1 Calcium sulfate scaling on RO membranes**

The batch process of the SWRO pilot testing encountered calcium sulfate scaling on the membrane surface. Therefore, prior to starting the continuous SWRO pilot test process a brief literature search was conducted on calcium sulfate scaling on the membrane surface.

Calcium sulfate scale formation on the membrane surface could be homogeneous or heterogeneous. In the homogeneous process, ions of opposite charge cluster together in large groups to form nuclei, which ultimately give way to the growth and formation of crystalline precipitates. In heterogeneous precipitation, nuclei associate with suspended or colloidal particles in solution, and precipitate on the membrane surface. In the past decades, several studies were performed to identify the nature of the homogeneous and heterogeneous precipitation of calcium sulfate. Sarig and Mullin (1982) reported that induction periods for the precipitation of calcium sulfate 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 calcium sulfate decreased when either temperature or supersaturation increased. Alimi et al. (2003) found that the induction period was strongly dependent on the solution supersaturation and temperature, with the activation energy decreasing with increasing supersaturation and temperature. Shih et al. (2005) found that calcium sulfate 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 calcium sulfate scaling on membranes is controlled by both surface/heterogeneous crystallization and by deposition of bulk materials (Mi and Elimelech, 2010). Mi and Elimelech (2010) found that calcium sulfate scaling on a polyamide membrane was dominated by heterogeneous/surface crystallization while calcium sulfate scaling of a cellulose acetate membrane was dominated by bulk crystallization and subsequent particle deposition.

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 (HERO) and Optimized Pretreatment and Unique Separation Technology (OPUS) processes for high product recovery in RO systems. 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 hours under reverse flow conditions with a calcium saturation index of 5.4 without fouling the last element. 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. Corbett et al. (2003) also concluded that the addition of 2 mg/L of sodium hexametaphosphate to RO feed water was successful in avoiding scale at recoveries of 93%. The addition of proprietary antiscalants is perhaps the most common method of calcium sulfate control in RO systems. Antiscalants 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 antiscalants, Amjad (1985) concluded that formulated polyelectrolytes were the most effective, with the induction period affected by the antiscalant molecular weight, its concentration, and the nature of the functional groups.

## **2.2 Concentrate characterization**

Concentrate from the KBH desalting plant served as the raw water feed in this project. The overall recovery of the KBH Plant is 80 Percent (%). 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 Average for the Year 2010. As shown in the *% change columns* of the average values, all parameters related to salt content in the water have continued to increase over the past two years. This is due to the general degradation of the raw water supply over time.

The proprietary antiscalant, Pretreat Plus Silica from King Lee Technologies, was 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 the antiscalant in the KBH concentrate should be approximately 20 mg/L, based on an average KBH recovery of 80 %.

**Table 2.1. Characteristics of KBH concentrate**

Parameter*	2008			2009			% Change <sup>+</sup> '08-'09	2010			% Change <sup>+</sup> '09-'10
	Min	Avg.	Max	Min	Avg.	Max		Min	Avg.	Max	
<b>Cl<sup>-</sup></b>	265	4,699	9,710	2,590	5,439	6,740	16%	4,040	6,099	10,100	12%
<b>SO<sub>4</sub><sup>-2</sup></b>	127	1,039	2,110	1,020	1,410	2,260	36%	1,020	1,410	2,260	0%
<b>ALK-T</b>	418.8	417	498	400	427	445	3%	130	397	477	-7%
<b>EC</b>	10,400	16,267	22,100	8,740	18,712	21,500	15%	10,700	20,565	28,800	10%
<b>Fe-T</b>	0.03	0.13	0.57	0.032	0.121	0.42	-7%	0.06	0.16	0.41	32%
<b>Hard-T</b>	528	2,089	3,030	1,200	2328	3,050	11%	540	2,490	4,330	7%
<b>Mn<sup>+2</sup></b>	0.1	0.16	0.21	0.08	0.18	0.22	11%	0.14	0.21	0.34	17%
<b>PO<sub>4</sub><sup>-3</sup></b>	0.1	0.16	1.26	0.1	0.16	0.49	1%	0.1	0.17	0.96	6%
<b>pH</b>	7.1	8.0	8.3	7.5	7.8	8.0	-2%	7.6	8.0	8.0	3%
<b>Ca<sup>+2</sup></b>	376	589	793	281	640	937	9%	176	722	1,220	13%
<b>K<sup>+</sup></b>	4.9	76	99.7	38.9	77	114	18%	61	78	89	-12%
<b>Mg<sup>+2</sup></b>	0.9	153	208	85.7	171	250	12%	146	191	260	12%
<b>Na<sup>+</sup></b>	172	2,674	4,200	1,490	3,097	4,140	16%	2,710	3,412	4,440	10%
<b>SiO<sub>2</sub></b>	28.7	148	228	23.4	128	173	-14%	36.7	145	187	13%
<b>TDS</b>	6,740	10,412	13,200	5,730	11,520	13,600	11%	9,750	12,763	17,900	11%
<b>Temp</b>	18.9	22.0	26.0	22.6	25.0	26.3	4%	22.1	26.0	26.6	4%

Note:

\*All values are in mg/L except EC (µS/cm), pH (pH units), and Temp (°C)

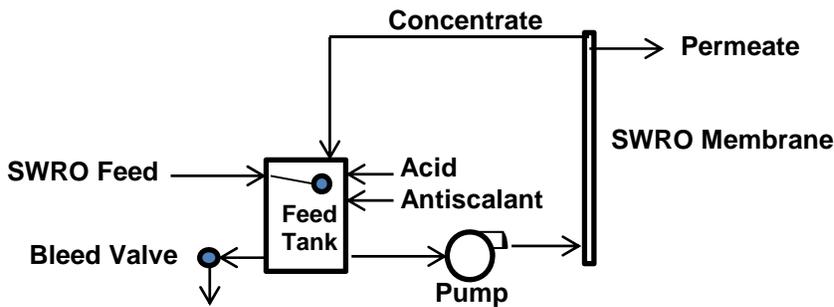
<sup>+</sup>% change is based on average values.

1. Temp refers to Temperature in °C.
2. ALK-T refers to Total Alkalinity.
3. Fe-T refers to Total Iron Concentration.
4. Hard-T refers to Total Water Hardness.

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### 3 Experimental set-up and research methods

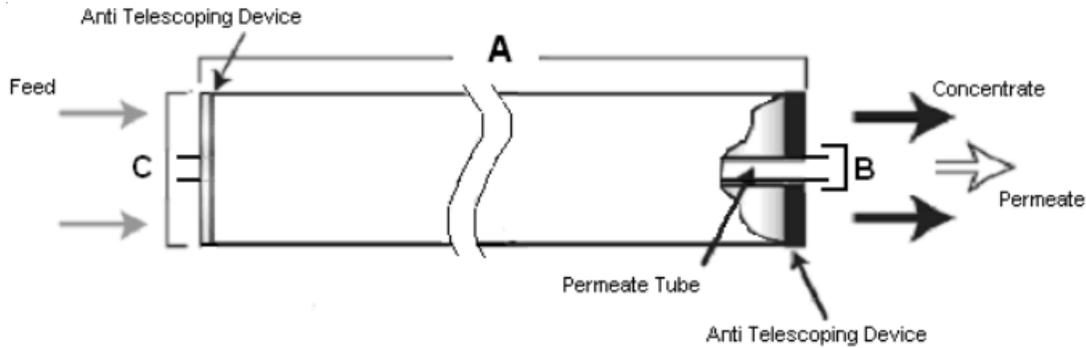
A simplified schematic diagram of the experimental set-up is shown in Figure 3.1. Concentrate from the KBH plant continuously flows into the 200 gallon feed tank through a float valve. An antiscalant for controlling sulfates (Pretreat Plus 400 from King Lee Technologies) is added at a pre-set concentration based on the flow rate into the feed tank (i.e. 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 set-up.**

Sulfuric acid was added to the system to ensure that there would be no fouling due to carbonates. In RO systems, either sulfuric or hydrochloric acid are used to control the pH of the feed. In this study, sulfuric acid was selected because sulfuric acid is cheaper than hydrochloric acid. Acid addition was automatically controlled by feeding a 25% sulfuric acid solution to maintain the pH at a pre-set value of 3.9.

The feed tank was equipped with a pH probe, and a conductivity probe. The conductivity probe enabled the feed tank to be maintained at any pre-set conductivity value by opening and closing the bleed valve as necessary. The pressure vessel housed a single GE-Osmonics seawater RO thin-film membrane Model AD 4040FM. The active area of the membrane was 86 square foot (sf) with an average NaCl rejection capacity of 99.6%. The operating parameter of maximum pressure drop over a single element was 12 psi. A schematic diagram of the membrane is shown in Figure 3.2, with the membrane length (A) equal to 40 inches, the diameter of the permeate tube (B) equal to 0.75 inch, and the outside diameter of the membrane (C) equal to 3.9 inches.



Source: Fact Sheet, AD HR Series, GE Water & Process Technologies

**Figure 3.2. Schematic of SWRO membrane.**

The high-pressure feed pump was a Cat Pumps Model 820 triplex positive displacement pump that is driven by a 7.5 horsepower motor. A flow meter in the permeate discharge line allowed for the system to be operated in a constant-permeate-flow mode (i.e. the speed of the pump was automatically adjusted to maintain the pre-set permeate flow rate). At the beginning of the project, the permeate flow rate was set to 0.5 gallons per minute (gpm), which translated to a membrane flux of 9 gallons per square foot 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 one-minute intervals anytime the system is running in the automatic mode (i.e. 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 3.1 below.

The reader should note that on Table 3.1 the values for “Concentrate Flow” far exceed those shown for “Flow into the Feed Tank” or “Permeate Flow” because the booster pump forcing the concentrate through the membrane maintains a much higher flow rate than the rate at which the feed water is introduced into the Feed Tank. Consequently, the values for “Concentrate Conductivity” on Table 3.1 appear to be rather low, because each pass of the concentrate feed through the membrane only slightly increases the TDS of the concentrate.

**Table 3.1. Sample of data sheet (20 minutes on June 24, 2010)**

Time Stamp	Flow into the Feed Tank (gpm <sup>1</sup> )	Permeate Flow (gpm <sup>1</sup> )	Concentrate Flow (gpm <sup>1</sup> )	Feed Tank Cond. <sup>2</sup> (μS/cm) <sup>3</sup>	Permeate Cond. <sup>2</sup> (μS/cm) <sup>3</sup>	Concentrate Cond. <sup>2</sup> (μS/cm) <sup>3</sup>	Feed Temp. <sup>4</sup> (°C)	pH in Feed Tank	Feed Pressure (psi) <sup>5</sup>	Concentrate Pressure (psi) <sup>5</sup>
12:00	0.7	0.49	6.84	33339	1333	34530	29.7	3.7	377	375
12:01	0.7	0.51	6.79	33211	1321	34188	30.3	3.7	387	364
12:02	0.7	0.51	6.79	33571	1368	34676	29.5	3.8	383	365
12:03	0.7	0.5	6.87	33468	1358	34334	30.2	3.8	371	371
12:04	0.7	0.49	6.86	33417	1331	34383	30.5	3.8	381	372
12:05	0.7	0.49	6.84	33339	1333	34554	30.1	3.8	384	369
12:06	0.7	0.5	6.86	33622	1338	33968	31.4	3.8	392	369
12:07	0.7	0.51	6.86	33185	1363	34407	30.5	3.7	386	366
12:08	0.6	0.5	6.83	33725	1363	34383	31.3	3.8	378	370
12:09	0.7	0.49	6.87	34033	1365	34432	30.6	3.7	392	369
12:10	0.7	0.49	6.83	33417	1358	34554	30.6	3.8	391	371
12:11	0.7	0.49	6.87	33211	1355	34994	29.8	3.8	387	371
12:12	0.6	0.49	6.89	33931	1350	34652	30.7	3.8	391	378
12:13	0.7	0.5	6.83	33828	1387	35189	30.1	3.8	395	375
12:14	0.7	0.49	6.84	34085	1375	35042	30.2	3.8	389	377
12:15	0.7	0.48	6.89	33648	1350	35531	29.8	3.8	382	377
12:16	0.6	0.49	6.88	33699	1350	35628	29.9	3.8	388	374
12:17	0.7	0.49	6.9	33828	1358	34627	31.5	3.8	395	379
12:18	0.7	0.49	6.94	34085	1331	35067	30.7	3.8	393	382
12:19	0.7	0.5	6.88	34291	1355	36166	29	3.7	392	381
12:20	0.6	0.49	6.93	34008	1370	35311	30.6	3.8	405	382

*Note:*

1. gpm refers to gallons per minute
2. Cond. refers to conductivity
3. μS/cm refers to microSiemens per centimeter
4. Temp. refers to temperature
5. psi refers to pounds per square inch

The chemical analyses included in this report were performed at the International Water Quality Laboratory (IWQL) of El Paso Water Utilities. After samples were collected at the research site (i.e. KBH Desalting Plant Laboratory), a Chain of Custody form was filled out and the samples were transported to the IWQL within two hours. The QA/QC procedures associated with the test results are contained in section 23 of their 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 Company) 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 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.

Since 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 low recoveries (i.e. less than 50%) in order to get familiar with the vagaries of the system and to acclimatize the system. 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 antiscalant dosage, but none were 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* that is specific for silica and *Diamite CAL* that 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.

## **4 Operational 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 data-logging 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 replacing it with a completely new unit and getting the same result, it was determined that the Adjustable Frequency Drive (AFD) 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**

During the study, a problem with the water-level switch was identified: it indicated that the water level was above the sensor even when the tank was empty. This would allow the high-pressure pump to run even when there was no water in the tank (i.e. in a dry condition), 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 running 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 data-logging systems were not functioning smoothly. After the contractor re-calibrated 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 located 13 miles from the University of Texas at El Paso.

#### **4.5 Problems with the pH sensor**

Approximately one-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 located on the suction side of the pump, and it used to measure the pH of feed water to the membrane. It was also connected to the acid feed pump and was used to control the pH of the feed water. Therefore, it was 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 located (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 was located on the discharge side of the positive displacement pump and it used to protect the unit from over-pressurization. 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**

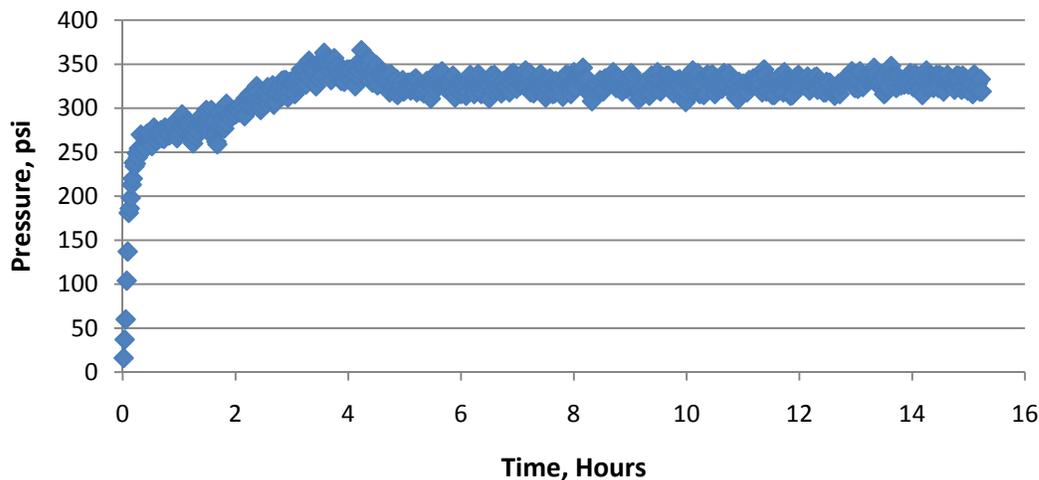
The low-pressure feed pump that provided water to the feed tank was a residential water pressure booster pump. Although it was not made to handle salty water, the contractor thought it would last for at least the duration of the project. It did not, and neither did a subsequent replacement. The third 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.

Most of the problems encountered in conducting this research were specific to this project. Operational problems encountered in this research project illustrates that as pilot plants increase in complexity by incorporating automatic controls and data logging, more frequent process interruptions will likely occur.

## 5 Data collection, data analysis, and data 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 the system 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 one 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 hours* after start-up at a recovery of 48%. The pressure was essentially constant after the feed tank conductivity reached the pre-set value of 30,000 microSiemens per centimeter ( $\mu\text{S}/\text{cm}$ ) (corresponds to a TDS of approximately 20,000mg/L), approximately 4 hours after the system was started. For the purposes of this research project, since the feed tank and discharge point are both operated at atmospheric conditions, the terms “transmembrane” pressure and “feed” pressure are essentially the same.

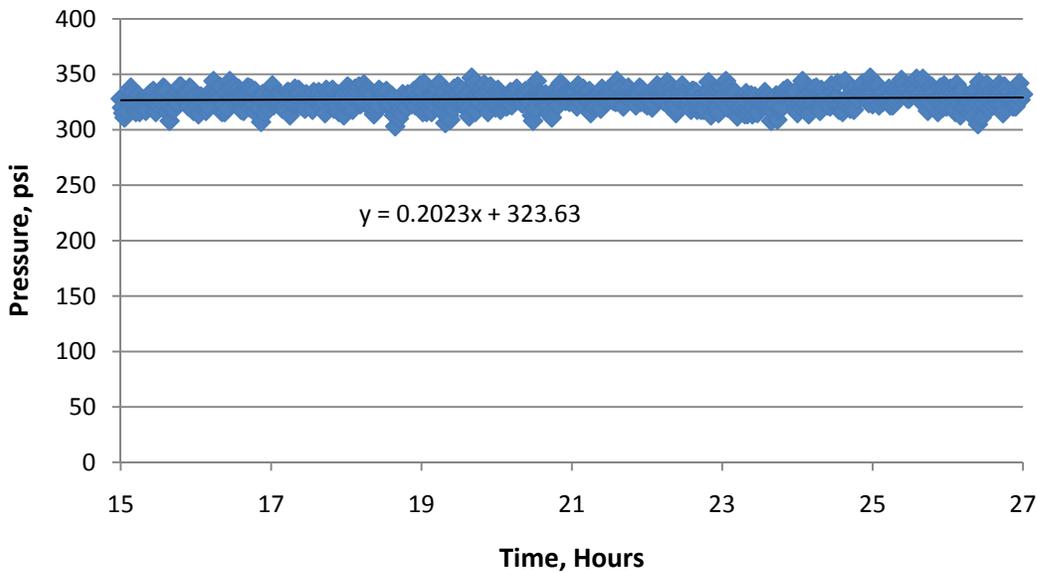


**Figure 5.1. Transmembrane pressure for first 15 hours after start-up at 48% recovery, with a pre-set conductivity value of 30,000  $\mu\text{S}/\text{cm}$ , in terms of feed tank concentration.**

Figure 5.2 is a plot of transmembrane pressure versus time *after the first 15 hours* of run time and the pressure was already starting to increase as evidenced by the slight upward slope (i.e.

0.2023) of the best fit line. The pilot membrane system was designed to automatically shut down operation at a prescribed, or pre-set, 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 that there was 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 the 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 (i.e. the raw feed water to the SWRO unit), the permeate, and in the concentrate from the SWRO pilot plant. The rejection for all parameters was very high, averaging over 98%, yielding a very high quality permeate having a TDS concentration of 268 mg/L.



**Figure 5.2. Transmembrane pressure *after first 15 hours* at 48% recovery, with a pre-set conductivity value of 30,000  $\mu\text{S}/\text{cm}$ , in terms of feed tank concentration.**

**Table 5.1. Results at 48% recovery**

<b>*Parameter</b>	<b>SWRO Feed</b>	<b>SWRO Permeate</b>	<b>SWRO Concentrate</b>	<b>% Rejection</b>
<b>Hardness</b>	2,610	5	5,250	99.8%
<b>Ca<sup>+2</sup></b>	570	1	1,060	99.8%
<b>Mg<sup>+2</sup></b>	186	1	364	99.5%
<b>Na<sup>+</sup></b>	2,950	66	5,780	97.8%
<b>K<sup>+</sup></b>	86	2.7	164	96.9%
<b>Cl<sup>-</sup></b>	6,050	167	11,200	97.2%
<b>SO<sub>4</sub><sup>-2</sup></b>	1,800	23	3,320	98.7%
<b>Silica</b>	155	2	266	98.7%
<b>Conductivity</b>	19,875	541	36,600	97.3%
<b>TDS</b>	12,970	268	24,100	97.9%

Note:

\*All values in mg/L except conductivity ( $\mu\text{S}/\text{cm}$ ) and hardness (mg/L of  $\text{CaCO}_3$ ).

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 is probably because the samples were analyzed at EPWU's International Water Quality Laboratory, where the turn-around is normally two weeks. The supersaturated silica in the SWRO concentrate polymerizes with time, and the test that measures silica only measures 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.

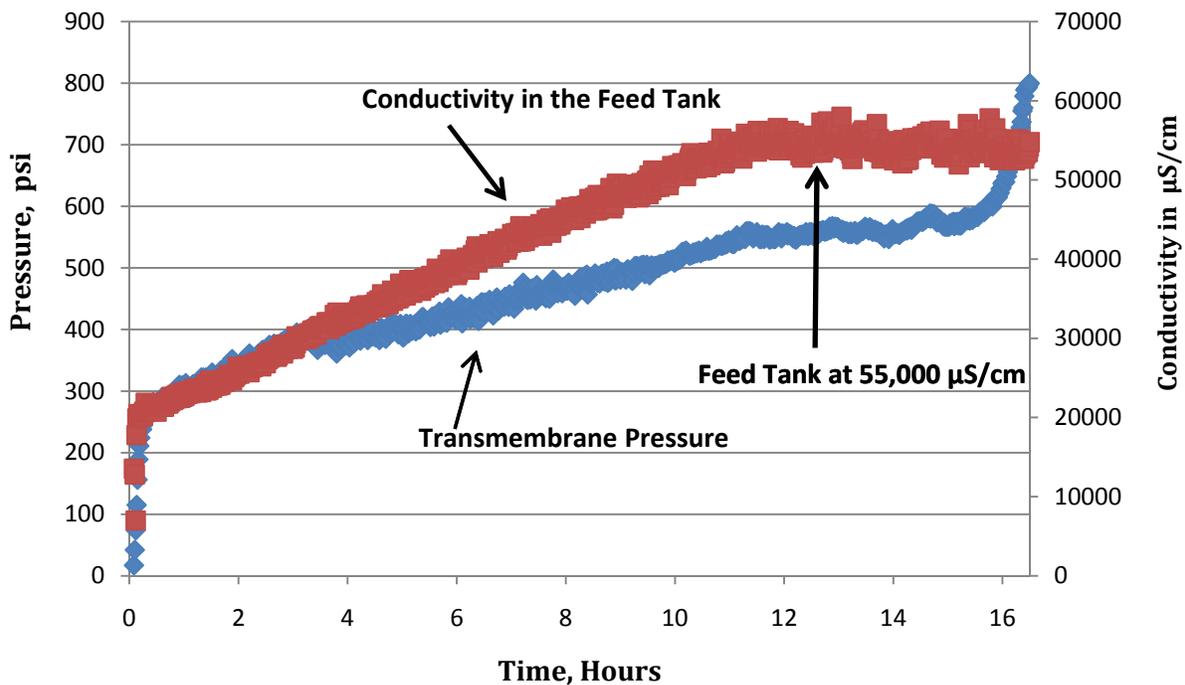
**Table 5.2. Mass balance at 48% recovery**

<b>*Parameter</b>	<b>Mass In</b>	<b>Permeate Mass</b>	<b>Concentrate Mass</b>	<b>Permeate and Concentrate</b>	<b>Diff</b>	<b>% Diff</b>
<b>Hardness</b>	10,620	10	10,586	10,595	-25	0%
<b>Ca<sup>+2</sup></b>	2,319	2	2,267	22,69	-50	-2%
<b>Mg<sup>+2</sup></b>	757	2	778	780	23	3%
<b>Na<sup>+</sup></b>	12,003	128	12,361	12,489	486	4%
<b>K<sup>+</sup></b>	350	5	351	356	6	2%
<b>Cl<sup>-</sup></b>	24,617	322	23,951	24,274	-343	-1%
<b>SO<sub>4</sub><sup>-2</sup></b>	7,324	45	7,100	7,145	-179	-2%
<b>Silica</b>	631	4	569	573	-58	-9%
<b>Conductivity</b>	80,869	10,44	78,270	79,314	-1,555	-2%
<b>TDS</b>	52,773	517	51,538	52,056	-717	-1%

Note:

Mass values are in mg/min, except Conductivity ( $\mu\text{S}\cdot\text{L}/\text{cm}\cdot\text{min}$ ) and hardness as mg/min of  $\text{CaCO}_3$ .

Figure 5.3 is a plot of transmembrane pressure and conductivity versus time for a feed tank conductivity setting of 55,000  $\mu\text{S}/\text{cm}$ , which corresponds to a TDS of approximately 35,000 mg/L. This test case resulted in a recovery of 74%. It took approximately 12 hours for the feed tank conductivity to reach the pre-set value of 55,000  $\mu\text{S}/\text{cm}$ , as shown on the graph. Thereafter, the transmembrane pressure remained relatively constant (at about 530 psi) for only 3 hours, after which it steadily increased. During the last 30 minutes of operation, the transmembrane pressure increased very rapidly, going from 600 psi to 800 psi, causing the system to automatically shut down due to high pressure.



**Figure 5.3. Time vs. transmembrane pressure and feed tank conductivity.**

The chemical analyses of the SWRO permeate and concentrate are shown in Table 5.3. The rejection of the divalent ions remained high at over 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 or not the fouling and subsequent chemical cleaning had anything to do with the lower than anticipated rejection rates is not known.

**Table 5.3. Results at 74% recovery**

<b>*Parameter</b>	<b>KBH Conc</b>	<b>SWRO Perm</b>	<b>SWRO Conc</b>	<b>% Rejection</b>
<b>Hardness</b>	2,010	99	7,260	95.1%
<b>Ca<sup>+2</sup></b>	538	26.4	1,960	95.1%
<b>Mg<sup>+2</sup></b>	142	5.9	505	95.8%
<b>Na<sup>+</sup></b>	2,560	236	9,100	90.8%
<b>K<sup>+</sup></b>	57.6	7.8	223	86.5%
<b>Cl<sup>-</sup></b>	4,510	465	15,800	89.7%
<b>SO<sub>4</sub><sup>-2</sup></b>	1,050	50.3	4,980	95.2%
<b>Silica</b>	107	6.1	342	94.3%
<b>TDS</b>	9,600	850	34,200	91.1%
			<b>Average</b>	<b>92.6%</b>

*Note:* \*All concentrations in mg/L; hardness is expressed as mg/L of CaCO<sub>3</sub>

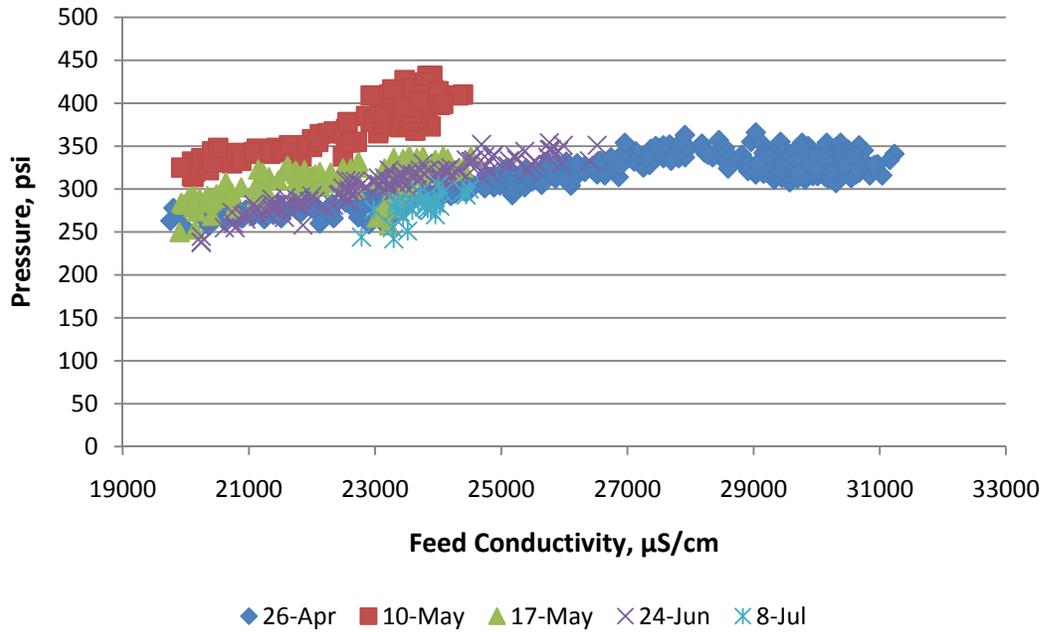
The results were the same for all recoveries above 48% (i.e. conductivities above 30,000  $\mu\text{S}/\text{cm}$ ), so the conductivity set point was reduced to 27,500  $\mu\text{S}/\text{cm}$ , which represents a recovery of about 35%. The system operated at this recovery almost continuously over a time span of 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 was the highest recovery for which no pressure increase was observed. This disappointingly low recovery led to modifications of the operating conditions as discussed in Chapter 7.

## 6 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* (specific for calcium sulfate) cleaning solution at a 1:40 dilution was used. The solution was circulated through the membrane for about one hour and then allowed to soak overnight. Recirculation was resumed the next morning for about 10 minutes (the solution had turned yellow-orange by that time), and then the membrane was flushed with RO permeate. When the system was restarted, the pressure was down to its pre-fouled values. The result is shown in Figure 6.1.

Figure 6.1 shows the transmembrane pressures for 5 different events: (1) prior to any membrane fouling events (i.e. 26-Apr), (2) while the membrane was fouled (i.e. 10-May), and (3) after the membrane had been cleaned following three different fouling events (i.e. 17-May, 24-Jun, and 8-Jul). 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 possibly 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 those 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  $\text{CaSO}_4$ , all the nucleation sites may not be removed during cleaning. Thus, the membrane may subsequently foul quicker under super-saturation conditions due to the presence of these nucleation sites, resulting in fouling of the cleaned membrane surface faster than fouling of a new one.



*Note:*

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

**Figure 6.1. Transmembrane pressures before and after chemical cleaning.**

## 7 Operating modifications

In an attempt to get recoveries in the range of at least 50%, several modifications were made to the operating conditions of the pilot plant. Since calcium sulfate was apparently the foulant that was limiting recovery in the system, two steps were taken to address this problem. The first step 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%. Since the solubility of calcium sulfate decreases with increasing temperature, a heat exchanger (fashioned out of stainless steel tubing) was installed in the feed tank to keep the temperature below 26°C. However, precipitation occurred again in day two, even at the lower conductivity setting of 30,000 µS/cm. Since addition of a heat exchanger was not able to control salt precipitation in the system, the second step was implemented to control the precipitation. The second step was to replace the stainless steel tubing with an all-plastic heat exchange unit, but precipitation still occurred on the second day.

The second modification of the system involved increasing the antiscalant feed rate so that its concentration would raise from 4 mg/L to 20 mg/L. The increased concentration did not work, as precipitation occurred again within two days.

The third modification involved reducing the membrane flux from 9 gfd 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 gallons to 15 gallons. Researchers hoped that the shorter retention time in the feed tank would decrease the tendency for calcium sulfate to precipitate, but it did not. Thus, none of the changes in operating conditions beneficially affected system performance from the standpoint of reduced membrane fouling.

The main problem appears to be associated with the design of the treatment system itself. That is, the feed tank is *maintained* at a solids concentration that is above 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, which ultimately fouled the membrane. In the batch-treatment system, which was the fore-runner to this project, the supersaturated condition lasted for only a relatively short period of time (i.e. less than the induction time for calcium sulfate precipitation) before the super-concentrate was dumped, and thus the precipitation problem was avoided. It would seem that a possible solution to the fouling problem in a continuous flow SWRO system is to run the system in a single-pass mode with no RO concentrate recirculation back to the feed tank.

## **8 Economic considerations and evaluations 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 the KBH Plant in El Paso, Texas. Nevertheless, researchers and engineers investigating projects at other locations exhibiting conditions similar to those existing in El Paso, should be able to make prudent adjustments in terms of scale and regional wages and other cost components in order to derive estimated costs for water produced 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 on 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/kgal selling price for water represents the approximate cost to EPWU for obtaining, treating, and delivering a new source of water to EPWU customers. This represents a conservative (i.e. low) estimate of income because a substantial amount of water is sold at higher prices, especially in the summer months. Cost-related components that are not included in this project that may have to be included in projects considered elsewhere are land costs, pipeline costs, and solids disposal costs. These components are not included herein because of circumstances that are unique to this project.

**Table 8.1. Unit values for calculating cost of continuous flow SWRO system at 35% recovery**

<b>Item</b>	<b>Value</b>
<b>Initial RO concentrate volume (gpd<sup>1</sup>)</b>	3,600,000
<b>Interest rate (%)</b>	6%
<b>Evaporation rate (in/yr)</b>	50
<b>Liner cost (\$/sq ft)</b>	0.73
<b>Liner life (yrs)</b>	20
<b>Excavation (\$/cubic yard)</b>	\$3.00
<b>Fence (\$/LF)</b>	\$10.00
<b>Flow storage (months)</b>	6
<b>Excavation and fence amortization time (yrs)</b>	20
<b>Equipment life (yrs)</b>	20
<b>Water selling price (\$/1000 gal)</b>	\$2.00
<b>Blending well depth (ft)</b>	300
<b>Blending ratio (blend volume/total volume)</b>	33%
<b>SWRO membrane life (yrs)</b>	5
<b>RO capital cost (\$/MGD)</b>	\$1,000,000
<b>Power cost (\$/kw-hr)</b>	\$0.08
<b>Pump &amp; Motor efficiency (%)</b>	75%
<b>Building cost (\$)</b>	\$320,000
<b>RO membrane cost (\$/ sq ft)</b>	\$2.00
<b>SWRO flux (gpd<sup>1</sup>/sq ft)</b>	9
<b>SWRO operating pressure (psi)</b>	300
<b>Antiscalant dosage (ppm<sup>2</sup>)</b>	4
<b>Antiscalant (\$/gal)</b>	\$11
<b>Sulfuric acid cost (\$/gal)</b>	\$2.53
<b>Sulfuric acid feed rate (mL/gal)</b>	1.0
<b>Personnel cost (\$/year)<sup>3</sup></b>	\$105,000
<b>Contingencies (\$/yr)</b>	\$100,000

*Note:*

1. gpd refers to gallons per day
2. ppm refers to parts per million
3. Assumed three personnel for the plant at \$35,000 per year for each person

**Table 8.2. Total values for calculating cost of continuous flow SWRO system at 35% recovery**

<b>Item</b>	<b>Value</b>
<b>Membrane area (sq. ft)</b>	140,000
<b>RO equipment capital cost (\$)</b>	\$1,260,000
<b>Amortized RO equipment cost (per year)</b>	\$109,853
<b>Membrane cost (\$)</b>	\$280,000
<b>Amortized membrane cost (\$ per year)</b>	\$66,471
<b>Flow rate to RO (cubic feet per second)</b>	5.54
<b>Horsepower for RO unit (HP)</b>	579
<b>Power cost (\$)</b>	\$304,327
<b>Disposal volume (gallon per day)</b>	2,340,000
<b>Pond area (sq. ft)</b>	27,262,872
<b>Pond volume (cubic feet)</b>	56,953,260
<b>Liner cost (\$)</b>	\$19,901,897
<b>Excavation cost (\$)</b>	\$6,328,410
<b>Fencing cost (\$)</b>	\$185,093

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 an amortized Capital, and Operations and Maintenance (O&M) costs, 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 above 35% were proven to be achievable, then more potable water would be available to sell to the public, thus decreasing the cost of the SWRO process. The breakeven 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.

Breakeven 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 \$5.80/1000 gallons and \$2.00/1000 gallons, 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. At higher recoveries, more water is available to sell, and that generates more income. Furthermore, the cost of final disposal of the super-concentrate (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 thousand gallons of product water for each \$1 million of capital cost.

## 9 Conclusions

Based on the results from 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 continuous 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 for preventing calcium sulfate precipitation, including installing a heat exchanger in the feed tank, raising the antiscalant dosage from 4 mg/L to 20 mg/L, reducing the membrane flux from 9 gfd to 5.4 gfd, and decreasing the feed tank volume from 200 gallons to 15 gallons.
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 selected. That is, the RO concentrate flow pattern of the system was such that the solid concentration in the feed tank was maintained same as that of the final concentrate of the system. Since one or more compounds were in a supersaturated state in the final concentrate, precipitation (sooner or later) was inevitable, resulting in a fouled membrane. This is the result of exceeding the induction period previously described in detail in Chapter 7.
5. A better design for a continuous flow system would be a once-through, single-pass 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 full-scale system and would minimize the time that the concentrate with the supersaturated salts stays in the SWRO membrane and the treatment system.
6. At 35% recovery, the cost of the produced water would be \$5.80 /1000 gallons 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 breakeven, 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.

## **9.1 Recommendations**

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 order to avoid exceeding the previously described induction time for calcium sulfate. A single-pass SWRO system should be designed and tested to prove the concept, and then evaluated from the standpoint of economic feasibility.

## **10 Acknowledgements**

This project was made possible through funding provided by The Texas Water Development Board and The WaterReuse Research Foundation. Special thanks go to El Paso Water Utilities for providing the physical facilities and personnel for a variety of tasks including equipment installation, plumbing and electrical work, equipment maintenance/replacement, and extensive laboratory analytical services. Without the unwavering support of many, many individuals at all levels in the Utility (particularly Mr. Ralph Padilla), we would have had no chance of carrying out this project.

The research team also thanks Dr. Bob Ning of King Lee Technologies for providing antiscalants and cleaning solutions, and for always being available to discuss new ideas. The research team would also like to thank Frank Leitz of the United States Bureau of Reclamation for providing valuable technical review of the report.

Finally, the research team would like to thank the project advisory committee for offering their insightful suggestions about how to overcome the obstacles that were encountered along the way. The members of the committee were

Patrick Brady, Sandia National Laboratories

Juan Gomez, CH2M HILL, Inc.

Scott Irvine, Bureau of Reclamation

Keith Rutherford, Parkhill Smith & Cooper, Inc.

Bruce Thompson, University of New Mexico

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# APPENDIX A

## SCREEN SHOTS OF CONTROL PANEL



**Figure A. 1** Main Screen

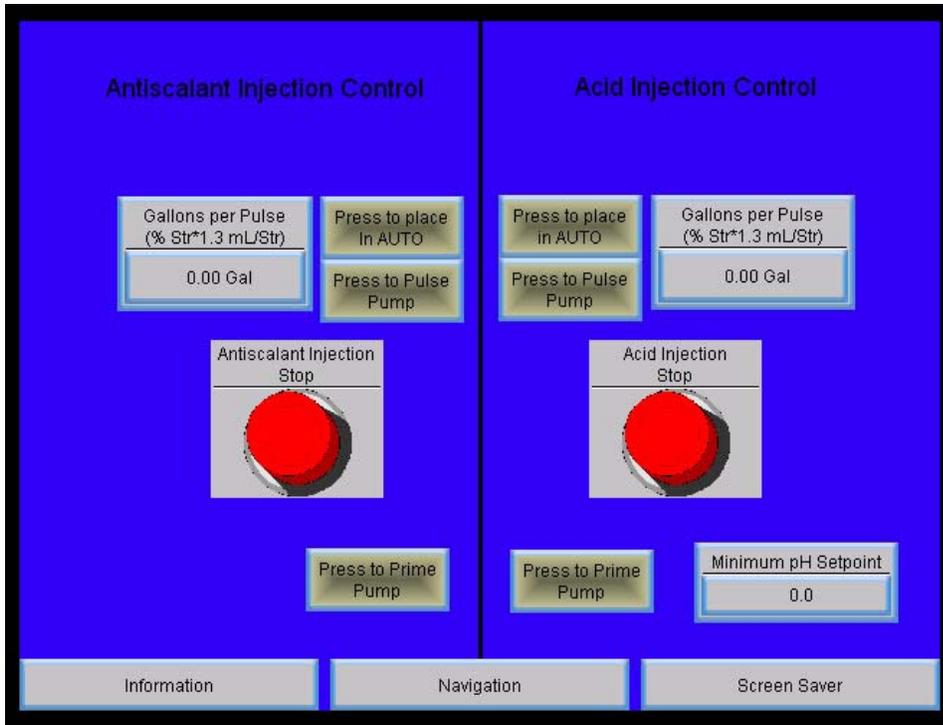


Figure A.2 Chemical Injection Screen

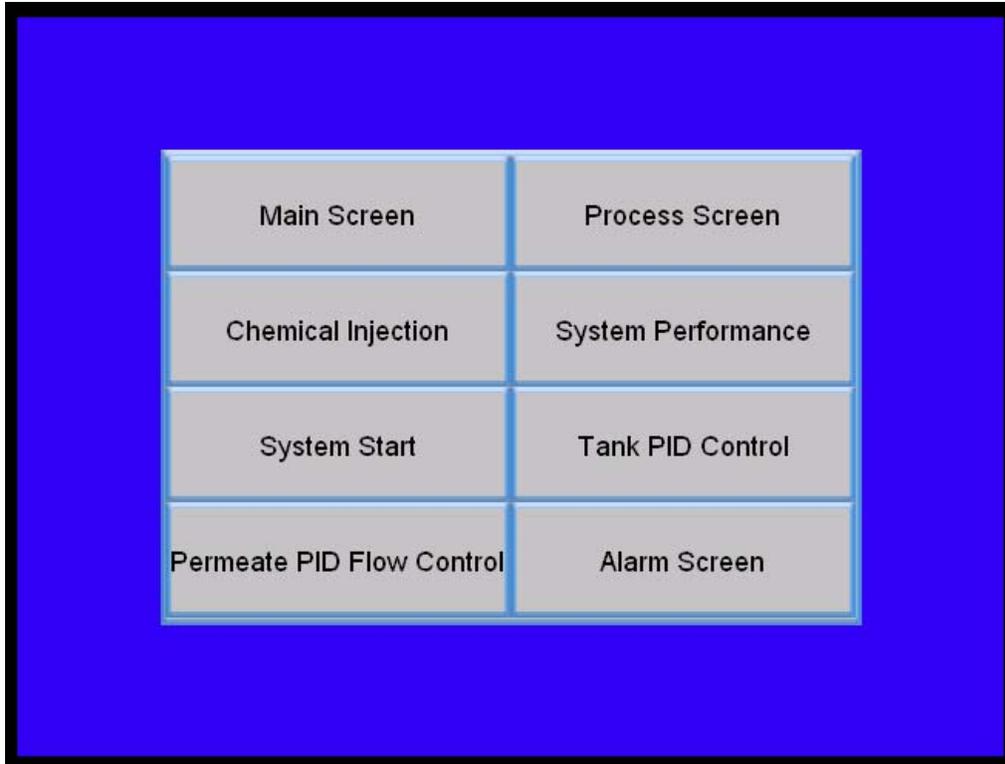


Figure A.3 Navigation Screen

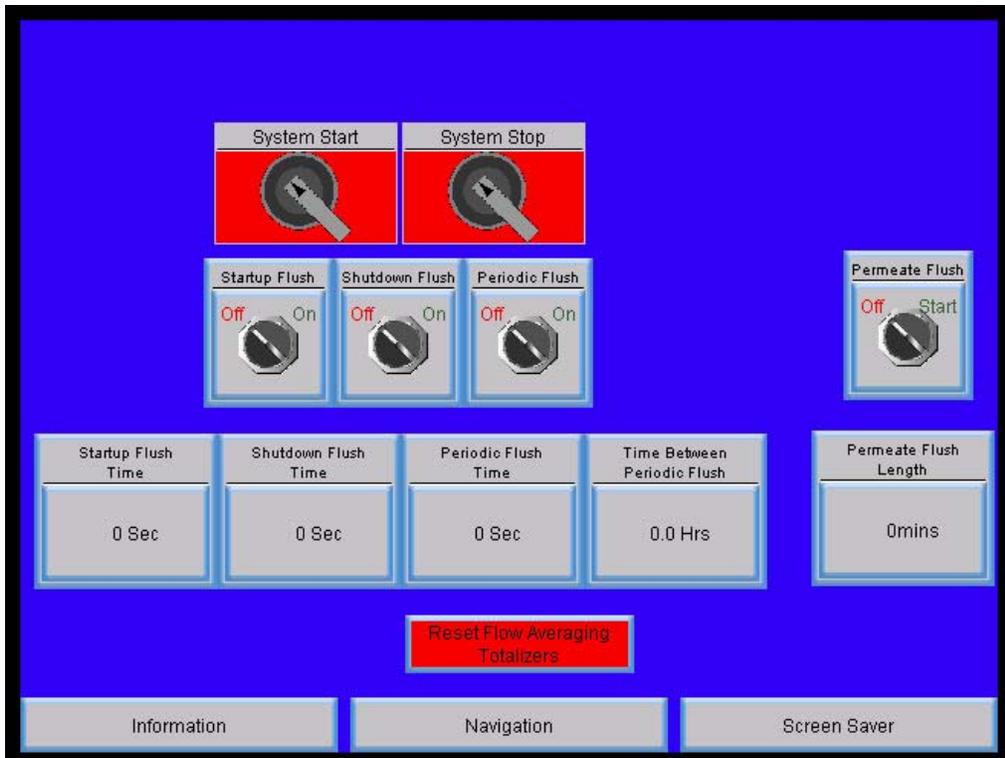
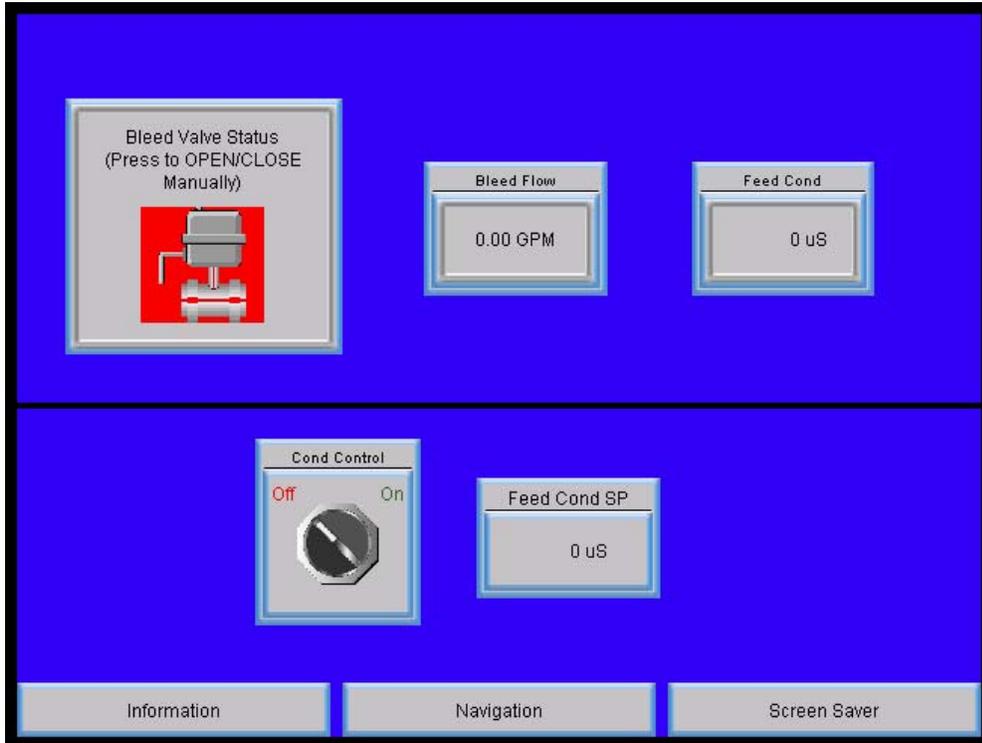


Figure A.4 System Start Screen



**Figure A. 5 Tank PID Control Screen**

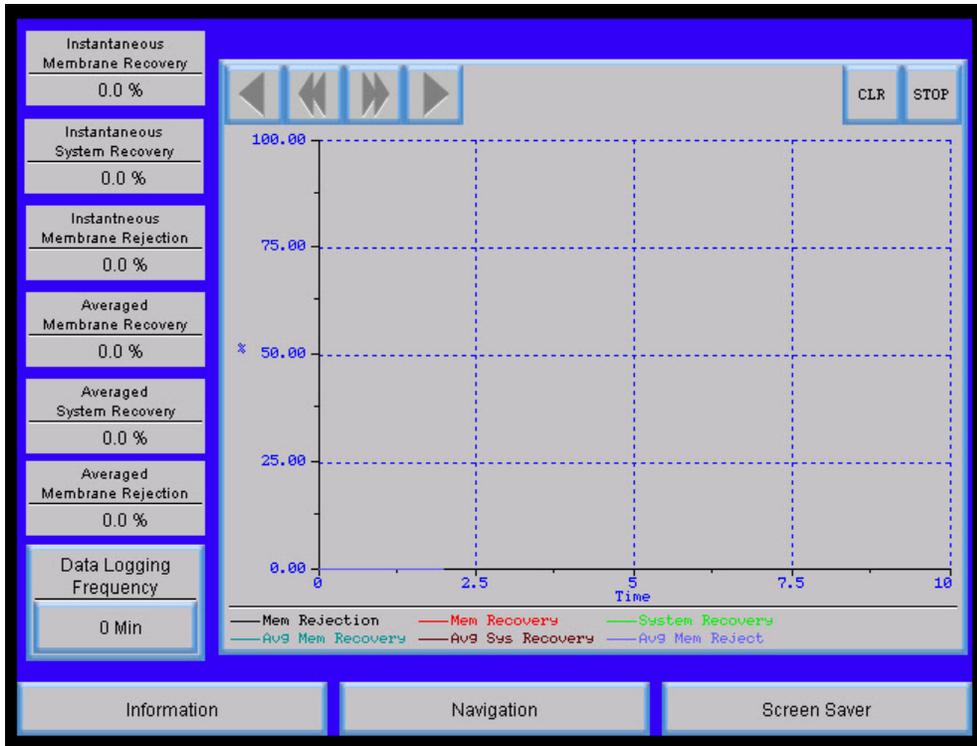
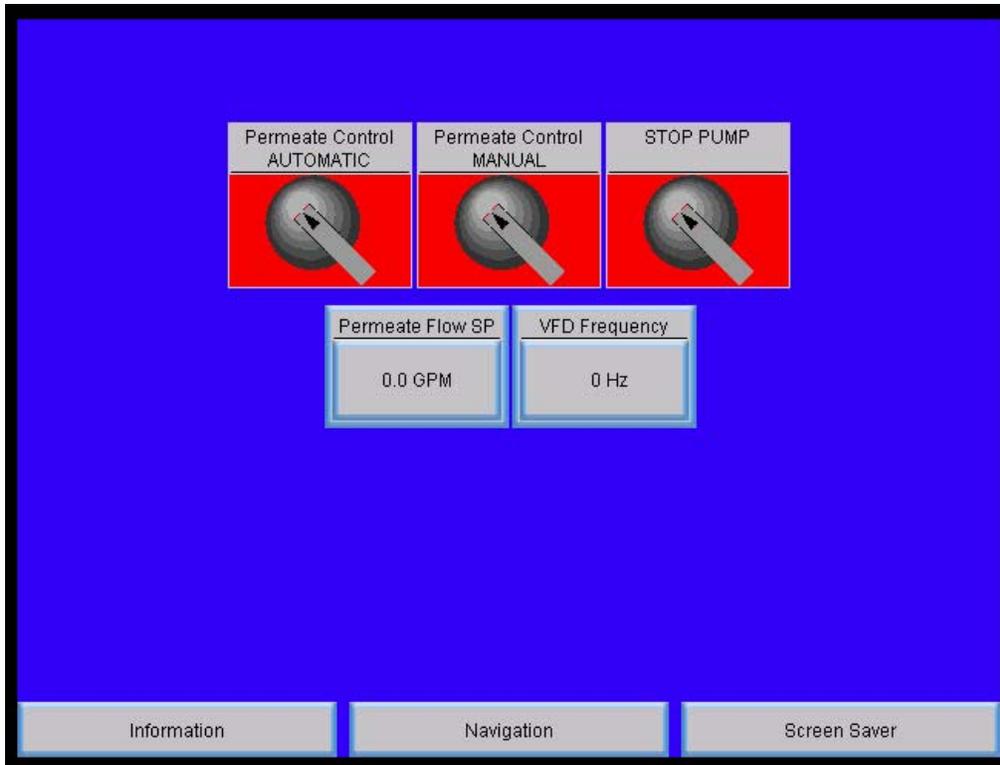


Figure A.6 System Performance Screen



**Figure A. 7 Permeate PID Flow Control Screen**

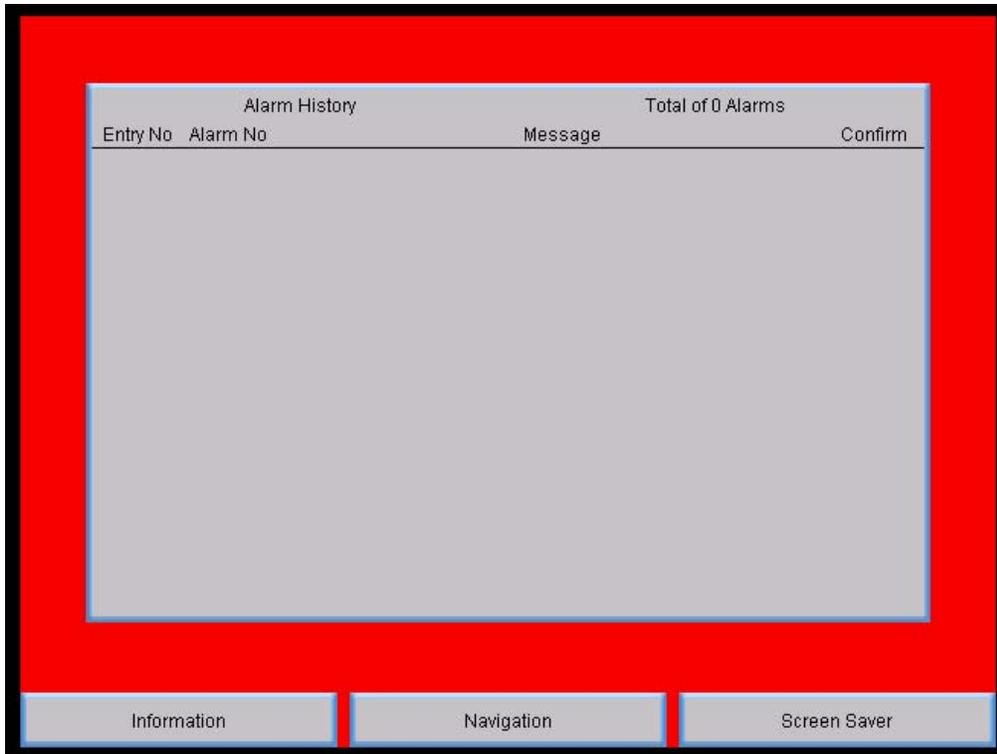


Figure A.8 Alarm Screen

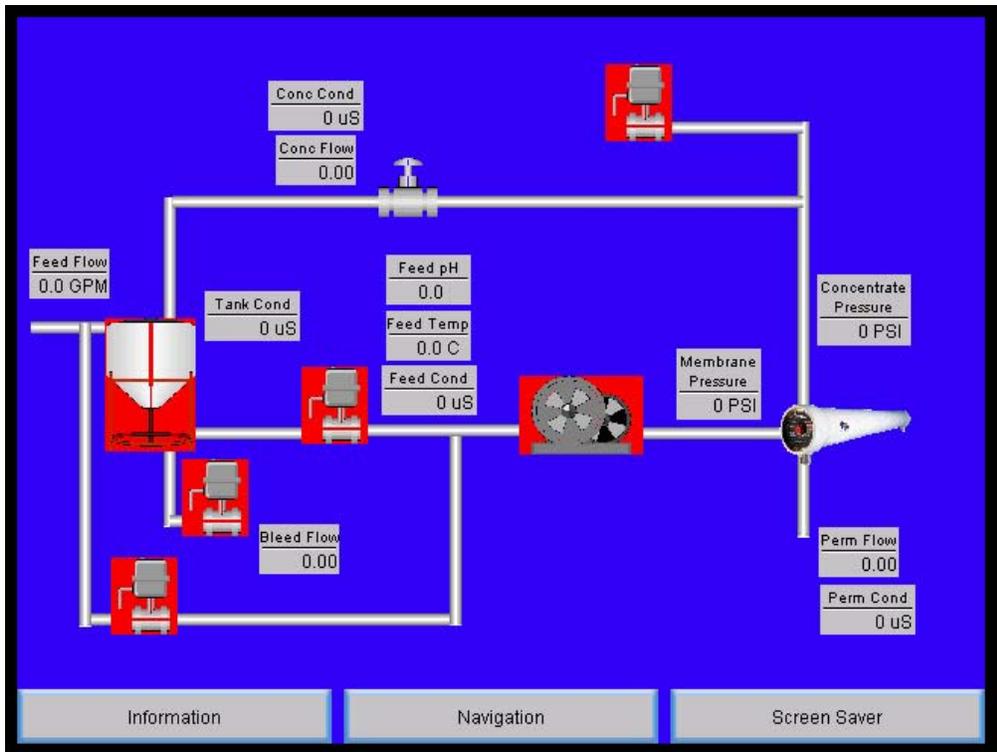
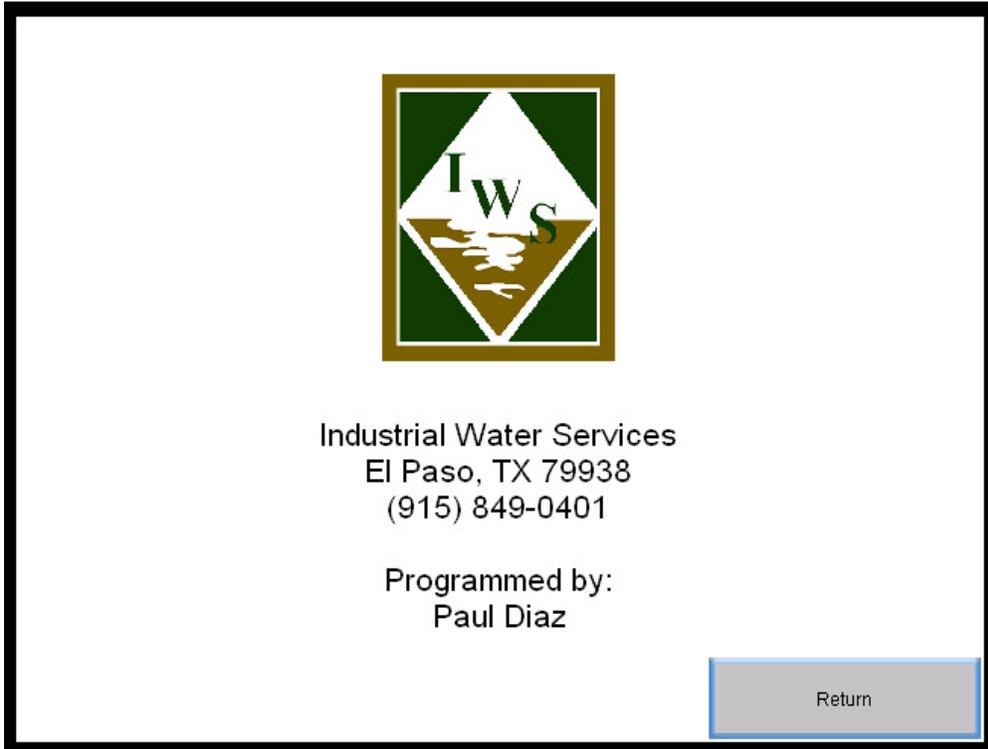


Figure A.9 Process Screen



**Figure A. 10** Information Screen



**Figure A.11** Emergency Stop Screen

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## APPENDIX B

### SAMPLE CALCULATIONS

(used to generate the unit cost of water produced)

$$\begin{aligned}\text{Membrane Area} &= \text{perm volume/flux} \\ &= 1,260,000/9.0 \\ &= 140,000 \text{ ft}^2\end{aligned}$$

$$\begin{aligned}\text{RO Equipment Capital Cost} &= \text{concentrate volume} \times \text{cap cost/MGD} \\ &= 1.26 \times 1,000,000 \\ &= \$1,260,000\end{aligned}$$

$$\begin{aligned}\text{Amortized RO Equipment Cost} &= \text{RO equipment cost} \times (\text{A/P}, 6\%, 20) \\ &= 1,260,000 \times 0.07358 \\ &= \$109,853 \text{ per year}\end{aligned}$$

$$\begin{aligned}\text{Membrane Cost} &= (\text{perm volume/flux}) \times \text{membrane cost/ft}^2 \\ &= (1,260,000/9) \times 2.00 \text{ (membrane cost can be estimated based on unit cost per} \\ &\text{membrane element. A membrane manufacturer can provide a budget quote for SWRO} \\ &\text{membrane elements, i.e. \$550 – 600 per membrane element)} \\ &= \$280,000\end{aligned}$$

$$\begin{aligned}\text{Amortized membrane cost} &= \text{membrane cost} \times (\text{A/P}, 6\%, 5) \\ &= 280,000 \times 0.22463 \\ &= \$66,471 \text{ per year}\end{aligned}$$

$$\begin{aligned}\text{Flow Rate to RO in cfs} &= \text{RO water volume} \times \text{conversion to cfs from gpd} \\ &= 3,600,000 \times 0.133 / (24 \times 60 \times 60) \\ &= 5.54 \text{ cfs}\end{aligned}$$

$$\begin{aligned}\text{Horsepower for RO Unit} &= \text{flow} \times \gamma \times \text{head} / (550 \times \text{Eff}) \\ &= 5.54 \times 62.4 \times 300 \times 2.303 / (550 \times 0.75) \\ &= 579 \text{ HP}\end{aligned}$$

$$\begin{aligned}\text{Power Cost} &= \text{horsepower} \times 0.75 \text{ Kw/HP} \times \text{hrs/day} \times \$/\text{kW-hr} \times \text{days/yr} \\ &= 579 \times 0.75 \times 24 \times 0.08 \times 365 \\ &= \$304,327 \text{ per year}\end{aligned}$$

$$\begin{aligned}\text{Disposal Volume} &= \text{raw concentrate volume} - [\text{raw concentrate volume} \times \text{recovery}] \\ &= 3,600,000 - [3,600,000(0.35)] \\ &= 2,340,000 \text{ gallons per day}\end{aligned}$$

$$\text{Pond Area (ft}^2\text{)} = (\text{disposal volume} \times 0.133 \times 365 / (\text{evaporation rate}/12))$$

$$\begin{aligned} &= 2,340,000 \times 0.133 \times 365 / (50/12) \\ &= 27,262,872 \text{ ft}^2 \end{aligned}$$

$$\begin{aligned} \text{Pond Area (acres)} &= ((\text{disposal volume} \times 0.133 \times 365 / (\text{evaporation rate} / 12)) / 43,560) \\ &= ((2,340,000 \times 0.133 \times 365 / (50/12)) / 43,560) \\ &= 626 \text{ acres} \end{aligned}$$

$$\begin{aligned} \text{Liner Cost} &= \text{pond area} \times \text{liner cost} / \text{ft}^2 \\ &= 27,262,872 \times 0.73 \\ &= \$19,901,897 \end{aligned}$$

$$\begin{aligned} \text{Pond Volume (ft}^3\text{)} &= \text{disposal volume} \times 0.133 \times 30.5 \times \text{months of storage} \\ &= 2,340,000 \times 0.133 \times 30.5 \times 6 \\ &= 56,953,260 \text{ ft}^3 \end{aligned}$$

$$\begin{aligned} \text{Excavation Cost} &= (\text{pond volume} / 27) \times \text{excavation cost} \\ &= (56,953,260 / 27) \times \$3 / \text{CY} \\ &= \$6,328,140 \end{aligned}$$

$$\begin{aligned} \text{Fencing Cost} &= \text{pond circumference} \times \text{fence cost} / \text{ft} \\ &= (((27,262,872 \times 4) / \pi)^{0.5}) \times \pi \times \$10 / \text{LF} \\ &= \$185,093 \end{aligned}$$

## APPENDIX C

### RESPONSE TO REVIEWER COMMENTS

#### TWDB Comments on the Draft Report and Responses to Comments

##### TWDB Contract #0704830769

The following plain text is a record of comments on the draft version of this report, and the *italics text is a record of responses to those comments.*

##### Comments

1. Since the report is submitted to the TWDB, the WaterReuse Research Foundation's Disclaimer page should not be included in the report.

*Water Reuse Research Foundation's Disclaimer page was removed.*

2. Please add acronyms in the report.

*An acronym page has been inserted into the report.*

3. The term 'membrane pressure' is present throughout the report. Please replace the term 'membrane pressure' with the term 'transmembrane pressure'.

*Text was corrected.*

4. Please use 'Rotated Title' to align the 'Y-axis' of the figures.

*The 'Y-axis' of the figures were aligned as suggested.*

5. The second Draft report did not include responses of the following comments that were made for the first Draft report:

- a. Data Collection, Data Analysis, and Findings: Operating a single membrane element at 70% recovery appears to be atypical. Please include a range of recovery that is usually applied in a single element for the regular operation of a seawater reverse osmosis system.

*Please refer to the following:*

*A single membrane system operated at over 70% recovery was the usual operating method for all of the batch-treatment systems that we had previously used. All of those systems were based on recycled flow through a single membrane, with recoveries up to 90%. We started with a system having a single 2 ½” membrane and a permeate flow of 0.25 gpm. We then built a fully automated, larger system with four- 4” membranes (in parallel) that had a permeate flow rate 20 times larger than the small system, and it performed exactly like the small system.*

*If we were to build a full scale system, it too would be based on a single membrane system, with recycle, at a recovery of between 85% and 90% of the concentrate (from the KBH plant). Thus, our batch treatment system design is based on single membrane pressure vessels, with recycle, to achieve a high recovery. This project was based on our experience with those systems. Since we have not done any work with regular seawater RO systems, we are not in a position to comment about what recoveries are appropriate for a single membrane in a regular system.*

- b. Membrane Cleaning: Please mention the dose (in mg/L) of ‘High Flux A’ during cleaning.

*The Final report incorporated the dose of ‘High Flux A’ in terms of the dilution ratio, which is common practice in the water treatment industry.*

- c. Operating Modifications: The problem lies in the understanding of what is being simulated. If there were no recycle, the study would be simulating the element located next to the last element of the KBH plant. By recycling the concentrate, the study is simulating an element farther downstream. Please address the issue.

*The Final report mentioned that the problem could be fixed by allowing a single pass design with no RO concentrate recirculation.*

## **Specific Comments**

### **1. Chapter 2 Concentrate Characterization**

- a. Second sentence of the first paragraph: Please mention clearly that **average** concentrations of TDS, silica, calcium, and sulfates **for 2010** are 1273 mg/L, 145 mg/L, 722 mg/L, and 1410 mg/L, respectively.

*Text was added.*

- b. Second sentence of the first paragraph: To help understand the reason for increasing the antiscalant dosage from 4 ppm in the feed stream to 20 ppm in the concentrate stream, please mention the overall recovery of the KBH plant.

*Text was added.*

- c. Last sentence of the first paragraph: Please explain the consequence of the continuous increase of salt contents in the KBH concentrate over the past two years.

*An explanation was added to the Final report.*

- d. Table 2.1: Please spell out the following parameters

- ALK-T
- Fe-T
- Hard-T
- Temp

*The parameters were spelled out in the Table.*

## **2. Chapter 3 Experimental Set-Up**

- a. Figure 3.1:

- The figure is blurred. Please improve the quality of the figure.

*Figure quality was improved.*

- Please replace the word ‘Anti’ with the word ‘Antiscalant’ in the figure.

*Text was corrected.*

- b. Last sentence in Page 5: Please describe the terms, A, B, and C.

*A description was added.*

- c. Page 6, first sentence, second line: Please add a period after the word ‘motor’.

*Text was corrected.*

- d. Last sentence in Page 6: Please replace the word '6/24' with the words 'June 24, 2010'.

*Text was corrected.*

- e. Table 3.2: Please consider showing the Table as a component in Appendix A.

*Table was improved and left in the Chapter.*

### **3. Chapter 4 Equipment Problems Encountered**

- a. Please consider changing the Chapter Title from 'Equipment Problems Encountered' to 'Operational Problems'.

*The Chapter Title was changed.*

- b. Please consider dividing the chapter into the following sub-sections:

- Error in feed reading
- Error in water level reading
- Issues with positive displacement pumps
- Problems in process control and data logging system
- Problems with accessories
- Problems with high pressure relief valve
- Problems with feed pump

*The suggestion was accepted and the chapter was divided into the above mentioned sub-sections.*

### **4. Chapter 5 Data Collection, Data Analysis, and Findings (Pages 9-12)**

- a. Figures 5.1 and 5.2: In the Title of the figure, please clearly identify the parameter for 30,000  $\mu\text{S}/\text{cm}$ .

*The figure was modified as suggested.*

- b. Second sentence, last paragraph, Page 9: Please mention clearly that the system was pre-set in a way that it would be shut down automatically if the pressure reaches at 800 psi.

*The text was added.*

- c. Table 5.1:

- Please consider replacing the term ‘KBH conc’ with the term ‘SWRO Feed’.

*Text was modified.*

- Please spell out the words ‘Perm’ and ‘Conc’.

*Text was modified.*

- d. Table 5-2:

- Please spell out the words ‘Perm’, ‘Conc’, and ‘Perm+Conc’.

*The terms were spelled out in the report.*

- Please re-check the values for the first two rows of ‘Perm+Conc’.

*The values were corrected.*

- e. Figure 5.3:

- Please revise the Title of the figure.

*The Title of the figure was revised.*

- Please consider showing the numbers in ‘X-axis’ as a whole number.

*‘X-axis’ of the figure was modified.*

## **5. Chapter 6 Membrane Cleaning (Page 12)**

- a. According to Figure 6.1, five sets of experiments were carried out (on 26-Apr, 10-May, 17-May, 24-Jun, and 8-Jul). However, the text explained the results for only two sets of experiments (for 26-Apr, and 10-May). Please include explanations for other three sets of experiments in the text.

*Text was modified.*

- b. Figure 6.1: The figure contains too much information. Please revise the figure so that it can be interpreted easily.

*This figure was simplified; the data for early start-up was eliminated from each event.*

## **6. Chapter 7 Operating Modifications**

Please add a space between the second and the third paragraph.

*The edit was done.*

## **7. Chapter 10 References**

Please format the references.

*Reference was formatted.*