RECLAMATION

Managing Water in the West

Desalination and Water Purification Research and Development Program Report No. 149

Evaluation and Selection of Available Processes for a Zero-Liquid Discharge System for the Perris, California, Ground Water Basin

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14. ABSTRACT ***(Maximum 200 words)***
Eastern Municipal Water District faces a serious disposal issue for brine produced by their existing and planned primary ground water desalination facilities. The cost of brine disposal, especially for an inland agency, is very expensive and is expected to dramatically increase. As a result, a wide range of existing and emerging water treatment technologies were evaluated for the design of a zero-liquid discharge system. Alternative processes were tested; at pilot-scale, secondary reverse osmosis and electrodialysis reversal; at bench-scale, forward osmosis, membrane distillation, and seeded reverse osmosis; and by desktop modeling, brine concentrators, crystallizers, evaporation ponds, and SAL-PROC™. A cost analysis model based on individual treatment modules was developed for each process alternative with 14 process trains evaluated. Total annual costs ranged from $5,396,278 to $8,326,201. These figures represent a sum of the amortized capital annual costs plus operation and maintenance costs, and do not consider the revenue generated from recovered water.

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Desalination and Water Purification Research
and Development Program Report No. 149

Evaluation and Selection of Available Processes for a Zero-Liquid Discharge System for the Perris, California, Ground Water Basin

Prepared for Reclamation Under Agreement No. 05-FC-81-1153 Task F

by

Eastern Municipal Water District
Carollo Engineers

U.S. Department of the Interior
Bureau of Reclamation
Technical Service Center
Water and Environmental Services Division
Water Treatment Engineering Research Team
Denver, Colorado

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MISSION STATEMENTS

The mission of the Department of the Interior is to protect and provide access to our Nation's natural and cultural heritage and honor our trust responsibilities to Indian tribes and our commitments to island communities.

The mission of the Bureau of Reclamation is to manage, develop, and protect water and related resources in an environmentally and economically sound manner in the interest of the American public.

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The success of this research is a direct result of the cooperative effort between private, public, and Government agencies that have been generous in providing their time, expertise, and knowledge.

Sincere thanks are extended to the Desalination and Water Purification Research and Development Program, Bureau of Reclamation, for their vision and support in sponsoring this research.

Eastern Municipal Water District (EMWD) would like to thank their partner, Carollo Engineers, for operating the pilot-scale facilities, performing the secondary reverse osmosis, electrodialysis reversal, and seeded reverse osmosis processes, and developing the process combinations with associated costs for this research.

In addition, EMWD would like to thank the University of Nevada, Reno, for performing the forward osmosis and membrane distillation processes, Geo-Processors USA for performing the SAL-PROC processes, HPD-Veolia Water Solutions for performing the brine concentrator and crystallizer processes, Ionics for their assistance performing the electrodialysis reversal process, and the J.R. Filanc Construction Co. for constructing the pilot-scale facilities for this research.

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Acronyms

$BaSO_4$	barium sulfate
BC	brine concentrators
BC/XLZR	brine concentrator/crystallizer
Ca	calcium
$CaCO_3$	calcium carbonate
CaF_2	calcium fluoride
Carollo	Carollo Engineers
$CaSO_4$	calcium sulfate
Cl	chlorine
DC	direct current
DCMD	direct contact membrane distillation
DS	draw solution
DWR	Department of Water Resources
EDAX	energy dispersive x-ray
EDR	electrodialysis reversal
EDTA	ethylenediaminetetraacetic acid
EMWD	Eastern Municipal Water District
FO	forward osmosis
ft^2	square foot
Geo	Geo-Processors USA, Inc.
gpd	gallons per day
g/L	grams per liter
gpm	gallons per minute
gfd	gallons per square foot of membrane per day
H_2S	hydrogen sulfide
kWh	kilowatthour
L/min	liters per minute
m	meter
m^2	square meter
MD	membrane distillation
MF	microfiltration
Mg	magnesium

Acronyms (continued)

mgd	million gallons per day
$Mg(OH)_2$	magnesium hydroxide
mg/L	milligrams per liter
mmHg	conventional millimeter of mercury
Na	sodium
NaCl	sodium chloride
NF	nanofiltration
NTU	nephelometric turbidity unit
O&M	operation and maintenance
PCC	precipitated calcium carbonate
PLC	programmable logic controller
psi	pounds per square inch
psig	pounds per square inch gauge
PTFE	polytetrafluoroethylene
Reclamation	Bureau of Reclamation
RO	reverse osmosis
ROSA	reverse osmosis system analysis
SARI	Santa Ana Regional Interceptor
SDI	silt density index
SEM	scanning electron microscopy
Si	silicon
SiO_2	silicon dioxide
SPARRO	slurry precipitation and recycle reverse osmosis
TDS	total dissolved solid
VEDCMD	vacuum enhanced direct contact membrane distillation
VOC	volatile organic compound
XLZR	crystallizers
ZLD	zero-liquid discharge
°C	degrees Celsius
°F	degrees Fahrenheit
µg/L	micrograms per liter
µm	micrometer
µS/cm	microsiemens per centimeter

1. Executive Summary

1.1 Overview and Study Objectives

Eastern Municipal Water District (EMWD) faces a serious disposal issue for brine produced by their existing and planned primary ground water desalting facilities. The cost of brine disposal, currently via the Santa Ana Regional Interceptor (SARI), is expensive. As a result, Carollo Engineers (Carollo) was contracted by EMWD to research a wide range of existing and emerging water treatment technologies for the design of a zero-liquid discharge (ZLD) system under a Bureau of Reclamation (Reclamation) grant. In this report, alternative processes such as secondary reverse osmosis (RO), electrodialysis reversal (EDR), forward osmosis (FO), membrane distillation (MD), and seeded RO (slurry precipitation and recycle reverse osmosis [SPARRO]) were studied by either bench-scale experiments or pilot plant testing. Subsequent brine minimization techniques, including brine concentrators (BC), crystallizers (XLZR), evaporation ponds, and SAL-PROC™, were incorporated into the treatment processes; and overall economics were calculated through desktop modeling. A cost analysis model based on individual treatment modules was developed for each process alternative with a total of 14 process trains evaluated. This cost model will not only benefit EMWD but will also benefit other local, State, regional, and national agencies facing brine disposal issues from inland desalination facilities.

The least expensive alternative evaluated was primary RO+softening+EDR+BC+Evaporation Pond-Disposal (Option 6) at a total annual cost of $5,839,736. It is interesting to note that for the volume and quality of the brine considered in this study, there was only a 5- to 7-percent difference in total annual cost between all treatment trains that result in BC/Evaporation Pond-Disposal and BC/XLZR/Landfill options. The treatment costs for secondary RO and EDR are nearly equivalent at $5,868,570 and $5,839,736, respectively, for BC/Evaporation Pond-Disposal and at $5,954,690 and $5,945,360 for BC/XLZR/Landfill, respectively. For inland communities where access to the SARI line is not a viable option, brine minimization by brine concentrators and further crystallization prior to land filling are comparable to thermal evaporation ponds. Although the capital costs of BC/XLZR are more expensive than evaporation ponds, operation and maintenance (O&M) costs for the BC/XLZR alternatives are slightly cheaper ($200,000). In the long run, it could most likely be a “greener” solution than the upkeep of an evaporation pond—which can occupy as much as 12 acres of land for the proposed treatment alternatives.

The total costs calculated are the sum of the amortized capital annual costs plus O&M costs and did not deduct the possible revenue generated from recovered water through additional brine minimization. Excess SARI capacity was not sold in the cost calculations; and therefore, the total annual project costs could be less costly than projected.

Although emerging technologies such as MD and FO are evaluated in this study, further development remains before these processes can be commercially available at more competitive pricing in the future.

2. Background

2.1 General Introduction

Eastern Municipal Water District (EMWD) faces a serious disposal issue for brine produced by their existing and planned primary ground water desalting facilities. The brine (total dissolved solids [TDS] around 6,000 milligrams per liter [mg/L]) is presently discharged to the Santa Ana Regional Interceptor (SARI) pipeline that transports it about 50 miles to the Pacific Ocean.

The problem is twofold:

1. The cost of brine disposal via SARI line is expensive and will become more expensive in the future.

2. There is not enough capacity in the SARI line to handle EMWD's future brine production needs.

Total future brine flow is projected to be 6.6 million gallons per day (mgd), which includes brine from EMWD desalters (4.2 mgd), truck disposal (0.2 mgd), and two future power generation facilities (2.2 mgd).

As a result, EMWD has decided to further investigate recovering drinking water from the primary reverse osmosis (RO) brine stream and converting the entire system to zero-liquid discharge (ZLD). Today, the technology usually associated with ZLD systems is expensive and energy-intensive brine concentrators.

The purpose of this project is to evaluate five promising technologies that, individually or in combination, could act as an intermediate brine treatment step to further concentrate the existing brine and recover more potable water at a lower cost. This study involved desktop modeling and bench-scale testing to evaluate individual technologies, and combinations of technologies, from which the most appropriate treatment combination could be selected by EMWD for potential testing. The outcome of this project will not only benefit EMWD but also other local, State, regional, and national agencies facing brine disposal issues from inland desalination facilities.

2.2 Justification for Research

As with other inland communities, EMWD is faced with the challenge of assuring long-term availability of water to a rapidly growing population. One way in which EMWD has sought to address this need is to increase the number of brackish wells and treat ground water with RO for potable water supply.

However, as a finite resource, well water cannot be relied on as the sole means of addressing the water challenges of the Inland Empire. Even in the unlikely event of inexhaustible water resources, other treatment processes deserve exploration. This is, in part, due to conventional RO recovery limitations and concentrate disposal issues such as the cost and environmental well being of receiving bodies. These alternative means of water recovery are discussed in the following sections.

2.2.1 Recovery Limitations

Currently, EMWD operates two RO plants at a recovery rate of 70 percent. RO facilities have been known to operate at recoveries in excess of 85 percent using a typical two-stage membrane array. However, for EMWD's plants, recovery is limited by the presence of high concentrations of sparingly soluble salts in the brackish feed water. The water quality of the well water feed stream is presented in table 2.1. Key sparingly soluble salt constituents and foulants are highlighted (e.g., calcium, silica, magnesium [Mg], and sulfate).

Even with additing antiscalant chemicals and operating at 70 percent recovery, EMWD has seen significant scaling of their membranes, particularly in Train No. 2 of the Menifee Desalter. The lead and lag elements of the second stage membranes in this train have lost about 50 and nearly 100 percent of their permeability, respectively, in less than 2 years of operation.

2.2.2 Concentrate Disposal

Coupled with recovery limitations is the related issue of concentrate disposal. At the design recovery of 70 percent, a significant volume of water is discharged to the SARI line. Currently, this is about 2 mgd and will increase to 4.6 mgd when the third RO facility is constructed. Disposal to the SARI line represents a significant annual operating cost. Also, due to the concentrating effect on components such as heavy metals and arsenic (table 2.1), it is conceivable that EMWD may be required to pay increased costs for ocean discharge via the SARI line if stricter regulations are applied to discharge water quality in the future.

2.2.3 Economic Value

Historically, RO concentrate has been regarded as an unfortunate waste byproduct. Emerging technologies for recovery enhancement have begun to generate a paradigm shift in regards to this waste. Newer technologies provide the opportunity to recover more usable water and convert the salts to a revenue-generating product. As a result, the brine "waste" can now be viewed as a "resource." With respect to the RO brine at EMWD's facilities, preliminary

Table 2.1 Summary of Water Quality Data: Menifee Water Quality[1]

Parameters	Units	Primary RO Plant Feed Water
pH		6.1
Conductivity (at 25 °C)	µS/cm	3,476
TDS (at 180 °C)	mg/L	2,330
Hardness	mg/L	1,300
Alkalinity	mg/L as $CaCO_3$	537.5
Calcium	mg/L	331
Magnesium	mg/L	284
Sodium	mg/L	278.7
Potassium	mg/L	9.8
Bicarbonate	mg/L	662
Sulfate	mg/L	580
Chlorine	mg/L	565
Fluoride	mg/L	0.32
Aluminum	µg/L	78
Arsenic	µg/L	7.5
Barium	µg/L	129
Boron	mg/L	0.193
Iron	µg/L	387
Manganese	µg/L	115
Silica	mg/L	61.2
Strontium	µg/L	1,600
Zinc	µg/L	52
Nitrate as N	mg/L	3
Ammonia as N	mg/L	<0.5
Nitrite as N	mg/L	<0.01
Total Inorganic N (TIN)	mg/L	3.3
TKN	mg/L	0.7
Total Phosphate as P	mg/L	0.2

[1] °C = degrees Celcius; µS/cm = micorsiemens per centimeter; $CaCO_3$ = calcium carbonate; ug/L = micrograms per liter; N = nitrogen.

studies have shown that significant quantities of precipitated calcium carbonate and magnesium hydroxide could be harvested. These may find use in the paper and building industry. Generated revenue could be used to offset the costs of operating brine minimization facilities. What was formerly an operational liability has the potential to be transformed, if not fully, into a usable resource. In the long term, this approach is more sustainable when considering environmental impacts.

Although recovery limitations, concentrate disposal challenges, and economic concerns provide compelling arguments for exploring brine minimization technologies, concerns over their high associated costs have limited their testing and implementation. Most commercially available technologies for brine minimization are driven by thermal processes, which can result in significantly greater capital and operations and maintenance (O&M) costs than RO plants. In addition, while there are nonthermal processes that show promising results, for now, they remain unavailable for commercial use. These include technologies such as forward osmosis (FO) and membrane distillation (MD). However, for inland communities such as those served by EMWD, the cost argument may be less of a deterrent. Often, the costs to lay pipelines or pay for SARI access from these remote locations can be comparable to, or even greater than, the costs for these recovery enhancement technologies. Additionally, SARI disposal costs are expected to increase over the next few years; and EMWD's purchasable capacity will be outstripped by its population growth. These factors underscore the need to investigate recovery enhancement technologies for future consideration.

The results of this study will assist EMWD in meeting its current and future water treatment goals in a more environmentally and economically sustainable manner.

2.3 Project Goals and Objectives

The ultimate goal of this project is to research a wide range of existing and emerging water treatment technologies for the design of a ZLD system, which may be pilot tested and then ultimately constructed for the production of drinking water for inland communities within EMWD's service area. This research will identify the most feasible of technologies and then evaluate them in terms of process, applicability, economics, and robustness. It is desirable to develop a ZLD system that will be more cost effective than current brine discharge to the SARI line.

Our preliminary investigations have shown there are several candidate processes that have potential in addressing the brine discharge issues facing EMWD:

- Conventional lime/soda softening followed by secondary desalting treatment using RO or EDR.
- Seeded RO (or slurry precipitation and recycle reverse osmosis [SPARRO] process) for treatment of first and second stage RO brine.
- MD or FO for treatment of first stage RO brine.
- Recovering salts using the SAL-PROC™ process by Geo-Processors for beneficial use in chemical processing industry or water treatment plants.
- Brine concentrators and/or crystallizers to process desalting concentrate to form a salt that can be disposed of as a solid waste while recovering additional water.

The objectives for this study are:

1. Provide capital and operating cost estimates for each treatment technology.

2. Develop a mass balance model for each process combination.

3. Based on capital and operating cost estimates for each process combination, identify preferred alternatives for EMWD's ZLD system.

2.4 Description of Unit Processes for Brine Minimization

Historically, desalting technologies have assumed one of two forms: thermal desalting processes or membrane processes. The former includes brine concentration and crystallization, which are high-energy technologies that have traditionally not been feasible for the high-volume/low-cost product generated in the drinking water industry. The chief membrane desalting technologies have been restricted to RO and EDR. Their use in the water industry has been well established, but they produce large volumes of environmentally undesirable saline waste when operating at standard recoveries of 50 to 85 percent. This has led to renewed interest in lesser recognized, but equally valid membrane technologies (FO and MD), which are capable of operating at much higher recoveries and achieving ZLD opportunities. These and other brine minimization processes are discussed in detail below.

2.4.1 Chemical Softening and Secondary Desalting

This approach uses a combination of chemical and physical steps to enhance recovery from brine produced by a primary RO plant. Primary brine is treated with conventional softening chemicals (such as lime, sodium hydroxide, and soda ash) to precipitate hardness and other minerals. Recent pilot studies have confirmed that, at the appropriate pH, silica may adsorb to magnesium hydroxide precipitates and be removed through co-precipitation. Following settling, the supernatant is filtered to remove solids carried over from the precipitation step. Depending on operational conditions, softening pretreatment can remove up to 90 percent of some sparingly soluble salts (Carollo, 2006), often returning the hardness and silica concentration to that of the original feed to the primary RO plant.

After softening, water is then fed to a secondary desalting process such as EDR or RO. The TDS of the softened water is higher than that of primary RO feed and so the secondary membrane processes are forced to operate at a higher feed pressure, in the case of RO, or higher electrical potential in the case of EDR. Despite the increase in the feed water TDS concentration, higher recovery is sometimes possible in the secondary desalting step because the upstream softening may result in lower concentrations of scaling precursors than the primary feed.

Limitations of this treatment approach are the production and disposal of large volumes of solids from chemical softening, the need for high dosages of chemicals, and the presence of fine solids from the softening step that can impact downstream process performance.

2.4.2 Reverse Osmosis

High-pressure membrane processes such as RO are typically applied for the removal of dissolved constituents including both inorganic and organic compounds. RO is a process in which the mass-transfer of ions through membranes is diffusion controlled. Consequently, these processes can remove salts, hardness, synthetic organic compounds, disinfection-byproduct precursors, etc. However, dissolved gases such as hydrogen sulfide (H_2S) and carbon dioxide, monovalent ions such as chlorine (Cl) and sodium (Na), and some pesticides pass through RO membranes. Nanofiltration (NF), like RO, is a diffusion-controlled process; but NF membranes have a higher salt passage and lower rejection of monovalent ions (Na^+, Cl^-) than RO membranes. Figure 2.1 presents the process and instrumentation diagram of a high-pressure RO pilot plant.[1]

These technologies are usually expensive in terms of capital and operational costs for single contaminant removal but can be cost effective and provide substantial benefit when multiple contaminants are present in a water source. It is noted that the concentrate stream from NF/RO processes will contain high levels of rejected species and will require proper management and disposal.

NF/RO processes that use spiral-wound membranes are easily compromised by the presence of particulate matter. Particulate matter can become entrained within the interstitial spaces of membrane channels and result in colloidal fouling. This material may be removed through membrane cleaning, but this cleaning is done at the expense of operation time. To prevent colloidal fouling, feed water is required to have a turbidity value below 0.5 nephelometric turbidity unit (NTU). At this low value, a more precise measure of the suspended solids content of RO feed water is the unit-less silt density index (SDI). An SDI of less than 3 is considered optimum for RO operation.

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[1] Source of all figures is Corollo Engineers and Eastern Municipal Water District, unless otherwise noted.

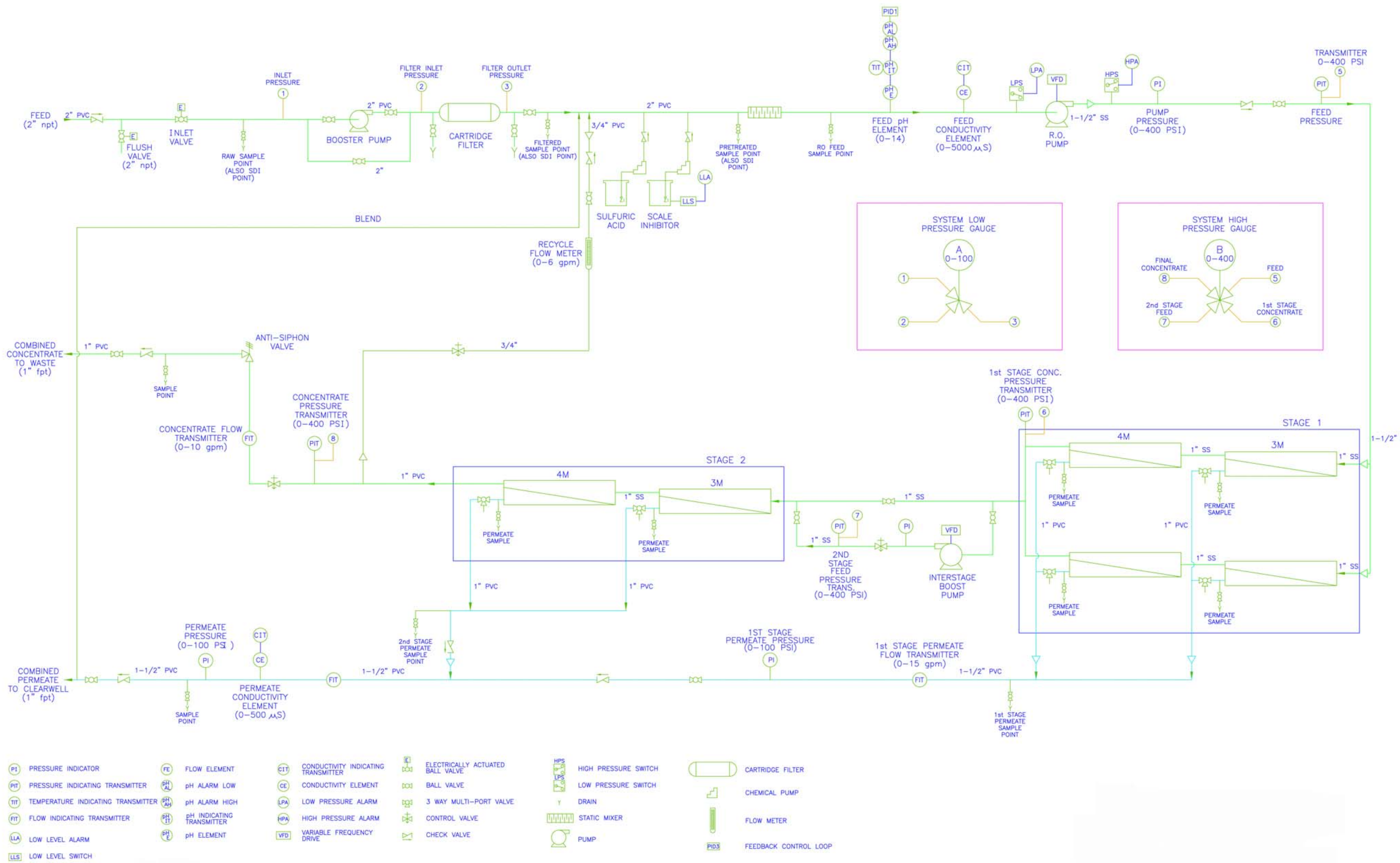


Figure 2.1 High-pressure membrane schematic.

Major limitations of RO processes are the limited recoveries (50-85 percent) dictated by the presence of large concentrations of sparingly soluble salts in the feed water and concentrate disposal challenges (both with respect to volume and components of the waste).

2.4.3 Electrodialysis Reversal

Electrodialysis reversal (EDR) is an electrochemical separation process that allows selective passage of ions, or charged species, in solutions. Only anions, or negatively charged ions, can pass through an anion exchange membrane, while cation exchange membranes transport positively charged ions, or cations. Ions are transferred through ion exchange membranes by means of direct current (DC) voltage and are removed from the feed water as the current drives the ions through the membranes to desalinate the process stream.

EDR units utilize membrane stacks with electrical stages. Each electrical stage also has two corresponding hydraulic stages. Water passes through each electrical stage twice to provide greater residence time for ion transfer. Essentially, an electrical stage is composed of one cathode and one anode separated by a series of cationic and anionic membranes and spacers. Electrodes are comprised of platinized titanium with a rare earth paint layer. Both cation and anion-transfer membranes are acrylic backed. Anionic membranes deflect positively charged cations and attract negatively charged anions, while the cationic membrane allows the passage of cations and rejects anions. Membranes are separated by spacers to separate both brine and product water streams.

EDR systems reduce the fouling tendencies of the water by reversing the polarity of the electrodes every 15 to 20 minutes. This change in polarity causes the scale to disassociate from the membranes. Figure 2.2 depicts the overall schematic of the EDR process.

Along with driving forces for demineralization, an important distinction between EDR and RO processes is that EDR membranes are not scaled by silica. As an uncharged molecule (less than pH=9), silica flows past the membranes with the permeate water and is, therefore, not concentrated in the brine stream. Consequently, EDR can be more cost effective for treating water where silica levels limit recovery. The EDR process is more tolerant of suspended solids than spiral-wound membrane RO systems and require feed water with a turbidity less than 2 NTU. As with RO, the concentrate stream of EDR processes contains ions well in excess of their solubility product. As a result, further recovery by membrane processes is not economically feasible without some post-treatment of this brine.

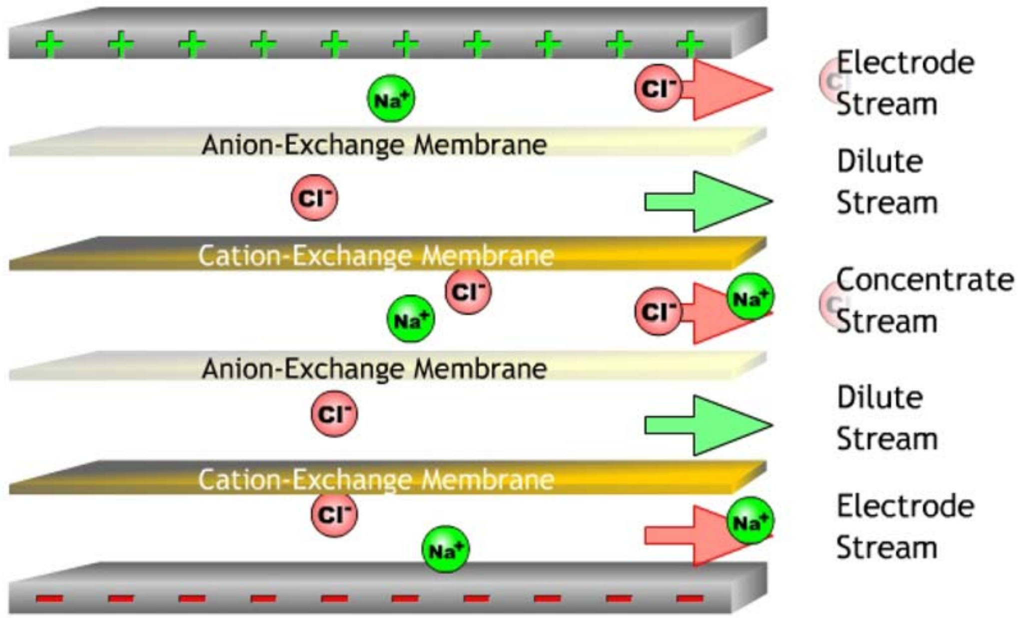


Figure 2.2 Schematic of electrodialysis process.

2.4.4 SPARRO (Seeded RO)

Seeded RO is a hybrid of conventional RO technology. It incorporates the recirculation of seeded slurry through the RO system, promoting homogeneous nucleation and precipitation from a solution. This process was developed to concentrate mine process water high in calcium and sulfate ions (Ca^{2+} and SO_4^{2-}) and was termed the SPARRO process. Seed crystals (typically gypsum) are introduced to the feed stream, which is then pumped into tubular RO membranes. As the water is concentrated along the membranes, the solubility products of calcium sulfate ($CaSO_4$), silicates, and other scaling salts are exceeded; and they preferentially precipitate on the seed material rather than on the membranes. A schematic of the seeded RO process concept is shown in figure 2.3. In water treatment, food-grade gypsum (anhydrous calcium disulfate) is used as the initial source of seed crystals to precipitate $CaSO_4$. The growing seed is recirculated through the system and removed in a controlled manner to maintain the desired concentration.

Concentrate containing seed crystals is processed in a cyclone separator to separate the seeds, and the desired seed concentration is maintained in a reactor tank by controlling the rate of wasting the upflow and/underflow streams from the separator. The technology has been tested at pilot scale for treating cooling tower blowdown (O'Neail et al., 1981) and highly scaling mine water (Juby, South Africa, 1996) and more recently using primary and secondary brine from the EMWD ZLD pilot project (September to November, 2006).

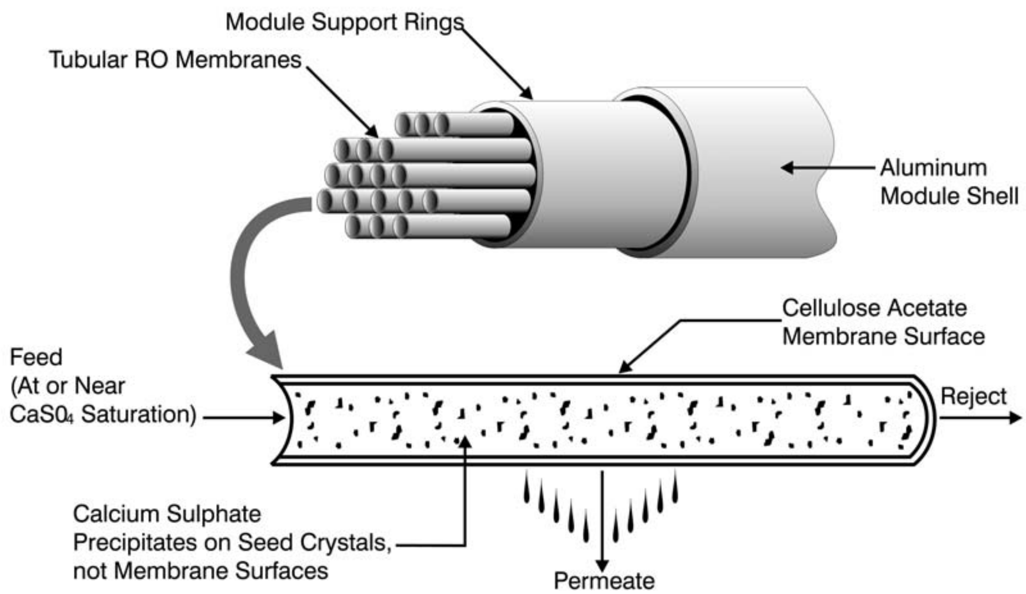


Figure 2.3 Schematic of seeded RO process concept.

2.4.5 Membrane Distillation

MD is a low-temperature separation technology that takes place through the pores of a hydrophobic microporous membrane. The driving force for separation is a vapor pressure gradient, which is generated by facilitating a temperature differential across the membrane. The volatile components of a heated-feed solution evaporate and pass through the pores to condense in a cold distillate stream on the permeate side. Typically, the process is used to separate volatile solutes such as volatile organic compounds (VOCs) from aqueous solutions. However, for aqueous feeds with nonvolatile solutes, only the volatile solvent (water) passes through the membranes; and the distillate is comprised of demineralized water. Fundamental criteria for MD are that the membrane must not be wetted and only vapor and noncondensable gases can be present within its pores. MD has demonstrated excellent ability to retain nonvolatile solutes and generate a nearly pure demineralized stream. Figure 2.4 presents an overview of the MD process.

MD technology has been around for 40 years, has been the subject of numerous academic studies, but has yet to see commercial use. Primary limitations to commercial application are lower flux compared to more conventional membrane separation technologies and the lack of membranes optimized for MD processes.

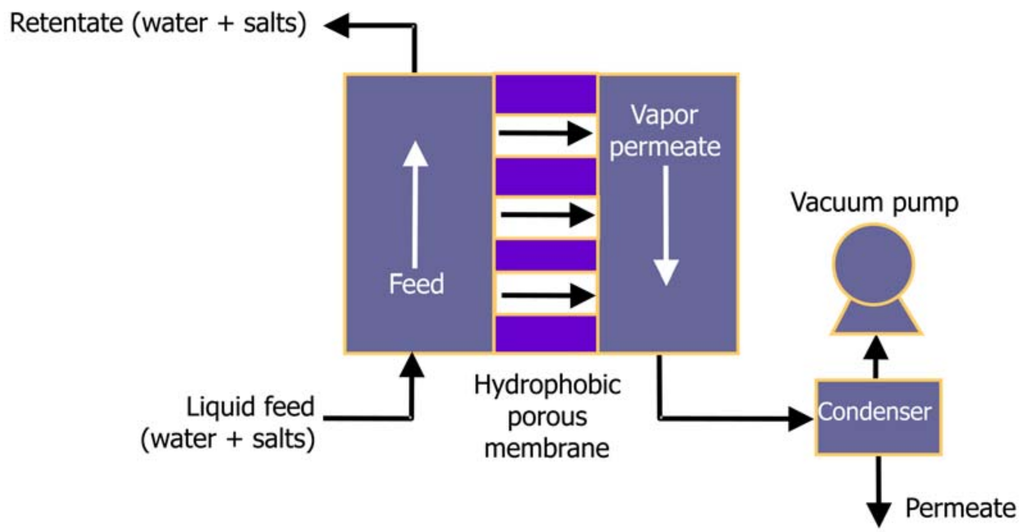


Figure 2.4 Schematic of membrane distillation process.

The recent resurgence of interest in MD may be attributed largely to the opportunity it presents to achieve ZLD. Two modifications in MD technology, direct contact membrane distillation (DCMD) and vacuum enhanced direct contact membrane distillation (VEDCMD), have increased its efficiency in demineralizing a concentrate stream. In DCMD, a hydrophobic membrane separates the hot feed from the cold distillate; and the volatile component of the feed evaporates through the membrane pores. In VEDCMD, a vacuum is placed on the permeate side to lower absolute permeate pressure. Since transfer occurs down the vapor pressure gradient, from high to low, the vapor pressure gradient is increased resulting in higher water (vapor) flux.

It is important to note that unlike RO and FO, MD processes are not selective for any particular ions; membranes are only a physical support for the vapor-liquid interface.

2.4.6 Forward Osmosis

Osmosis is the movement of water through a selectively permeable membrane from a region of low solute (high water) concentration to one of higher solute (low water) concentration. Two streams (a concentrate and dilute stream) are generated because the membrane rejects ions and most solute molecules but allows the passage of water. The driving force for mass transport is the osmotic potential gradient, which dictates that water moves from low to high osmotic potential. In reverse osmosis, hydraulic pressure is used to oppose and exceed the osmotic pressure of the "concentrated" solution so that water moves in the reverse direction against the osmotic potential gradient (from high to low solute concentration). FO is another separation technology that is based on the osmotic pressure of a feed solution. However, unlike RO, FO generates pure water by enhancing, rather than impeding, the osmotic pressure gradient between two aqueous solutions. Separation is driven by the osmotic pressure differential between a feed and a draw solution, and water moves according to its natural tendency from low to high osmotic potential. Figure 2.5 presents an overview of FO technology.

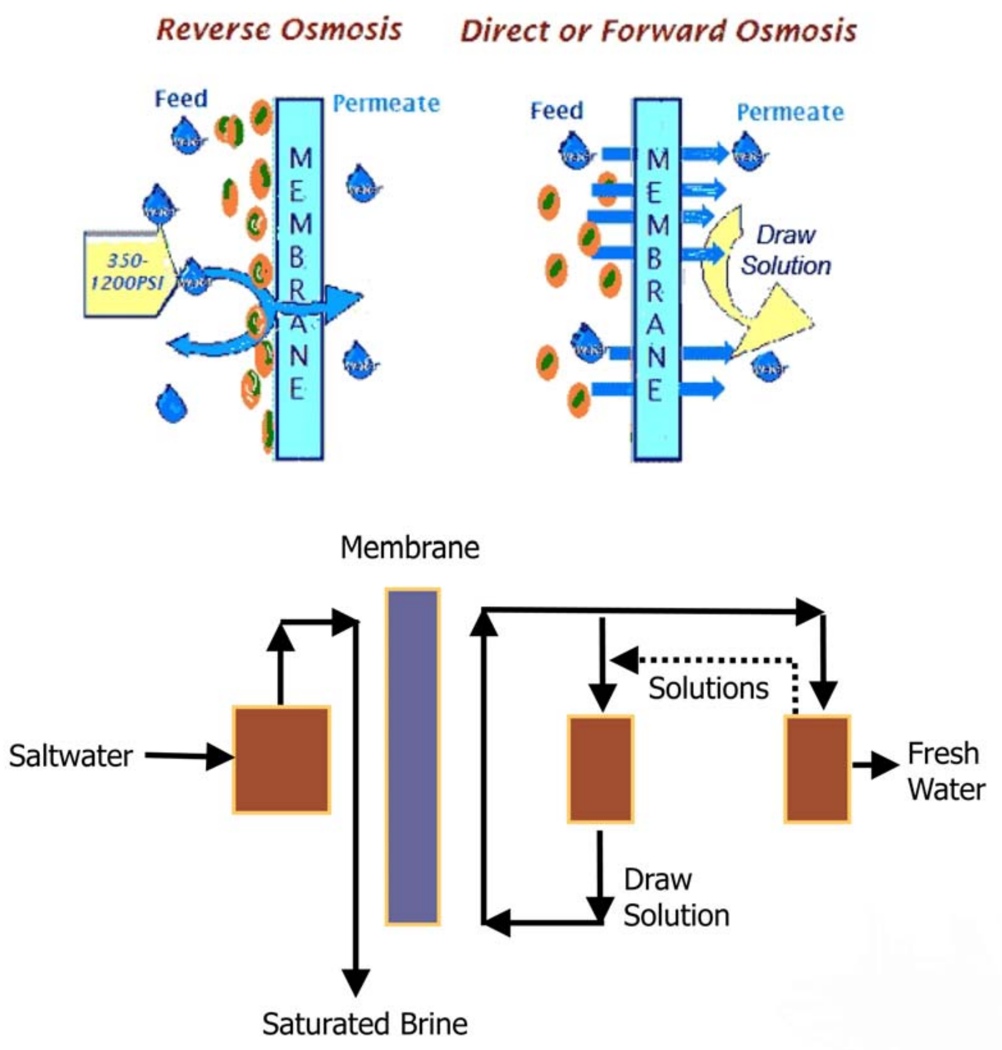


Figure 2.5 Schematic of FO process.

The draw solution is placed on the permeate side of the membrane and is diluted as water diffuses from the feed side into the permeate stream. The main criteria in the choice of a draw solution are as follows:

1. It must have a higher osmotic pressure than the feed.
2. It must have low or no toxicity.

3. It must be chemically nonreactive with polymeric membranes.
4. It must easily separate the water from the solvent.

Sodium chloride (NaCl) is often used in studies because of its high solubility and the relative ease with which it may be separated from the product fresh water, enabling the draw solution to be re-concentrated and reused. FO presents a number of advantages over RO:

1. Low to no hydraulic pressure required.
2. Higher rejections for a wide range of contaminants.
3. FO has been shown to have a lower propensity for membrane fouling.

Like MD, the study of FO has been restricted to the bench-scale level where flat sheet membranes and, to a lesser extent, tubular membranes have been used to demonstrate the effectiveness of FO. These membranes lend themselves to the FO application because they allow liquids to flow freely on both sides, which is a necessary flow characteristic for FO. A review of the literature shows that using FO technology in large-scale applications suffers from two major limitations:

1. The lack of available robust optimized membranes.
2. The slow development curve in the production of draw solutions that can induce separation and yet be amenable to low-energy separation from the product fresh water.

With increased attention on addressing these limitations, FO will emerge as a viable alternative in the suite of membrane technologies used in the drinking water industry.

2.4.7 Residual Recovery (SAL-PROC)

SAL-PROC is one of the newer entries into the drinking water industry. It is a proprietary technology combination developed by Geo-Processors USA, Inc. (Glendale, California) and includes several processing steps. The process enables the selective and sequential extraction of dissolved constituents from a saline feed in the form of valuable chemical byproducts in crystalline, slurry, and liquid forms. The process uses multiple evaporation and/or cooling steps, supplemented by conventional mineral and chemical processing steps. The technology is based on simple closed-loop processing and fluid-flow circuits, which enable the comprehensive utilization of inorganic saline streams to recover valuable mineral and chemical products. Recent large-scale pilot systems have demonstrated the technical feasibility of the process to produce a number of valuable chemicals from one or more waste streams while achieving ZLD. The chemicals typically harvested from saline streams are gypsum-magnesium hydroxide, magnesium hydroxide, sodium chloride, calcium carbonate, sodium sulfate, and calcium chloride. These can be sold to a number of industries, generating an income stream from what was formerly a waste product. In this way, the waste is transformed into a resource, and the revenue generated may be used to offset operational costs of the facility.

2.4.8 Brine Concentrators

Brine concentrators (BC) are mechanical evaporators that are commonly used in the power industry to further concentrate cooling tower blowdown before final disposal. Most concentrators work on single-effect evaporators, which use steam to heat brine solutions and promote water evaporation or operate on an electrically powered vapor compressor. Heat released from condensing steam is transferred to the brine solution via a heat exchanger, which boils the brine solution. Brine concentrators may use multiple stages to increase the overall efficiency, and economy, of the treatment process. Advantages of using brine concentrators include producing high-purity distilled water that has monetary value, as in the case of drinking water production. Figure 2.6 presents an overview of a brine concentrator. In addition, water evaporators are not dependent upon climatic conditions. Furthermore, evaporators are very effective at reducing brine solutions to very concentrated levels; TDS levels may be as high as 250,000 mg/L at a recovery of 90 to 98 percent.

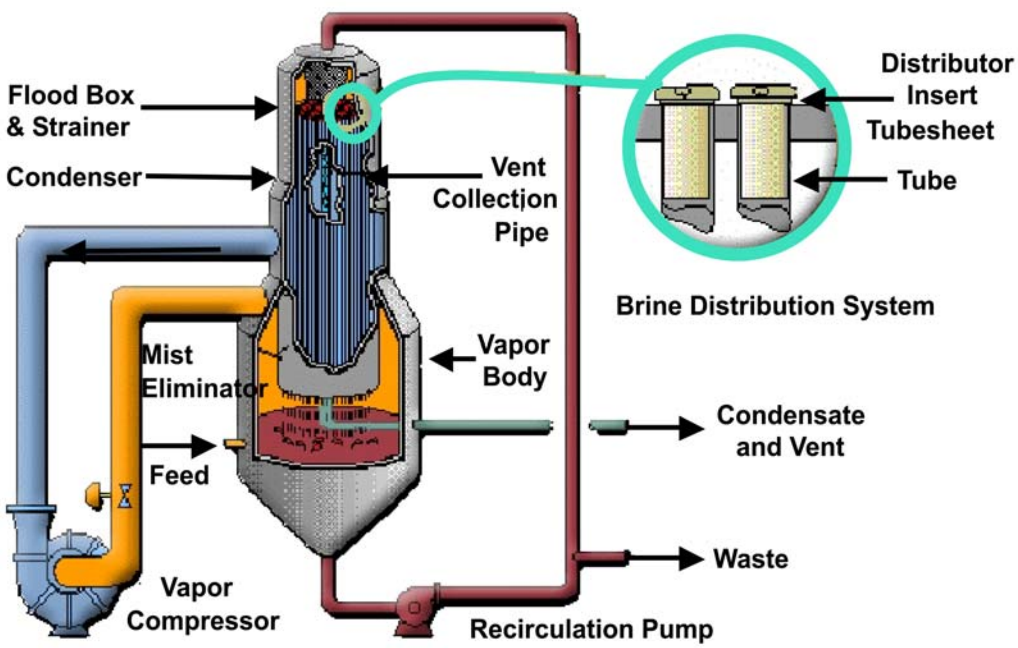


Figure 2.6 Schematic of brine concentrator (graphic courtesy of GE/Ionics).

Since such concentrated brines may be produced, the reject brine tends to be very corrosive and requires that evaporators be constructed of very durable and high-quality materials such as high-grade stainless steel and titanium. These costly materials drive up the capital cost of concentrators. Capacities of commercially available brine concentrators range from 10 to 700 gallons per minute (gpm) (1 mgd) with estimated energy consumption of approximately 90 kilowatthours per 1,000 gallons (kWh/1,000 gal). Resulting brine streams may be discharged to an evaporation pond, the SARI line, or trucked off site.

2.4.9 Crystallizers

Crystallizer (XLZR) technology has been used for many years to concentrate feed streams in industrial processes. More recently, as the need to concentrate wastewater has increased, this technology has been applied to reject from desalination processes, such as brine concentrate evaporators, to reduce wastewater to a transportable solid. Crystallizer technology is especially applicable in areas where solar evaporation pond (see section 2.4.10) construction cost is high; solar evaporation rates are low; and where deep-well injection is costly, geologically not feasible, or not permitted. The crystallizer converts the remaining waste to water that is clean enough for reuse in the plant and solids that are suitable for landfill disposal. Figure 2.7 presents an overview of a crystallizer.

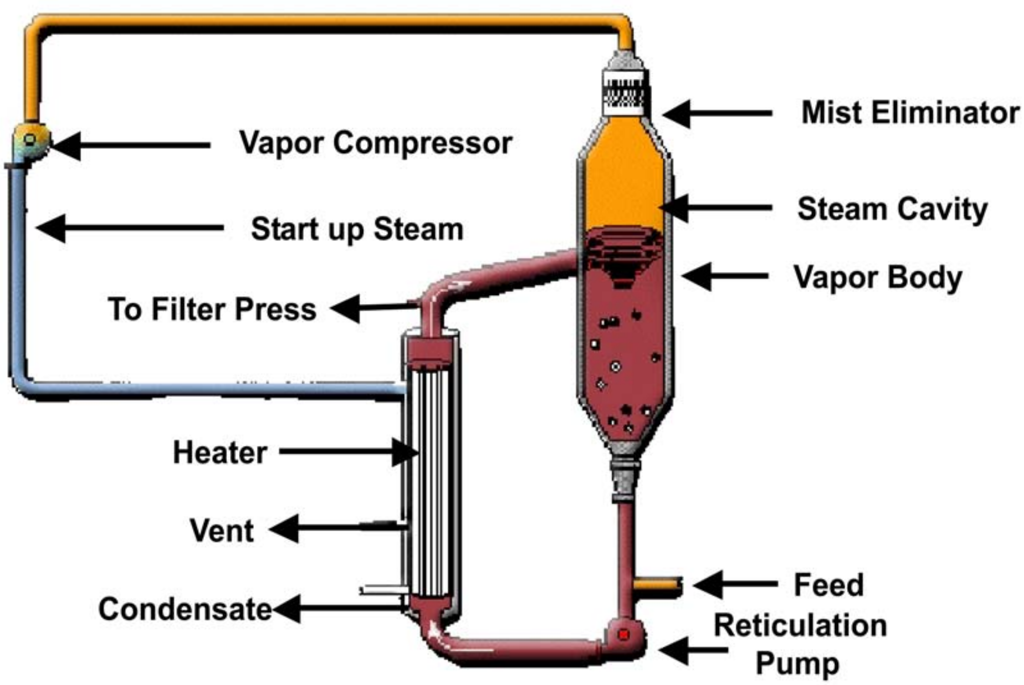


Figure 2.7 Schematic of brine crystallizer (graphic courtesy of GE/Ionics).

Crystallizers used for brine disposal range in capacity from about 2 to 50 gpm. These units have vertical cylindrical vessels with heat input from vapor compressors or an available steam supply. For small systems in the range of 2 to 6 gpm, steam-driven crystallizers are more economical. For larger systems, electrically driven vapor compressors are normally used to supply heat for evaporation.

For RO concentrate disposal, crystallizers would normally be operated in conjunction with a brine concentrator evaporator to reduce brine concentrator blowdown to a transportable solid resulting in a ZLD system. Crystallizers can be used to concentrate RO reject directly, but their capital cost and energy usage is much higher than for a brine concentrator of equivalent capacity (Mickley, 2001).

2.4.10 Evaporation Ponds

Evaporation ponds can be the final disposal point in a desalination treatment train once the brine stream has been concentrated to a manageable volume. Evaporation ponds use solar energy to heat and evaporate water from brine solutions, depositing salts in ever-greater concentrations on the pond floor. This form of treatment is advantageous in the Southwestern part of the United States where evaporation rates are high (50 to 100 inches per year) and in those areas where land is relatively inexpensive. Evaporation ponds are relatively easy to construct, easy to maintain, and have low operational costs (mainly the pumping of brine solution to ponds). However, evaporation ponds eventually have to be capped or have the solids removed for land filling once storage capacity has been exhausted.

Evaporation ponds are already used extensively in salt production facilities. However, when used in drinking water production, evaporation pond sizes may become relatively large, even in arid climates, if the volume of brine to be disposed is high. A large evaporation land area not only directly increases land purchase costs, but it can greatly increase construction costs due to the cost of pond liners, usually required for permitting. Monitoring wells are also likely to be required for meeting the permit requirements. Therefore, evaporation pond costs become excessive when the waste flow rate exceeds about 0.2 mgd (Mickley, 2001).

The evaporation rate for saline waters is lower than that for fresh water. Therefore, a newly created evaporation pond will experience a greater rate of evaporation than an aged system. However, this should not discourage the use of pond evaporation since that even at concentrations of greater than 250,000 mg/L TDS, brine solutions will evaporate at approximately 80 percent of that of fresh water. In addition, local wind velocities tend to increase evaporation rates in ponds.

Past work in the construction of evaporation ponds has shown that water depths should be about 1 to 18 inches for optimal evaporation of water. However, other factors must be considered in the construction of ponds including the provision of adequate storage for water surges, rain water, salt buildup, and adequate freeboard for wave action—the latter resulting from windy conditions at the pond site.

Cost drivers for evaporation pond systems are the local evaporation rate and climate, the concentrate volume, land and earthwork costs, liner costs, and the salinity of the concentrate, which determines the useful life of the ponds. The main cost variable is the evaporative area, and the largest individual cost is frequently the liner cost, where double layers are required.

2.5 Subcontractors

Carollo performed the technical investigations for this project including seeded RO, secondary softening RO, EDR, and overall analysis. To assist with the project, Carollo employed the following subcontractors:

1. Geo-Processors completed a desktop study of the application of its technologies on two different water qualities with the objective of providing a preliminary benefit/cost analysis for a comparative evaluation with other brine treatment alternatives.

2. HPD-Veolia Water Solutions completed an initial cost estimate of their concentrator and crystallizer system used to treat two different brine streams. Veolia provided information on operation and maintenance costs, energy requirements, and capital costs.

3. University of Nevada, Reno, performed bench-scale studies on MD and FO. The objectives of these studies were to confirm the technical viability and the operating and capital costs of the MD and FO processes.

3. Technical Approach

3.1 Project Implementation Plan

There are several processes in various stages of development that could potentially treat brine from a primary desalting process to produce a zero-liquid discharge system. These processes have been identified and were discussed in section 2.4. A six-step approach was used to achieve the goals on this project.

Step 1 – Identify the potential processes

Step 2 – Develop operating criteria

Step 3 – Develop operating cost data

Step 4 – Develop capital cost estimates

Step 5 – Develop process combinations

Step 6 – Evaluate process combinations and select preferred alternative

3.1.1 Introduction

The overall approach for this project was one of desktop modeling evaluations coupled with some bench-scale tests and pilot-scale tests to verify global process criteria and performance on different concentrations of RO brine.

The brine stream (Brine A) produced from EMWD's primary RO process had a TDS of approximately 6,000 mg/L and underwent softening to be used as feed water to a secondary RO pilot and/or EDR unit. The brine stream produced from the secondary RO pilot was designated "Brine B" and had a TDS concentration of approximately 18,000 mg/L. The brine concentrate from EDR was similar in composition to "Brine B" and was assumed to be the same for subsequent desktop modeling exercises. "Brine C" was the concentrate produced from the SPARRO treatment using "Brine B" as the feed water and resulted in a TDS concentration of approximately 22,000 mg/L.

As discussed in section 2.3, there are several other candidate processes to treat the primary RO brine produced from EMWD's desalter, such as forward osmosis and membrane distillation. Samples of "Brine A" and "Brine B" were used as feed water to FO and MD bench-scale testing and tested for maximum allowable recovery and membrane performance. The compositions for both "Brine A" and

"Brine B" were also sent to Geo-Processors USA to be modeled as feed to their SAL-PROC technology to obtain a recommended process stream for salt and water recovery.

Concentrate discharges from the pilot studies (RO, EDR, and SPARRO) and bench-scale experiments (FO and MD) could be processed downstream for further water recovery to achieve zero-liquid discharge. These downstream processes included a combination of brine concentrator and/or crystallizers, evaporation ponds, and direct discharge to the SARI line. Brine composition and volume information were sent to manufacturers to obtain accurate price quotes specific to this project.

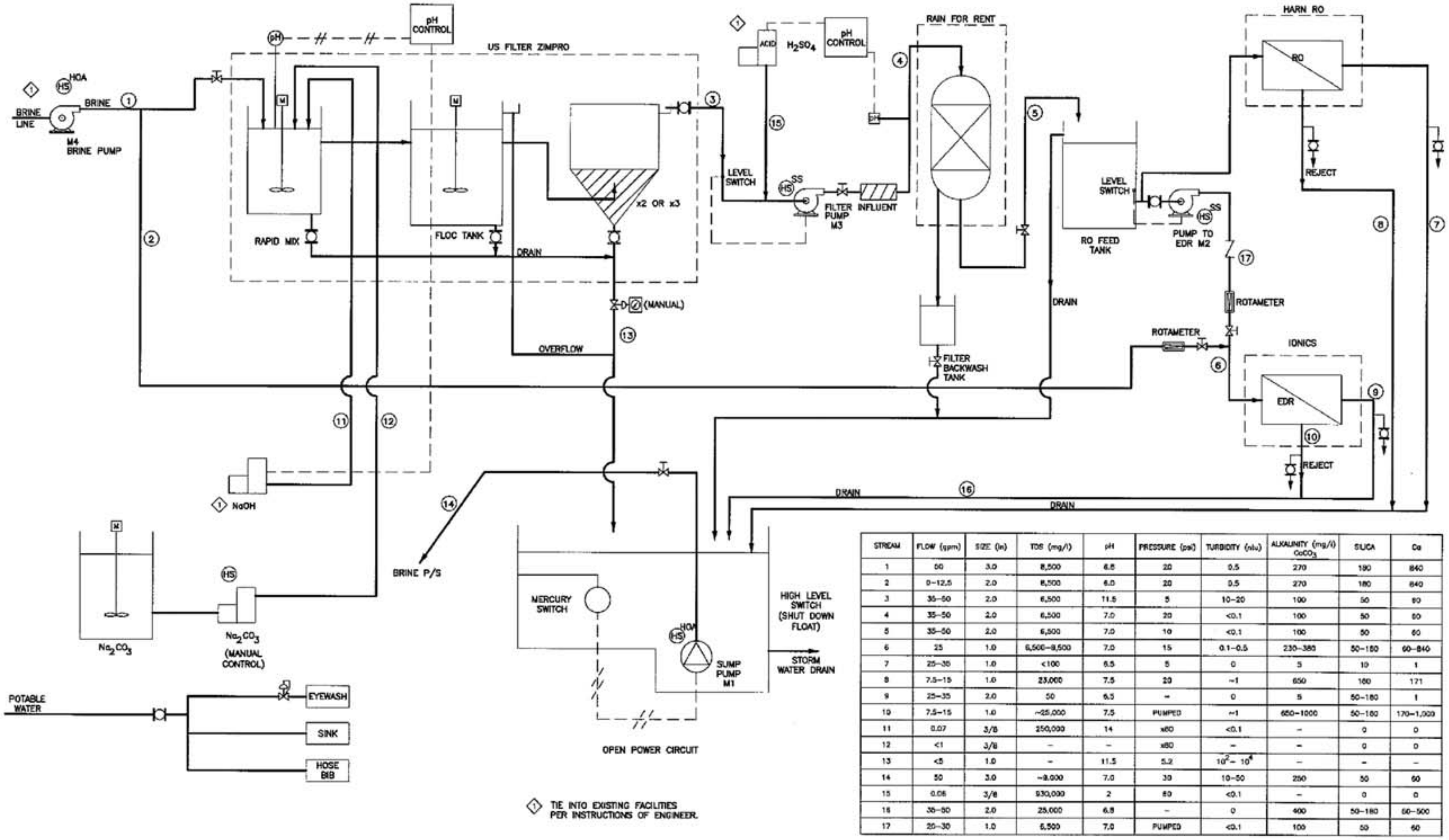
Each brine treatment process is treated as a block with associated capital and operating cost information in our desktop model. Different process blocks were combined to form alternate treatment trains in the model, and the resulting costs were compared for the most viable brine treatment train. The nature of the experiments conducted and the assumptions for our desktop cost model will be presented in detail in subsequent sections.

3.2 Pilot Studies

3.2.1 Chemical Softening and Secondary Desalting

Figure 3.1 presents the overall process flow diagram of the pilot plant that was part of the California Department of Water Resources Prop 50 "Desalination Recovery Enhancement and Concentrate Management" study conducted at EMWD. Menifee brine (Brine A) was pumped at 20-50 gpm through buried lines from the Menifee desalter building to the California Department of Water Resources (DWR) Prop 50 pilot plant slab. A preliminary softening and clarification step using a USFilter ZIMPRO unit was followed by conventional dual-media filtration to remove suspended solids. The ZIMPRO unit consisted of rapid mixing and flocculation tanks followed by a lamellar flow inclined plate settling basin mounted above a sludge hopper.

The brine stream first entered the rapid mixing chamber where it was dosed at between 600 and 650 mg/L with 50-percent caustic soda to achieve an elevated pH between 9.5 and 10.0. Anionic polymer at 0.5 to 1.0 mg/L was added to the flocculation tank to encourage agglomeration of flocs. As water from the flocculation tank flowed through the inclined plate settler, insoluble salts of calcium carbonate and magnesium hydroxide formed and settled in the clarification section. During the testing phase, additional downstream solids removal steps were added (quiescent settling and filtration) to remove solids that did not settle in the clarifier.



STREAM	FLOW (gpm)	SIZE (in)	TDS (mg/l)	pH	PRESSURE (psi)	TURBIDITY (ntu)	ALKALINITY (mg/l) $CaCO_3$	SILICA	Ca
1	50	3.0	8,500	6.8	20	0.5	270	180	840
2	0–12.5	2.0	8,500	6.0	20	0.5	270	180	840
3	35–50	2.0	6,500	11.5	5	10–20	100	50	60
4	35–50	2.0	6,500	7.0	20	<0.1	100	50	60
5	35–50	2.0	6,500	7.0	10	<0.1	100	50	60
6	25	1.0	6,500–8,500	7.0	15	0.1–0.5	230–380	50–180	60–840
7	25–35	1.0	<100	6.5	5	0	5	10	1
8	7.5–15	1.0	23,000	7.5	20	~1	650	180	171
9	25–35	2.0	50	6.5	–	0	5	50–180	1
10	7.5–15	1.0	~25,000	7.5	PUMPED	~1	650–1000	50–180	170–1,000
11	0.07	3/8	250,000	14	≈80	<0.1	–	0	0
12	<1	3/8	–	–	≈80	–	–	0	0
13	<5	1.0	–	11.5	5.2	$10^2 - 10^4$	–	–	–
14	50	3.0	~8,000	7.0	30	10–50	250	50	60
15	0.08	3/8	930,000	2	80	<0.1	–	0	0
16	35–50	2.0	25,000	6.8	–	0	400	50–180	60–500
17	20–30	1.0	6,500	7.0	PUMPED	<0.1	100	50	60

Figure 3.1 Process flow diagram for softening secondary desalting.

3.2.2 Reverse Osmosis

Softening effluent required the dosing of 25-percent sulfuric acid to adjust the pH to a more neutral value (pH 7.0) before treatment with RO and EDR units. The RO skid-mounted pilot plant was purchased from Harn RO Systems and consisted of two stages of RO membranes, organized in a 2:1 array of three- and four-element pressure vessels operated in series. The particular array was selected to mimic the configuration of primary desalter.

RO membranes were selected on the basis of chemical compatibility with the membranes in use at EMWD's Menifee and Perris desalters and Carollo's RO pilot-skid pressure vessels. As a result, the pilot membranes chosen were Dow Filmtec™ XLE-4040 spiral wound 40-inch (length) by 4-inch (diameter) membrane elements. Vitec 3000 from Avista was selected as the antiscalant for this study and added to the feed at a dose of 4.0 mg/L.

3.2.3 Electrodialysis Reversal

For the pilot project, EDR was tested using the mobile Ionics piloting platform, Aquamite V. This unit is in a trailer cargo container with dimensions 20 feet long by 8 feet wide by 12 feet tall. The unit used a single EDR membrane stack with two electrical stages. Each electrical stage also has two corresponding hydraulic stages. Water passes through each electrical stage twice to provide greater residence time for ion transfer. Water developed within the concentrate cell pairs is recirculated back to the concentrate system in a concentrate loop. A small booster pump is used to circulate the concentrate loop until the salts become supersaturated and a portion of the loop must be removed, which creates a reject stream. This process of brine removal is referred to as brine "blowdown," and the dilution and replenishment of the brine loop is referred to as "brine makeup."

EDR cathode and anode operation alternated every 15 to 30 minutes by reversing the polarity or direction of the current flow. This aided in preserving the integrity of the membranes by preventing scale buildup. During charge reversal, approximately 30 to 45 seconds, water was not to specification and was diverted as a waste stream.

Other specific attributes of the Ionics pilot EDR used for testing included:

1. A production rate of 18,000 to 20,000 gallons per day (gpd) with a water recovery of 75 to 80 percent.

2. Voltage regulator for simulating a two-stage, three-stage, or four-stage EDR unit.

3. Feed water pump, EDR membrane stack, and chemical feed systems.

4. Dedicated programmable logic controller (PLC) and data logger for EDR pilot control and operation. The EDR pilot will operate independently of the overall pilot process control panel.

Figure 3.2 is a photograph of the pilot plant equipment.

Figure 3.2 Site layout of RO skid, Zimpro softening unit, and EDR trailer.

3.2.4 SPARRO (Seeded RO)

In seeded RO experiments, the need to circulate gypsum slurry within membranes confines using membrane configurations to those that will not plug, namely, tubular membrane systems. There are a limited number of tubular membrane manufacturers; and for this work, tubular NF membranes were supplied by Koch Membrane Systems (San Diego, California).

Koch Membrane Systems does not currently manufacture a tubular RO membrane; hence, an NF membrane was used. Each 150-inch membrane module was a self-contained pressure vessel. The NF membrane module had an active membrane area of 28 square feet (ft^2) (2.6 square meters [m^2]) with typical operating pressures between 220 to 510 pounds per square inch (psi). To produce a typical permeate flow of about 1,000 gpd, 18 tube lengths were connected end to end within each module, with one inlet and one outlet per module.

Figure 3.3 presents an overview of the batch-testing arrangement used for the seeded RO process testing. Brines A and B were tested in the seeded RO system. Preliminary experiments conducted with unsoftened Brine A produced inconsistent results; and hence, Brine B was tested extensively. First, the batch-feed tank was filled with 120 gallons of the brine to be tested. Commercial gypsum powder ($CaSO_4.2H_2O$) was then added to the brine to produce the desired concentration of gypsum slurry (10-18 grams per liter [g/L]). The brine and gypsum slurry was then pumped though the Koch tubular membranes using a high-pressure positive displacement pump. The feed entered the membranes at a pressure of between 200 and 600 psi. Permeate produced from the membrane vessel was removed from the process and sampled periodically for laboratory analysis.

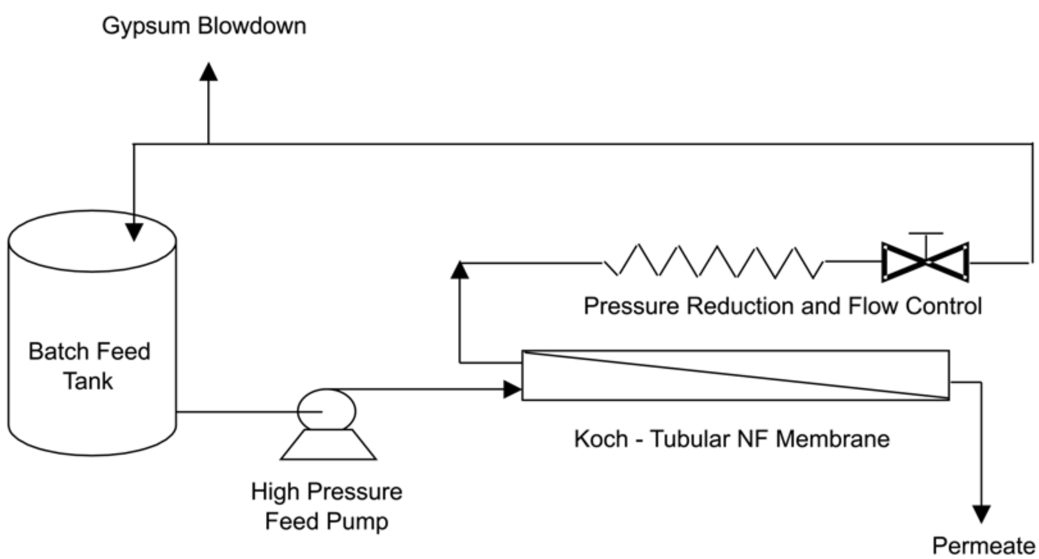


Figure 3.3 Seeded RO process testing setup.

The concentrate stream leaving the membrane vessel was piped through a pressure reducing system and then through a flow control valve. Low-pressure concentrate was returned to the batch tank. The solution in the batch tank became more concentrated with time to allow the system to simulate operation at different water recovery levels. Solid gypsum was not removed from the system, and the gypsum concentration in the feed solution increased with time.

3.3 Bench-Scale Studies

The scope of services performed by the Membrane Research Group at the University of Nevada, Reno, included bench-scale testing, evaluation of VEDCMD, and FO, for the concentration of Brines A and B. Modeling efforts on their part included cost estimates for a brine flow rate ranging from 0.4 to 1 mgd.

3.3.1 Membrane Distillation

Membrane distillation experiments were conducted with three types of membranes:

1. Capillary membranes in a tube-and-shell configuration to allow for tangential flow on both the feed and permeate sides of the capillaries.

2. Flat-sheet polypropylene membranes (GE Osmonics, Minnetonka, Minnesota).

3. Flat-sheet polytetrafluoroethylene (PTFE) membranes (GE Teflon® Laminated Membrane, GE Osmonics, Minnetonka, Minnesota).

The performance of the VEDCMD process was evaluated under various operating conditions using a bench-scale membrane test unit. A schematic of the bench-scale apparatus used in the VEDCMD investigation is illustrated in figure 3.4. Warm (40 and 60 °C) feed solution was circulated on the feed side of the membrane, and deionized water at a cooler temperature (20 °C) was circulated counter currently on the support side of the membrane. As water evaporated through the membrane, the concentration of the feed stream slowly increased, and the water level in the permeate reservoir slowly increased. Excess water overflowed the permeate reservoir and was continuously collected. The change in weight of the collection tank was recorded and used to calculate water flux and recovery. The permeate conductivity was monitored in order to calculate salt rejection. MD experiments were stopped when substantial flux decline was observed.

To treat scaling during experiments, four liters of deionized water was flushed through the feed side of the system to remove loosened precipitates. Two liters of an ethylenediaminetetraacetic acid (EDTA) cleaning solution was then recycled at 2 liters per minute (L/min) on the feed side of the membrane for 30 minutes. The system was then stopped, and the membrane was soaked in the cleaning solution for 1.5 hours. Subsequently, the feed channel was flushed with 8 liters of deionized water at 2.5 L/min.

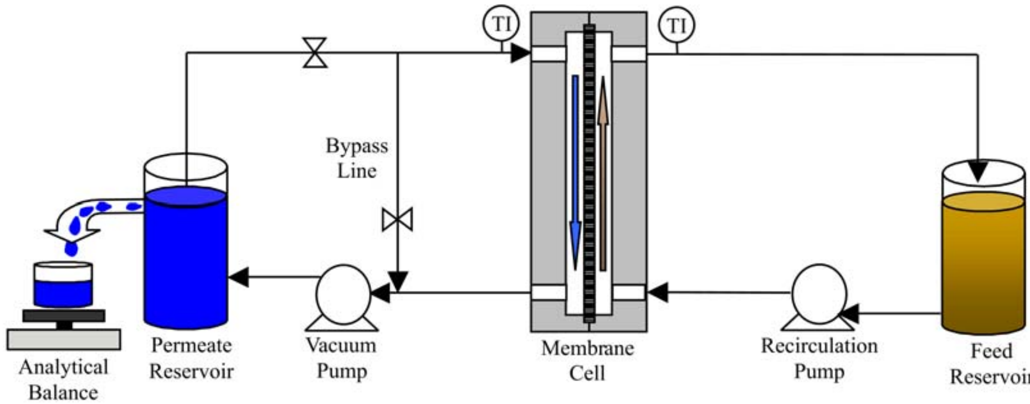


Figure 3.4 Bench-scale setup for vacuum enhanced direct contact membrane distillation.

A mathematical model was developed by the University of Nevada, Reno, to describe the pilot-scale VEDCMD process operated with commercial capillary polypropylene Microdyn membrane elements (MD020-CP-2N, Microdyn, Germany). Because there are no commercial membrane manufacturers for MD, a hydrophobic microfiltration membrane with a 0.2-micrometer nominal pore size was used. The numerical model developed considered both physical and thermodynamic forces and described the effects of both vacuum and flow velocity on the flux of water vapors across the membrane. Several membrane modules were modeled in series with feed reheating stages between membranes and once-through flow of cold water on the permeate side of each membrane element.

3.3.2 Forward Osmosis

The FO process performance was evaluated using a bench-scale membrane test unit coupled with a special pilot-scale RO system used to continually concentrate the draw solution (DS) at constant concentration (see figure 3.5). The feed solution, Brine A or Brine B, was circulated at 1.5 L/min on the feed side of the FO membrane, which was a flat sheet cellulose triacetate membrane. A sodium chloride draw solution at a concentration of 50,000 ±3,000 mg/L was circulated counter to the current at 1.5 L/min on the support side of the membrane.

The water level in the feed reservoir declined, and the TDS concentration of the feed stream increased as water from the brine diffused through the membrane into the draw solution. The change in brine volume was monitored and used to calculate water flux, recovery, and feed concentration. The feed solution was circulated on the feed side until water flux was substantially reduced. The DS inlet concentration was chosen to be 50,000 mg/L, and the feed concentration was 7,500 mg/L TDS for Brine A and 20,000 mg/L TDS for brine B, respectively. The DS and feed inlet flow rates were chosen arbitrarily and then adjusted to achieve the predetermined recovery.

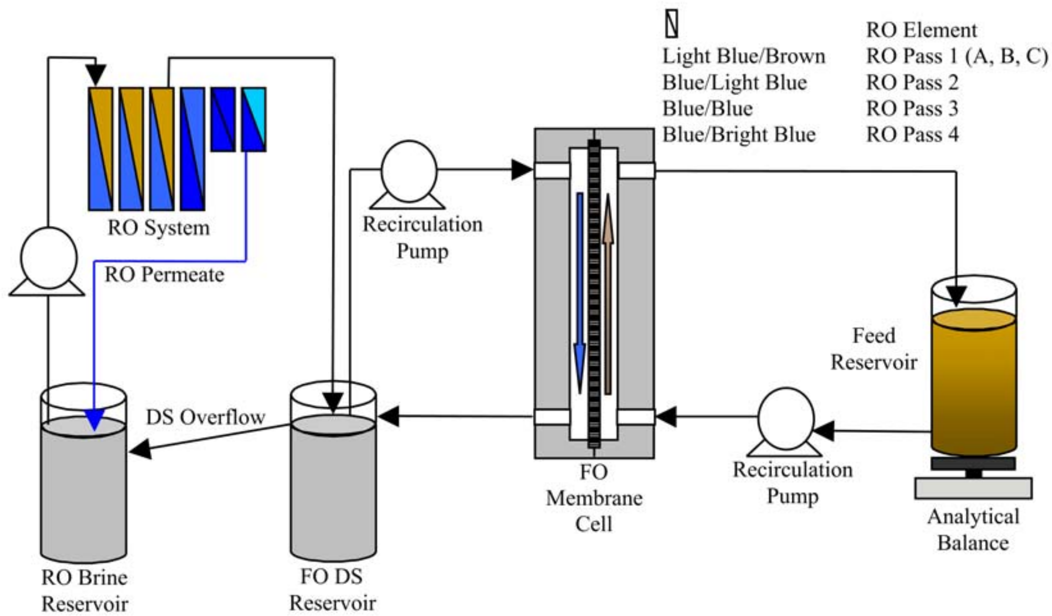


Figure 3.5 Schematic of forward osmosis testing system.

Membrane cleaning experiments conducted without the pilot-scale RO system to maintain the DS concentration were used to investigate the effectiveness of membrane cleaning. The membrane samples were analyzed by scanning electron microscopy (SEM) to determine the major constituents of membrane surface scale.

A model was developed by the University of Nevada, Reno, to estimate the energy demand and major operating costs required to treat 1 mgd of brackish water RO brine at a total water recovery of 70 percent. The Excel-based model was used in conjunction with a commercially available RO model (reverse osmosis system analysis [ROSA], Dow-Filmtec, Midland, Michigan). The FO model was developed as a vessel containing two membrane elements connected in a series. The number of vessels used in parallel was determined using the amount of brackish water to be processed and the expected water recovery. The model did not account for concentration polarization, temperatures, and scaling effects. Subsequently, the outlet DS concentration and flow rate were used as inputs into ROSA where a one-stage membrane array was modeled. The feed concentration was obtained from the FO model, and the permeate stream was the final product water. The RO system recovery was determined based on requirements to produce a reject stream concentrated to 50 g/L NaCl. The SW30HR LE-400i (Filmtec) spiral wound RO membrane element was selected for the model. The model RO system was then scaled up to produce product water at a total recovery of 68 percent with varying input capacities.

3.4 Desktop Modeling

Some of the ZLD processes considered for this project were analyzed as part of a desktop studies. Equipment suppliers and patent holders on the following processes were contracted to provide cost data and estimates of system effectiveness in reducing brine streams.

3.4.1 Residual Recovery (SAL-PROC)

As part of the EMWD/Reclamation component of brine concentration studies, Geo-Processors USA, Inc., (Geo) was engaged to perform a desktop pre-feasibility study of the application of the SAL-PROC™ process for recovery of useful byproducts and to achieve a ZLD. A nominal 1-mgd flow rate was used for analysis for this desktop study.

Brine water quality data was transmitted to Geo for evaluating the SAL-PROC™ process to treat and recover valuable byproducts from two reference brine streams with significant differences in their salinity and chemical makeup. In addition, Geo developed the recommended ZLD processing schemes for further consideration by EMWD as cost-effective RO brine management options.

Based on a 1-mgd flow rate, the annual salt load of Brines A and C were approximately 10,000 and 34,000 tons per annum, respectively. These salt loads were considered in this study as one of the parameters for conceptual design and sizing of the SAL-PROC plants and followup cost estimations.

The study was conducted by inputting the nominal concentrate flow and loadings into desktop modeling software developed by Geo to identify a number of technically feasible ZLD process systems. Each of the process systems defined by the SAL-PROC™ model is comprised of two subsystems, including one or more selective salt recovery steps that are linked with RO desalination, thermo-mechanical brine concentration, and crystallization steps. The desktop modeling exercise enabled the selection of the most appropriate ZLD process schemes as recommended options for the two reference brine streams. The selected ZLD systems were to utilize multiple reaction steps using lime and soda ash reagents to produce carbonated magnesium, calcium carbonate, and a mixed salt. The systems were to recover a large percentage of the flow as potable water. In addition, a crystallizer was proposed in preference to evaporation ponds to minimize space requirements.

3.4.2 Brine Concentration and Crystallizer

HPD-Veolia Water Solutions was provided with water quality data for Brine A (primary RO brine) and B (secondary RO concentrate produced from caustic/soda ash softening of Brine A). Veolia Water Solutions responded with budget estimates for treating both brine streams using a combination of BC only and brine concentrator/crystallizer (BC/XLZR) for treatment capacities ranging from 0.125 to 1 mgd.

3.4.3 Evaporation Ponds

An Excel-based model was developed based on a report from Reclamation (Mickley & Associates, 2001) and an in-house cost analysis previously developed at Carollo Engineers (2004). This model was used to estimate evaporation pond costs using the local (Eastern service area) net evaporation rate of 70 inches per year (Lee et al., 1992).

3.5 Development of Cost Estimates

Costs for treatment alternatives were developed in terms of capital and O&M costs. Costs were based upon recent projects, vendor information, and standard cost estimating curves. The actual cost of a project can vary based on the material specified, labor, competitive market conditions, and other variable factors. For these reasons, it is possible that actual construction costs may vary from the conceptual estimates shown herein.

3.5.1 Cost-Curve Assumptions

In the development of cost curves for each treatment option for various capacities, some common cost variables were assumed for each process. The list of assumptions is presented below:

1. Based on field tests and experimental results, 70-percent recovery was assumed for secondary RO, FO, and MD.

2. A recovery of 75 percent was assumed for EDR operation.

3. A recovery of 60 percent was assumed for the seeded RO process.

4. Market price for water regenerated: $549 per acre-foot based on EMWD quoted values.

5. The average annual interest rate is assumed to be constant at 6 percent over a loan period of 20 years for the calculation of annualized capital costs.

6. Cost of electricity: $0.12 per kWh.

7. Disposal cost of residuals solids to landfill: $50 per ton based on past EMWD projects.

8. Labor cost: $47.75 per hour.

Percentages (typically part of total direct costs) used in the development of cost estimates for this study were:

1. Interconnecting pipework: 12.50 percent.
2. Electrical and instrumentation: 18 percent.
3. Engineering: 15 percent.
4. Legal and administration: 10 percent.
5. Contingency: 25 percent.

The costs presented are in February 2007 dollars (Los Angeles ENR[2] = 8,871) and are not escalated for future construction or operation.

[2] ENR is a periodical, Engineering News Record, which published construction cost indicies. The ENR construction cost index for Los Angeles is 8,871 as of February 2007.

4. Results and Discussion

4.1 Primary Brine Water Quality

As expected, the Menifee brine (Brine A) is extremely hard water with significant scaling potential. Hardness and silica averaged 3,500 mg/L as calcium carbonate and 160 mg/L as silica, respectively. The greatest contribution to hardness was calcium ions, which contributed to over 70 percent of the hardness of the primary brine. Silica consisted almost exclusively of dissolved silica. Other constituents such as iron, manganese, and heavy metals were present in sufficiently low quantities in that they were not expected to pose a scaling threat to the secondary RO membranes, even at the desired recovery level of up to 75 percent. As anticipated, the TDS content of the raw brine was extremely high, ranging from 4,320 to 7,930 mg/L. Table 4.1 presents the average Brine A water quality based on samples collected over a 6-month period as part of the pilot testing of the softening and secondary desalting processes. The table also includes detailed chemical analyses for Brines B and C. Grab samples were collected three times per week over a 6-month period in half-gallon bottles by Carollo and delivered to the EMWD lab for water quality analysis.

Table 4.1 Summary Water Quality Data for Three Brine Types[1]

Parameter	Units	Detection Limit[2]	Brine A 1° RO Brine	Brine B 2° RO Brine	Brine C SPARRO Brine
pH	pH units	-	7	7.2	7.2
Total Alkalinity	mg/L as $CaCO_3$	3	652	188	259
Chlorine	mg/L	1	2,439	9,891	10,598
Sulfate	mg/L	1	462	2,202	3,338
Calcium	mg/L	1	994	2,200	1,550
Hardness[3]	mg/L as $CaCO_3$		3,470	6,222	11,000
Magnesium	mg/L	1	234	614	684
Dissolved Silica	mg/L as silicon dioxide (SiO_2)	1	165	166	230
Sodium	mg/L	10	873	4,142	5,476
Total Dissolved Solids	mg/L	25	5,701	18,605	22,264
Electrical Conductance	µS/cm	1	8,900	30,309	45,267

[1] Calculations assume values for nondetect results are at the detection limit. Calculations are based on EMWD lab analysis for the period June 23, 2006, to December 4, 2006. Reported values are average values from multiple data sets, wherever possible.
[2] Based on reporting detection limit as provided by EMWD laboratory.
[3] Reported hardness of 23,000 mg/L as $CaCO_3$ on November 1, 2006, omitted as aberration.

4.2 PILOT STUDIES

The results presented in this section are summarized from the preliminary results obtained from a project partially funded by the California DWR being undertaken by EMWD on testing of chemical softening followed by RO and EDR.

4.2.1 Chemical Softening and Secondary Desalting

The chemical softening process was described in section 3. Table 4.2 presents a summary of the average water quality before (Brine A) and after chemical softening for selected parameters. Values are based on 6 months of operating data.

Table 4.2 Summary of Selected Water Quality Results from Chemical Softening

Parameter	Units	Raw Primary RO Brine Feed (Brine A) (Average)	Softened Brine (Average)
pH	—	7.0	6.6
Bicarbonate	mg/L	792	73
Total Alkalinity	mg/L as $CaCO_3$	652	67
Chlorine	mg/L	2,439	2,341
Sulfate	mg/L	462	495
Calcium	mg/L	994	334
Magnesium	mg/L	234	138
Total Hardness	mg/L as $CaCO_3$	3,470	1,399
Dissolved Silica	mg/L as SiO_2	165	48
Total Dissolved Solids	mg/L	5,701	4,501

As summarized in table 4.2, the chemical softening process was very effective at reducing the influent calcium concentration by approximately 66 percent to an average of 334 mg/L. Likewise, the influent silica concentration was reduced by approximately 70 percent to an average of 48 mg/L. Magnesium was reduced by approximately 41 percent throughout the softening tests and was required for silica precipitation. The reduction of calcium and silica was key in determining the success of the softening step. The silica concentration in the softened water was at a level comparable to the raw well water that feeds the primary RO plant, which was important so that secondary RO recovery would not be significantly limited. Altogether, the reduction in calcium and magnesium resulted in an overall reduction in hardness of about 60 percent. Overall, the TDS reduction was more than 1,000 mg/L.

Softened water quality was used in the ROSA software to provide a preliminary estimate of the level of recovery that could be obtained by the downstream secondary RO process. The models predicted that the secondary RO process could operate at up to about 75-percent recovery. This recovery was higher than that of the primary RO plant.

Overall, the chemistry of the chemical softening process confirmed that reduction of calcium and silica could be obtained. Operation of the pilot plant did indicate that significant challenges associated with this process are the removal of the residual fine solids in the softened water that can impact the downstream treatment processes, as well as handling the large volumes of sludge produced by the softening step. Further work is planned to specifically address these challenges.

4.2.2 Reverse Osmosis

Initially, the RO pilot plant was operated at a modest recovery of approximately 40 percent. However, the recovery was quickly increased so that, for most of the RO operation, recovery was above 70 percent, with the highest recovery achieved around 78 percent. Figure 4.1 shows the variation in recovery with operating time.

Table 4.3 shows the average quality of RO permeate based on between 20 and 30 data points for the parameters shown. The secondary RO was able to produce a high quality permeate stream (TDS less than [<] 300 mg/L) which could be recovered as potable water.

Table 4.3 Summary of Selected Water Quality Results for RO Permeate

Parameter	Units	RO Permeate (Average)	Number of Samples
pH	—	5.6	14
Bicarbonate	mg/L	4	30
Total Alkalinity	mg/L as $CaCO_3$	4	31
Chlorine	mg/L	166	17
Sulfate	mg/L	2	17
Calcium	mg/L	1	27
Magnesium	mg/L	1	27
Nitrate (as NO_3)	mg/L	23	17
Sodium	mg/L	94	27
Dissolved Silica	mg/L as SiO_2	1	27
Total Dissolved Solids	mg/L	282	31

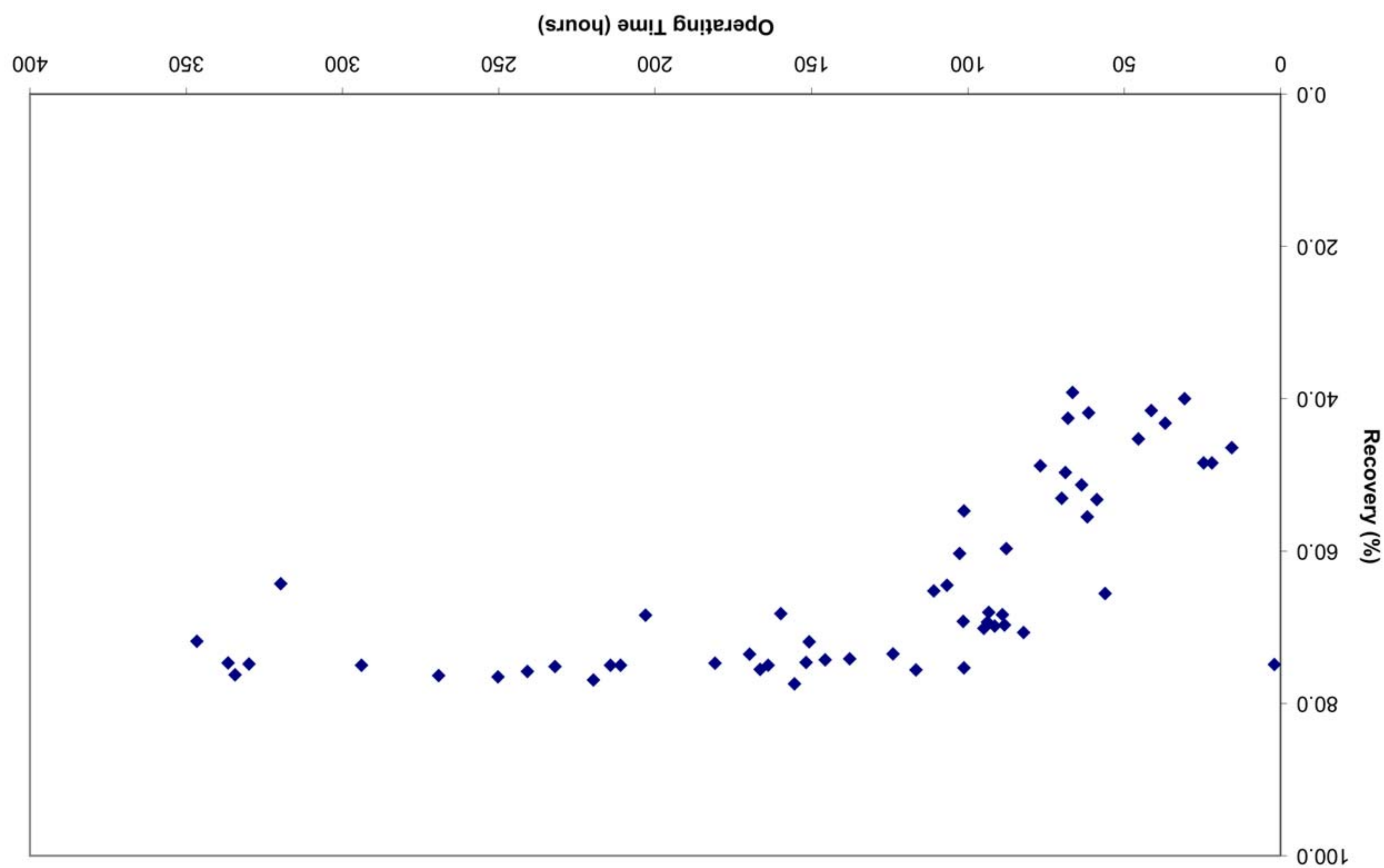


Figure 4.1 Recovery of secondary RO pilot operation.

Operating data showed that the first stage RO membranes did most of the treatment work. This was particularly true as the recovery level increased. Given that the feed TDS to the membranes was about 4,500 mg/L, the TDS leaving the first stage membranes was over 12,000 mg/L at a recovery of about 70 percent.

At the start of the project, it was not clear what recovery would be achieved on the RO pilot plant, and a recovery of 60 percent was considered reasonable. The RO pilot commissioned was designed for a 65-percent recovery, and the maximum operating pressure was pre-set to a maximum of 250 psi. The success of hardness removal in the softening step was beyond our expectation and increased RO recovery to 65 (and sometimes 70) percent. Estimates of the first stage osmotic pressure showed that operating pressures could be over 200 psi; thus, there was little driving pressure for the second-stage membranes. Pushing the feed pressure over 250 psi would result in triggering the alarm of the RO skid and, ultimately, shut off the system. Consequently, controlling the operation of the second stage was difficult, as the permeate flows were very low at the higher recovery values.

The first-stage membranes maintained salt rejection throughout the operating period. However, there was an upward trend in the feed pressure to the first-stage membranes, indicating that some fouling had occurred. The net permeate production from the first-stage membranes indicated a gradual downward trend, supporting the conclusion that the membranes were fouling.

4.2.2.1 Membrane Autopsies

Separate cleaning events were conducted on both membrane stages, although cleaning did not seem to noticeably improve the membrane performance for any significant time. Upon completion of pilot testing, the lead and lag elements from each of the two stages were removed from the vessels and sent to Filmtec Corporation for testing. These four elements were selected since they were expected to show the incremental deterioration of the membranes along the train.

The membranes were subjected to three different tests. The first test was a nondestructive test for observation of the physical integrity of the element and visually identifying potential foulants. Elements were visually inspected noting any differences from the new product. Stress marks were visible on the feed side product water tubes of both lead elements (stages 1 and 2). With the exception of stage 1-lead, which had a yellowish foulant on the membrane surface, the outer surface of the membranes and the membrane leaves were covered with a white, “gritty” foulant. The amount of white foulant was more visible on the second-stage elements than first, indicating that these membranes were subjected to a greater degree of colloidal fouling. The white foulant was partially removed with acid.

The second test was a Filmtec brackish water baseline test. This test uses a 2,000-mg/L NaCl solution at 77 degrees Fahrenheit (°F) with an applied pressure of 150 pounds per square inch gauge (psig). With the exception of stage 1-lead element, all elements exhibited a severe decline in permeate flow, as compared to product specification. Permeate flow of the stage 1-lead was nearly 98 percent of the expected value. By comparison, permeate flow from the stage 1-lag, stage 2-lead, and stage 2-lag elements were less than 30 percent of the expected value. The stage 2-lag element in particular was completely fouled, and permeate flow was only 1 percent of that of a new element. Similar trends were observed with the salt rejection capacity of the membranes; the stage 1-lead element preserved its salt rejection capacity (100 percent of new element), while the capacity of the other three elements was severely impaired, as low as 10 percent in the stage 2-lag element. The pressure drop across the lead elements compared favorably with specifications, while the lag elements experienced very high pressures under the test conditions, 1.6 times and 2.4 times specification for stages 1 and 2 elements, respectively.

Finally, membrane coupons were analyzed for metals via inductively coupled plasma in a third test. Levels of calcium, magnesium, sodium, and silica foulants were found to be 24.5, 21.6, 137.2, and 3,282 milligrams per square meter, respectively. The silica value is significant since it shows the severe amount of silica scaling experienced by the membranes. A section of the membrane was also scraped to accumulate enough of the gritty white foulant for testing. The white salt was determined to be calcium sulfate.

It is clear that the membranes experienced extreme fouling with respect to colloidal fouling (silica) and calcium sulfate scaling. During pilot testing, it was observed that early caustic soda and hydrochloric acid cleaning cycles were capable of restoring permeability to the membranes. This was not sustained for any significant period; and towards the end of testing, it was nearly impossible to restore permeability. The membrane autopsy results suggest an explanation for this observation. In the early phases of testing, calcium sulfate was the primary cause of permeate flow decline. Since it responded well to cleaning, permeability could be restored with cleaning. Towards the end of testing, silica scale dominated; and the cleanings were no longer effective in restoring membrane permeability.

In general, membrane fouling was attributed to poor RO influent water quality in terms of particulate matter. Frequent upsets of the upstream softening and filtration processes led to solids carryover into the RO unit. This resulted in the RO unit receiving water that was out of specification in terms of SDI (< 3). Towards the end of the pilot testing, the problems associated with upstream conventional softening were mostly resolved. However, by this time, colloidal and silica fouling had irreversibly damaged the membranes.

4.2.3 Electrodialysis Reversal

The EDR unit was operated for 106 hours in the 168 days of pilot operation. This unit arrived onsite later than the RO plant and required some maintenance soon after startup to deal with the high TDS feed water.

The initial pilot test plan had proposed blending softened water with raw brine to feed the EDR membranes. This was theoretically possible because, unlike the RO membranes, EDR membranes are not affected by the high silica level present in the primary RO plant brine (Brine A). However, this approach was not successful due to the release of gas from the primary RO brine in the feed lines of the EDR unit, which caused hydraulic control problems. Thus, for essentially all of the operating time on the EDR unit, the plant received softened water.

As discussed in chapter 3, the EDR plant was provided with a stack that consisted of four hydraulic stages and two electrical stages. For the first phase of operation, recovery exceeded 80 percent during both positive and negative polarities. In the later phase of operation, recovery was maintained at approximately 60 percent and 75 percent during negative and positive polarities, respectively.

During early operation, the rejection of the EDR process was relatively low and averaged only 70 percent. As a result, the permeate conductivity exceeded 2 µS/cm for this period, translating to a TDS of nearly 1,400 mg/L. However, after electrical setpoint changes were made to increase the voltage across the first stage of membranes, the rejection improved. Rejection increased to 85 percent; and permeate conductivity initially fell to less than 600 µS/cm (approximately 400 mg/L of TDS). Table 4.4 shows the average quality of the EDR permeate for the entire operating period. The average TDS is a lot greater than that of the RO process, due to some high values before the EDR electrical stages were modified. During the latter operating phase of the plant, the TDS averaged 415 mg/L, which could be recovered as potable water.

During startup operation, stage voltages were set at 75 and 65 volts for stages 1 and 2, respectively. Voltages were reset to 130 and 75 volts during the second stage of operation to facilitate increased rejection of the membranes. A natural consequence of the increased voltage across the membranes was an increase in the calculated resistance.

The EDR unit was not run long enough to determine the long-term operating impacts of the higher voltage on the membranes and potential fouling by the feed water. Future tests are planned to assess long-term fouling.

Table 4.4 Summary of Selected Water Quality Results for EDR

Parameter	Units	EDR Permeate (Average)	Number of Samples
pH	—	7.2	5
Bicarbonate	mg/L	116	7
Total Alkalinity	mg/L as $CaCO_3$	120	8
Chlorine	mg/L	611	7
Sulfate	mg/L	57	7
Calcium	mg/L	222	8
Magnesium	mg/L	66	8
Nitrate (as NO_3)	mg/L	13	7
Sodium	mg/L	396	8
Dissolved Silica	mg/L as SiO_2	52	7
Total Dissolved Solids	mg/L	415	9

Based on the results of this study, the EDR recovery is assumed to be 75 percent, and it is assumed that a similar permeate water quality (in terms of TDS) to the RO process could be obtained. EDR stack operating pressures, stack voltages, and amperage readings from the pilot plant were used as a basis for calculating the operating costs of the system.

4.2.4 SPARRO (Seeded RO)

A summary of the water quality for the feed water to seeded RO treatment (i.e., Brine B – prior to seeding), product and brine (i.e., Brine C) are presented in table 4.5.

Table 4.5 Summary of Water Quality Data for Seeded RO Process on Brine B

Parameter	Units	Feed (Brine B)	Product	Brine (Brine C)
TDS	mg/L	18,605	10,423	22,264
Na^+	mg/L	4,142	1,677	5,477
Ca^{2+}	mg/L	2,200	950	1,550
Mg^{2+}	mg/L	614	280	684
Cl^-	mg/L	9,891	5,723	10,599
SO_4^{2-}	mg/L	2,202	615	3,338
HCO_3^-	mg/L	223	97	316

From the secondary RO process, brine B has a calcium sulfate precipitation potential of 150 percent prior to seeded RO treatment, which indicated that the feed solution was supersaturated with respect to gypsum. In figure 4.2, the net driving pressure and the osmotic pressure of the system is plotted against operating time. In conventional RO processes, data analysis is predicated on the system operating at either constant flux or constant feed pressure. If constant flux is established, the feed and transmembrane pressures increase in response to membrane fouling. The reverse is also true; if feed pressure is kept constant, fouling is evidenced by a decrease in flux. For seeded RO experiments, the net driving pressure showed a linear increase from 210 to 290 psi as operating time increased from 0 to 180 minutes, since the increase in osmotic pressure of the solution from 160 to 216 psi necessitated higher pressure to push the brine through the membrane to obtain constant flux. Seeded RO experiments were performed for a total of 180 minutes, and aliquots of the feed, product, and brine were analyzed for the composition. There were instances where the solution level in the tank dropped to below operable limit by 180 minutes, and no sampling could be performed. At the time when brine composition data was sent to Geo-Processors for SAL-PROC™ desktop modeling, the most complete set of data available was for T=120 minutes; and hence, all experiments and desktop modeling based on Brine C composition was for T = 120 minutes.

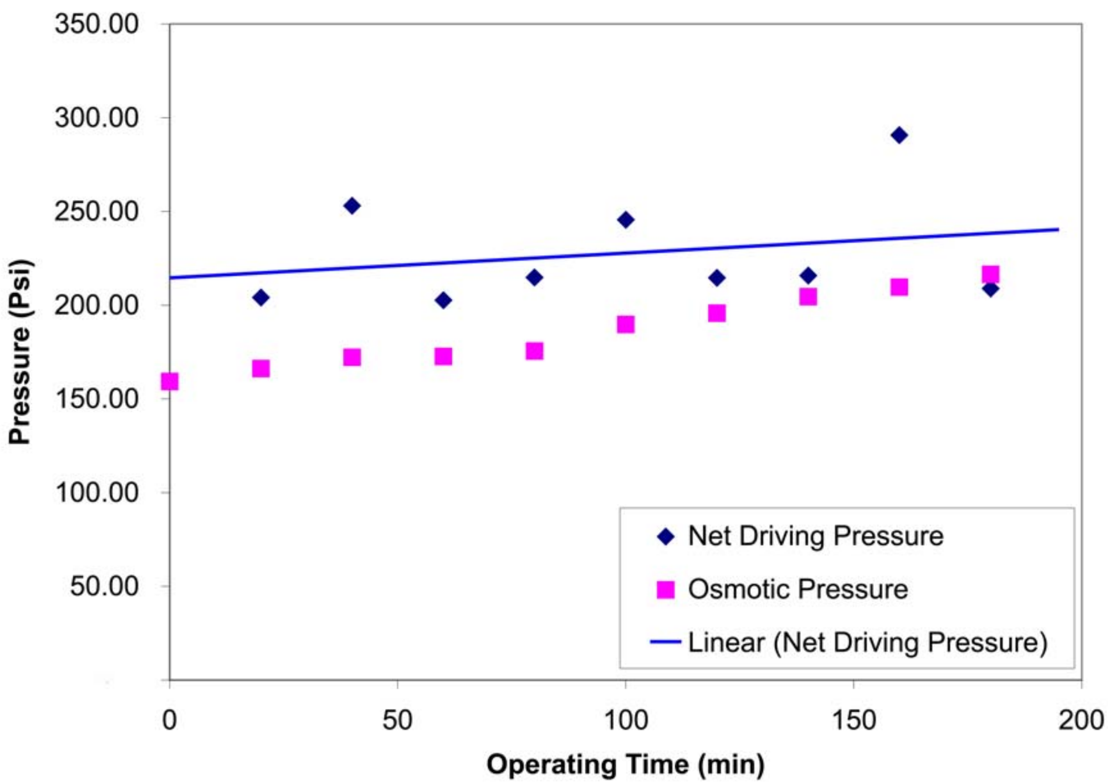


Figure 4.2 Seeded RO net driving and osmotic pressures for Brine B.

The normalized permeate flux in figure 4.3 showed a relatively constant production rate of water throughout the process.

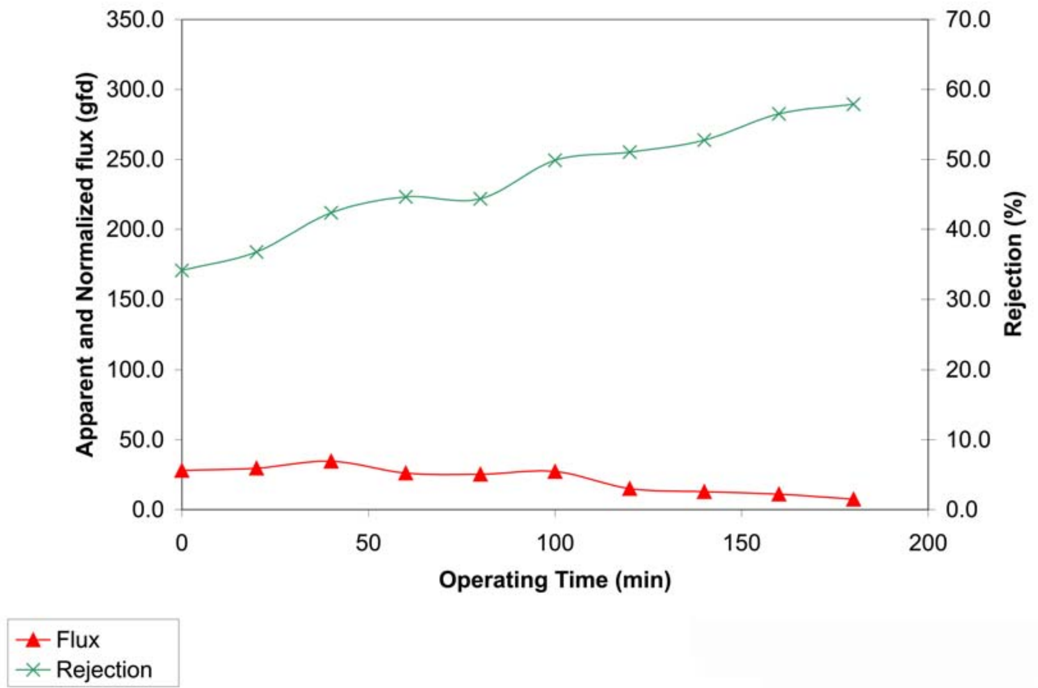


Figure 4.3 Seeded RO permeate flux and rejection for Brine B.

An overall rejection of 50 to 60 percent was achieved by the seeded RO system. The membrane specific flux is illustrated in figure 4.4. Membrane specific flux is defined as the normalized flux divided by the net driving pressure. While the flux remained relatively constant, the net driving pressure increased due to osmotic pressure; and as a result, the membrane specific flux decreased over time.

The highest recovery that was achieved during operation was about 60 percent, as shown in figure 4.5. The seeded RO is, in essence, a batch operation with recycle; and, hence, a linear trend in recovery from 0 to 60 percent was observed. The recovery was limited by the size of the equipment and not by membrane scaling. After 180 minutes of operation, the feed volume in the tank had decreased to about 40 gallons. At this tank volume, the impeller on the tank mixer was no longer totally submerged; and therefore, the system had to be shut down to prevent settling of the gypsum seed crystals. In a larger-sized plant, it is expected that recoveries of over 60 percent can be obtained. For cost modeling in the subsequent portions of this study, the cost analysis will be based on the 60-percent recovery that was achieved during pilot plant testing.

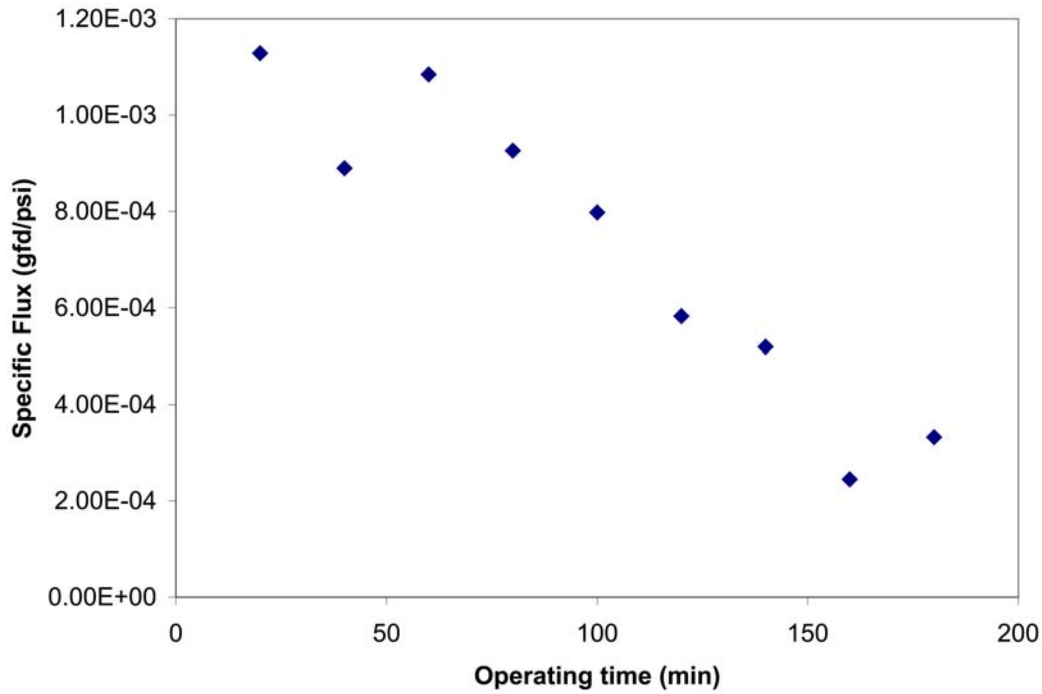


Figure 4.4 Seeded RO specific flux for Brine B.

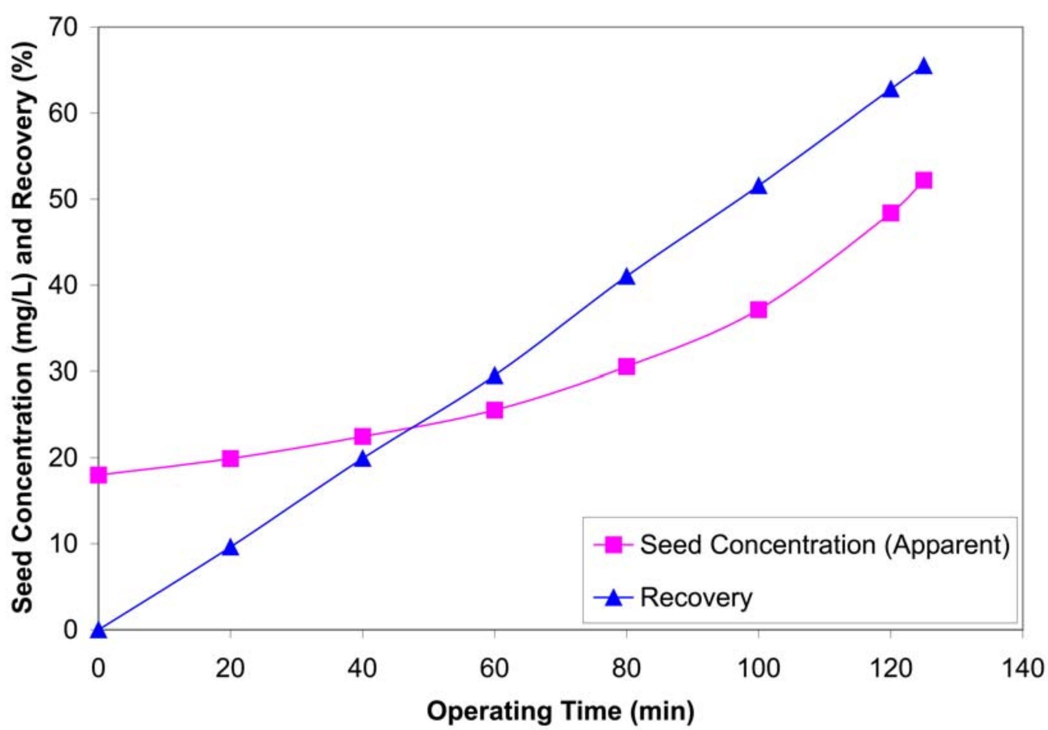


Figure 4.5 Seeded RO recovery and apparent seed concentration for Brine B.

The success of the seeded RO technique could be inferred not only from the apparent concentration increase of gypsum seeds in the system as shown in figure 4.5, but also from SEM imaging and energy dispersive x-ray (EDAX) analysis of the resulting gypsum seed. The presence of crystallites in the 1- to 5-µm size range on larger gypsum seeds (10 to 50 µm) (figure 4.6) indicated that mineral salts precipitated on the seed crystals. EDAX analysis (figure 4.7) confirmed that only calcium and sulfate precipitation occurred, shown by the large “Ca,” “O,” and “S” identification peaks, indicating a relatively pure byproduct.

The integrity of the tubular NF membrane was intact for the duration of the seeded RO process testing of 6 months, as demonstrated by the data shown in figure 4.8. The permeate conductivity was monitored throughout the pilot-testing duration and remained constant at around 21 µS/cm. In addition, the clarity of the permeate stream was monitored throughout plant operation. If the seeded slurry had punctured the membrane surface, the damage would translate to an increase in the permeate conductivity and/or visible turbidity in the water. No such observations were made during the course of testing.

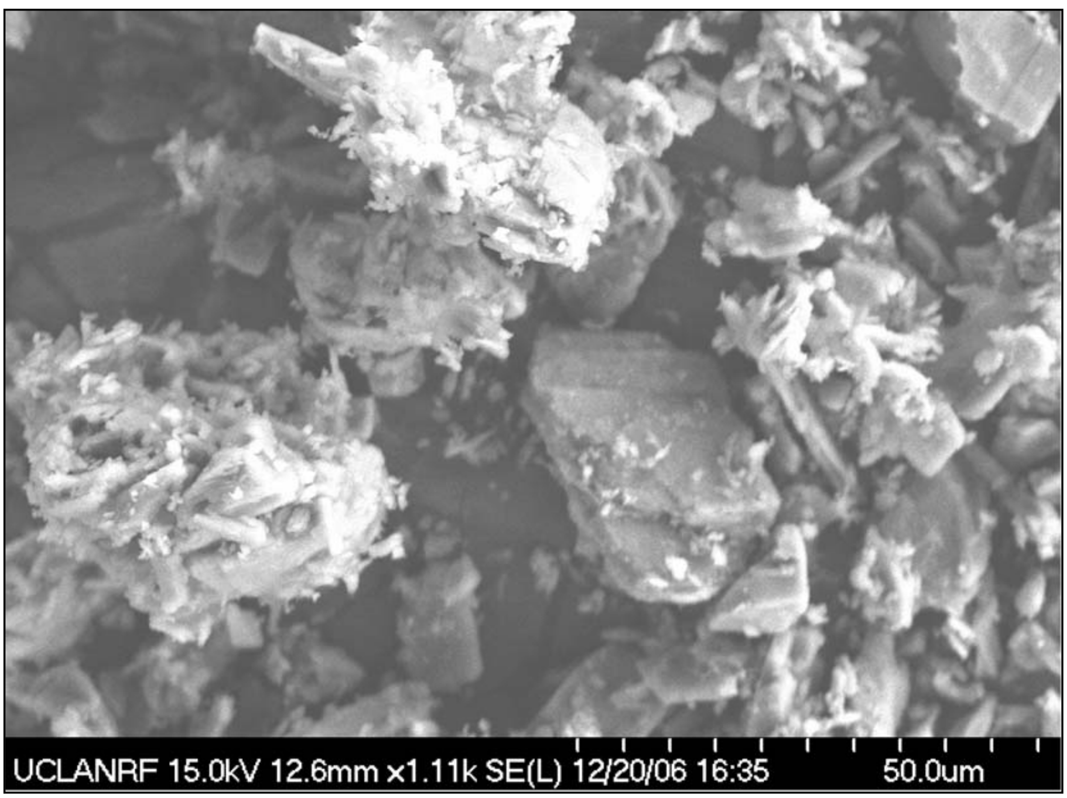


Figure 4.6 SEM of gypsum seeds from seeded RO tests.

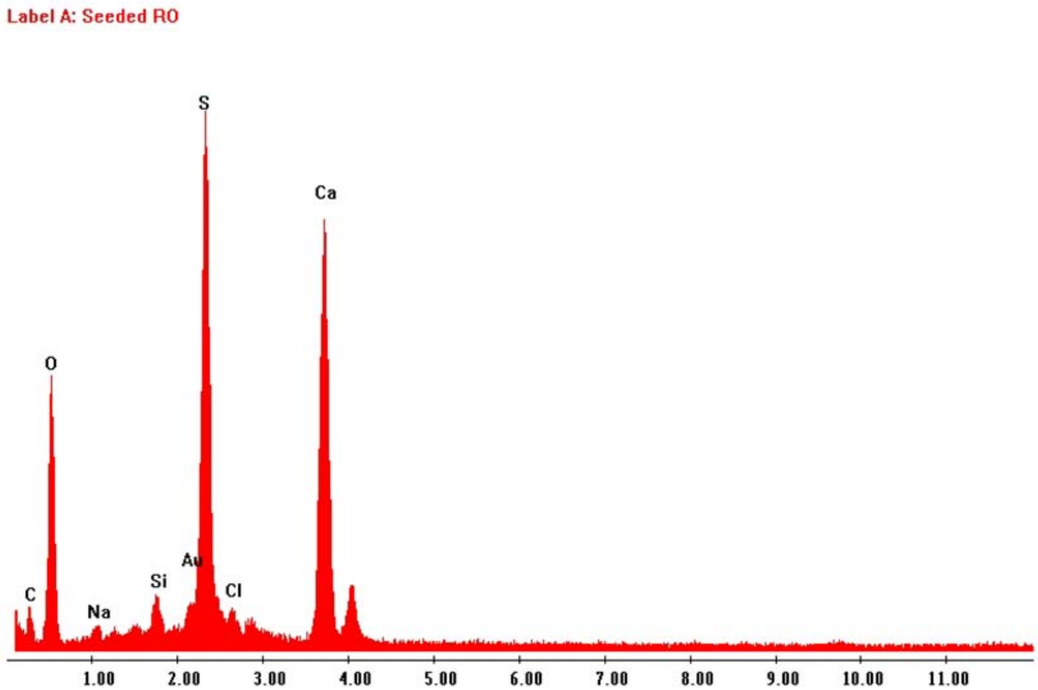


Figure 4.7 EDAX analysis of seed crystals from seeded RO tests.

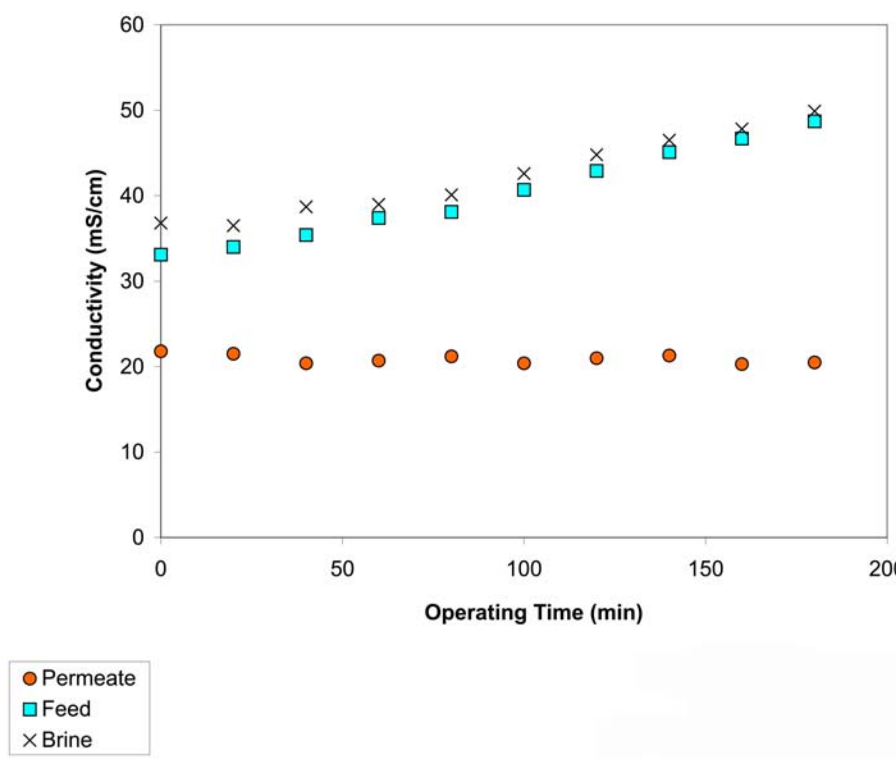


Figure 4.8 Seeded RO feed, permeate, and brine conductivity for Brine B.

4.3 Bench-Scale Studies

4.3.1 Membrane Distillation

VEDCMD experiments were performed using Brines A and B as feed solutions. The effect of feed concentration on water flux is illustrated on figure 4.9. Initial water flux was substantially greater in experiments conducted at ΔT = 40 °C than those conducted at ΔT = 20 °C. This is because a larger ΔT generates a greater vapor pressure gradient across the membrane and results in increased water flux. Water flux was further improved by decreasing the permeate vapor pressure from 660 to 360 mmHg.[3] By decreasing the permeate pressure, the partial vapor pressure gradient across the membrane is increased, resulting in higher water flux.

Membrane scaling at elevated feed concentrations (Brine B at about 20,000 mg/L TDS) severely reduced the water flux through the Microdyn capillary membrane. Water flux was only partially recovered following a membrane cleaning. Subsequent experiments were performed using the flat-sheet polytetrafluoroethylene membrane and polypropylene membrane.

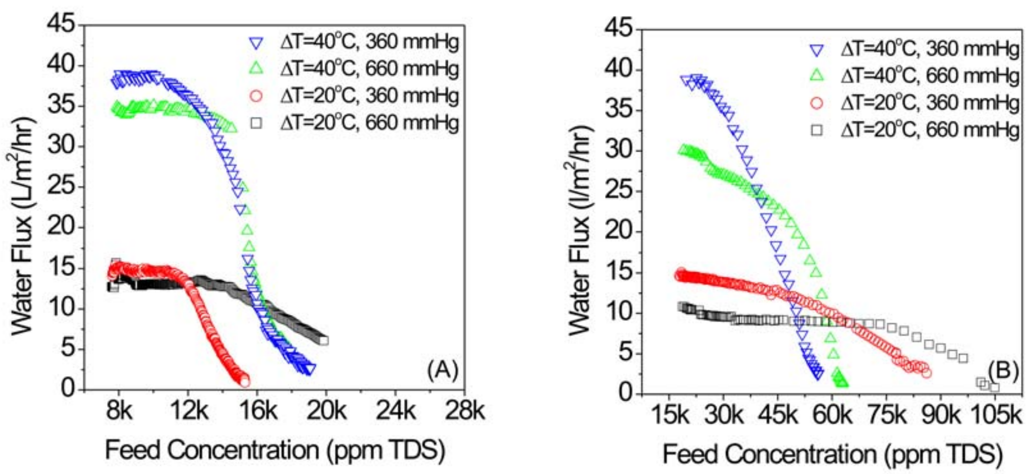


Figure 4.9 Water flux versus feed concentration for MD of (A) Brine A and (B) Brine B.

For Brine A (figure 4.9a), a relatively constant water flux was observed during the initial stage of the concentration experiments. Between 10,000 and 13,000 mg/L TDS, a sharp flux decline was observed. Because previous VEDCMD experiments conducted with two different feed solutions (NaCl and synthetic seawater salt) demonstrated a gradual (9 percent) water flux decline over a feed concentration range of 0 to 75 g/L (Cath et al., 2004), it is likely that the flux decline in the current investigation was due to scale formation on the

[3] mmHg = conventional millimeter of mercury.

membrane surface formed when sparingly soluble salts precipitated out of solution. The decline continued until the water flux was below 5 liters per square meter per hour and the experiment was terminated.

In the experiments conducted with Brine B (figure 4.9b), with the exception of one of the experiments conducted at the lower pressure gradient across the membrane, flux decline started after the feed was only minimally concentrated. The higher water flux associated with the increased pressure gradient likely increased concentration polarization at the membrane surface and resulted in earlier supersaturation and onset of water flux decline.

The results suggest that feed concentration has minimal effect on water flux in VEDCMD processes, but eventually scale formation at higher feed concentrations is detrimental. Based on previous VEDCMD experiments (which had higher feed concentrations), water flux was expected to be relatively constant; and approximately 90-percent total water recovery was anticipated (Cath et al., 2004). However, due to membrane scaling, the highest total water recoveries obtained for Brines A and B were 62 and 81 percent, respectively. The higher recovery achieved for Brine B appears counter-intuitive at first, but consistent experiment results suggest that the presence of residual antiscalant in the water composition is responsible for the delay of scaling when compared to Brine A.

When considering the full process recovery, overall recovery of 88.8 percent for the consecutive RO treatments (primary RO at 70 percent and secondary RO at 62 percent) plus recovery from VEDCMD, the numbers are more encouraging. Using equation (1), the overall system recovery for Brine A would be 88.6 percent; and the overall system recovery for Brine B would be 97.9 percent.

$$R_{tot} = R_{RO} + (1 - R_{RO})R_{VEDCMD} \tag{1}$$

Figure 4.10 presents water fluxes during VEDCMD of Brines A and B as a comparison of two feed temperatures and two different permeate vapor pressures. Initial water flux was very similar for Brines A and B at each temperature and pressure, which suggests that feed concentration has a minimal effect on water flux in VEDCMD. Total water recovery achieved in the experiments conducted at $\Delta T = 40$ °C ranged from 60 to 65 percent and from 60 to 81 percent at $\Delta T = 20$ °C. At $\Delta T = 40$ °C, the increased water flux lead to faster accumulation of salt ions on the membrane surface, and hence a lower recovery compared to experiments conducted at $\Delta T = 20$ °C.

Although MD processes have the substantial advantage of low impact of feed water salinity on membrane performance, there are no established large-scale applications of MD for seawater or brackish water desalination. Furthermore, there are currently no commercial membranes made specifically for

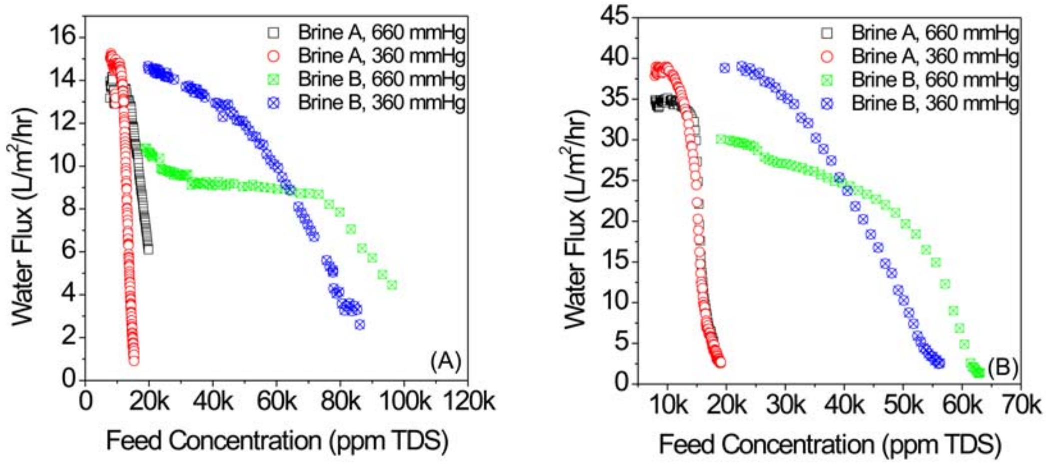


Figure 4.10 Comparison of Brine A and Brine B MD results at (A) ΔT = 20 °C and (B) 40 °C.

MD applications. Instead, membranes that are used for MD applications are microfiltration membranes that have specific pore size and are made of hydrophobic materials. For estimating the cost of brackish water brine desalination, an off-the-shelf membrane from Microdyn (MD 150 CP 2N) was used. This membrane has 1,800 polypropylene capillary membranes installed in a tube-and-shell configuration for a total membrane area of 10 m^2 at a retail cost of \$5,000 to \$7,000.

Under the conditions investigated in the current study, the Microdyn membrane can generate from 2 to 4 liters per square meter per hour (L/m^2-hr) (1.2 to 2.4 gallons per square foot of membrane per day [gfd]) at feed and permeate temperatures of 40 °C and 20 °C, respectively, which is equivalent to 750 liters per day (198 gpd) per membrane element. At feed and permeate temperatures of 60 °C and 20 °C, respectively, the membrane flux rate increases to a range of 6 to 8 L/m^2-hr (3.6 to 4.7 gfd), which produces approximately 1,600 liters per day (432 gpd) per membrane element. To achieve this performance, the flow rate of both the feed and permeate streams into each membrane element must be on the order of 250 liters per minute (66 gpm). Using a feed flow rate of 66 gpm and a daily production rate of 198 gpd when operating at a feed temperature of 40 °C, the recovery is approximately 0.2 percent. Likewise, using a feed temperature of 60 °C increases the expected recovery of the Microdyn element to 0.4 percent (432 gpd/66 gpm).

The above recoveries using MD are very low. Assuming a 70-percent water recovery from Brine A, approximately 31,000 Microdyn elements would be required to produce 0.7 mgd fresh water from feed water at 40 °C; approximately 15,000 Microdyn elements would be required to produce 0.7 mgd fresh water from feed water at 60 °C.

In each case, cold water (at approximately 20 °C) would have to be pumped to the permeate side of the membrane at a rate of 200 liters per minute (53 gpm) in every membrane element, which is equivalent to total pumping of more than 1,100 mgd at a pressure of approximately 10 psig. This flow rate is clearly not practical but can be reduced if several membrane elements can be connected in series and have cooling stages between elements to maintain temperatures of approximately 20 °C in the inlet of each element. On the feed side, assuming that 25 membrane elements can be connected in series with reheating stages in between, a feed water rate of approximately 118 mgd would have to be distributed between 1,240 vessels at a pressure of approximately 10 psig. This is a significant reduction but is still a very large volume of cooling water.

The drawbacks of the high energy required for the relatively low water flux and the very low single-module water recovery outweigh the advantages of very high salt rejection, chemical stability, ability to operate under very low pressure, and high life expectancy. Membranes designed specifically for MD applications may improve the outlook of MD for brine desalination. Additionally, if waste heat can be used to heat and reheat the feed water (feed water loses heat during the process and requires reheating to achieve high water recovery), MD would become a more viable alternative. For this reason, MD would not be considered as one of the process combinations in the treatment alternatives for this study. As a result, no capital and operating cost estimates were produced for MD.

4.3.2 Forward Osmosis

All FO experiments were carried out using flat-sheet cellulose triacetate membranes. The effect of feed water concentration and composition on the membrane flux rate for Brines A and B are illustrated in figure 4.11. The draw solution was maintained at a constant concentration of 50,000±3,000 mg/L using the pilot-scale RO system. The osmotic pressure difference between the feed stream and the DS stream was large enough to maintain a driving force capable of producing a relatively high flux. Water flux in the Brine A was found to range from 3 to 12 L/m^2-hr (1.8 to 7.1 gfd), which was higher than that for Brine B at 3 to 9 L/m^2-hr (1.8 to 5.3 gfd).

Water flux declined continuously in both experiments. As water crossed the membrane, the TDS concentration (and, thus, osmotic pressure) of the feed stream increased; the DS concentration was maintained at constant concentration by the RO system. Therefore, the osmotic pressure gradient across the membrane was reduced as Brines A and B were concentrated. The initial flux decline is not due to fouling or scaling (as commonly seen in pressure-driven fouling results) but mainly due to decreased driving force.

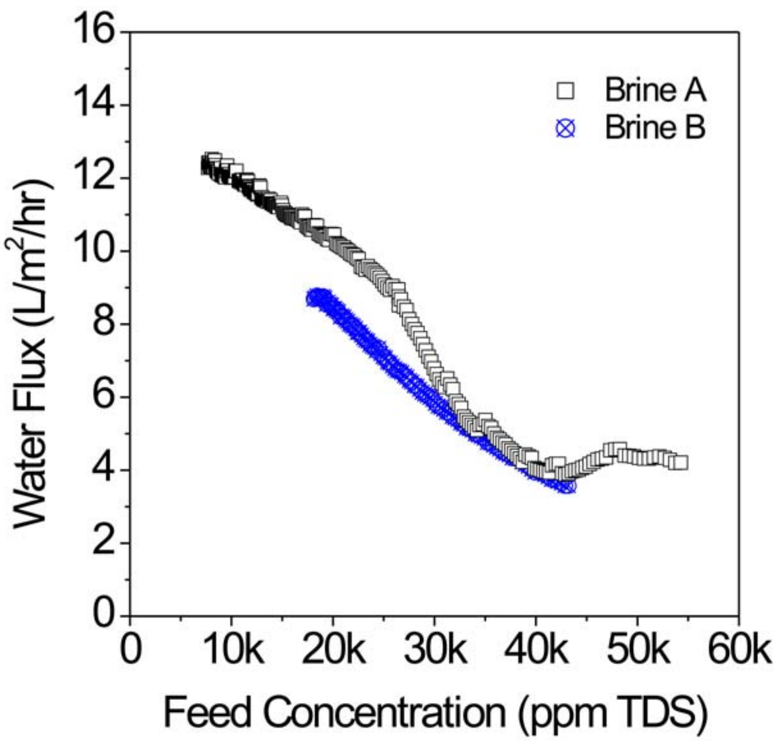


Figure 4.11 Water flux versus feed concentration for FO of Brine A and Brine B.

Scale was formed and deposited on the surface of the membrane during the experiments. Results for Brine A in figure 4.11 indicate a sudden change in the rate of flux decline at a feed concentration of approximately 25,000 parts per minute TDS. Further changes in the rate of flux decline/increase were repeatedly observed and are not well understood. It is believed that, due to scale inhibitor residues in Brine B, no noticeable change in flux rate was observed.

The water recoveries for Brines A and B were 86 and 51 percent, respectively. Brine B had a greater osmotic pressure than Brine A and, consequently, water flux declined to an unacceptable level at a relatively low water recovery.

Modeling was performed to estimate the size, specific energy and power consumptions, and other operating cost components required for producing 1 mgd of drinking water from brackish water RO brine. These parameters were estimated for a total water recovery of 75 percent in the FO subsystem. A membrane array was designed with ROSA (ROSA, Dow Filmtec, Midland, Michigan) that produced 1 mgd of purified water. Figure 4.12 shows a conceptual process schematic of an FO system.

In FO, as with other membrane processes, the amount of recovered water depends largely on the surface area of the membrane. Therefore, the model was iterated to obtain the most efficient amount of membrane area that would produce

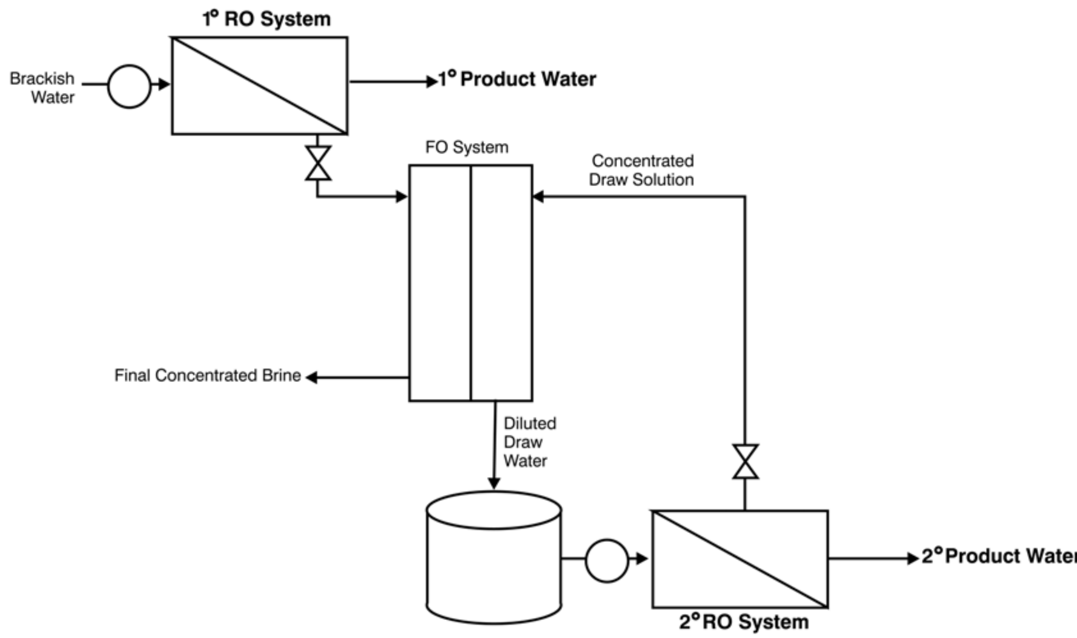


Figure 4.12 Conceptual process schematic for FO system.

approximately 75 percent water recovery from either Brine A or B. It was found that the FO membranes required the majority of the membrane area of the system because flux is lower in FO than in RO.

Ultimately, for the FO model, 352 FO vessels (704 elements) were arranged in parallel. Each element is 50 m^2 in membrane surface area. The power required to pump 1.34 mgd of water into the FO system using 14 1.5-horsepower pumps is 8.7 kilowatts (kW). The specific energy requirement for the FO system is, therefore, calculated by dividing 8.7 kW by 1 mgd water production rate, to give approximately 0.21 kWh/1,000 gal. The results from the ROSA model indicate that 10 kWh/1,000 gallon are required to produce 1 mgd using 150 RO membrane elements. The specific energy of the combined system is 10.4 kWh/1,000 gallon, with the FO system requiring approximately 2 percent of the energy. At a power rate of $0.12 per kWh, this would translate into a treatment cost of $1.25 per 1,000 gallon ($407 per acre-foot). The majority of the specific energy is required by the RO membranes due to the high-pressure operation of the DS separation system. It is also important to note that further savings can be realized by incorporating an energy recovery device into the RO system.

4.4 Desktop Modeling

4.4.1 Residual Recovery (SAL-PROC™)

The salinity variations between Brine A (TDS of 6,000 mg/L) and Brine C (22,000 mg/L) necessitated different approaches for brine treatment. The desktop modeling of these two feed compositions generated two treatment options for each of the 1-mgd brine streams—namely, SAL-PROC#1 for Brine A, and SAL-PROC#2 for Brine C. The conceptual flow diagrams of the ZLD options for these brines are shown in figure 4.13. All options start with one or two SAL-PROC^{TM} process steps followed by treatment of the spent water in RO membrane unit(s) and then the reduction of the RO brine in a brine concentrator. Options for commercial disposal of BC brine are provided to give an indicative base for economic evaluation of both the ZLD and partial treatment/disposal options. The BC reject is essentially comprised of sodium sulfate, calcium sulfate, and sodium chloride salts and, thus, amenable to further treatment for the separation and recovery of salt byproducts. However, in view of the relatively small volumes involved and the low market value of the sodium sulfate salt ($75 to $110 per ton), the production of noncommercial mixed salts for achieving ZLD or commercial disposal was considered more economical for the purpose of this study.

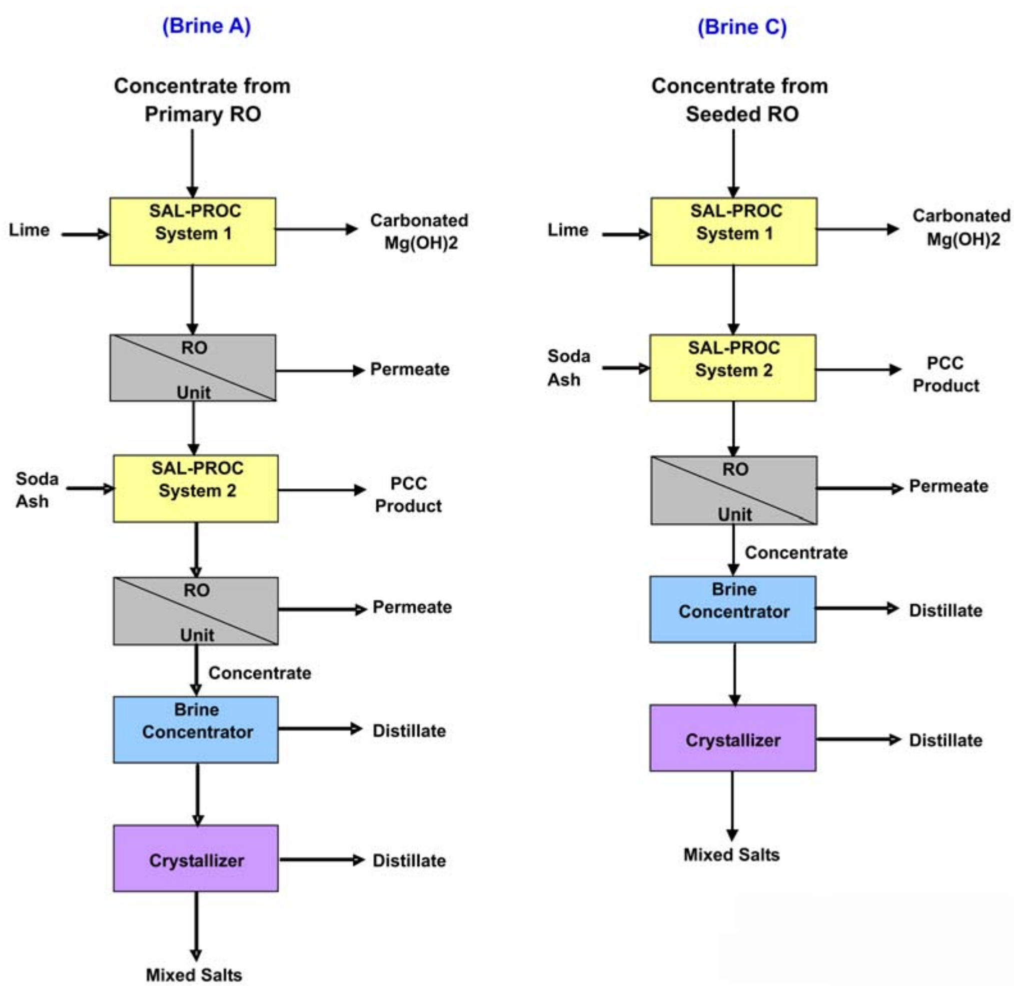


Figure 4.13 SAL-PROC modeling for 1-mgd ZLD treatment of Brine A and Brine C.

4.4.1.1 SAL-PROC#1 for Brine A

This component is comprised of multiple substeps including chemical dosing, additional RO units, and brine concentration. The brine stream is taken through a first salt recovery system whereby magnesium hydroxide ($Mg(OH)_2$) is harvested. The $Mg(OH)_2$ product, being a flocculating agent, will also remove most of the dissolved Si and other metal ions from the brine that are detrimental to functioning of the RO membranes. Subsequently, the supernatant from the $Mg(OH)_2$ recovery step passes through an RO process at 60-percent recovery for further concentration, and the brine will undergo a second salt recovery system whereby precipitated calcium carbonate (PCC) slurry is harvested. The PCC slurry would then be dewatered and processed via a dryer or sold as filter cake product as commercial-grade filler for paper and polyvinylchloride manufacturing. It is estimated that 14 tons per day of PCC would be produced from treatment of 1 mgd of brine from the primary RO. The supernatant from the PCC recovery system is treated by a second RO process at 60-percent recovery, and the resulting brine is then subjected to volume reduction to 2.3 percent of the original brine feed volume using a brine concentrator. The total recovery from the two RO units would be approximately 84 percent of the inflow (1 mgd).

4.4.1.2 SAL-PROC#2 for Brine C

This treatment process is similar to SAL-PROC#1 for Brine A except the supernatant from the $Mg(OH)_2$ recovery step is fed directly to the PCC recovery system, bypassing the need for an additional RO unit.

The final stage in the SAL-PROC™ system is concentrate minimization. This is achieved through using a thermo-mechanical crystallizer to produce dry-mixed salts for landfill disposal.

A summary listing of the cost estimates and revenue generation potential from salt byproducts is given in table 4.6. The cost estimates are presented without Carollo's modification to illustrate the selection of an optimized treatment based on the 1-mgd brine capacity that was provided to Geo-Processors. Table 4.6 is provided to serve as the basis for the rational elimination of other SAL-PROCTM alternatives to arrive at the total ZLD of Brine C. As the table shows, four alternatives were evaluated by Geo-Processors. The difference between ZLD and partial treatment for feed brine is the omission of a crystallizer following the brine concentration step. In all cases, except for Alternative C, the annual cost of the operation (including the benefit of revenue generated) would result in a net outlay of cash. In the case of Alternative C, it is estimated that about \$0.7 million could be generated per year.

Table 4.6 Conceptual Cost Estimates and Projected Revenue Base for the Recommended Brine Treatment Systems (1 mgd)

Alternative	Annua Revenue[1] ($m)	O&M ($m/yr)	Capital Cost ($m)	Net Annual Cost[2] ($m/yr)
(a) Brine A – ZLD	3.29	2.93	17.60	1.17
(b) Brine A – Partial Treatment and Commercial Disposal	3.29	3.79	14.63	1.17
(c) Brine C – ZLD	5.26	2.10	28.37	-0.67
(d) Brine C – Partial Treatment and Commercial Disposal	5.26	8.17	16.48	4.35

[1] Based on sale of byproduct chemicals.
[2] Sum of O&M and annualized cost at 6 percent over 20 years less annual revenue.

The O&M costs were determined to be higher for options involving partial treatment; this reflects the high disposal cost of reject from brine concentrators. The capital costs were influenced by the range of equipment and the salt load of the brine stream.

4.4.2 Brine Concentration

Water quality data for Brines A and B were analyzed by HPD-Veolia Water Solutions to estimate the budget for a BC/XLZR system for 0.125- to 1-mgd treatment capacity. The installation estimates provided are for mechanical and electrical construction on a customer-supplied foundation. No buildings are allowed for. It is assumed that the brine concentrators are outdoors and the PLC computer interface and motor control center are installed in the same electrical/control building as the RO plant.

For mass balance and cost estimate calculation purposes, the flow out of the BC was set to approximately 4 percent of feed Brine B (6,300 mg/L TDS) flow and 15 percent of feed Brine A (28,000 mg/L TDS) flow. The values for brine output were selected based on the water quality data and anticipated performance of the brine concentrators. The received quotes include cost estimates for a combined BC/XLZR system; however, to price the brine concentrator and crystallizer separately, a 10-percent increase was added to the individual unit price. The BC portion of the system makes up approximately 85 percent of the turnkey costs, power consumption, O&M costs, spare equipment, and the chemical/cleaning costs. In table 4.7, the total capital cost, O&M, and annualized capital is summarized for treatment capacities varying between 0.125 to 1.0 mgd.

Table 4.7 Conceptual Cost Estimates for Brine Concentration Treatment Systems (0.125 to 1 mgd) for BC only

Capacity	Total Capital Cost ($)		O&M Cost ($/yr)		Annual Costs[1] ($/yr)	
	Brine A	Brine B	Brine A	Brine B	Brine A	Brine B
1 mgd	$31,447,238	$31,447,238	$4,689,550	$4,870,350	$7,431,263	$7,612,063
0.5 mgd	$21,713,569	$21,713,569	$2,571,900	$2,675,475	$4,464,988	$4,568,563
0.25 mgd	$15,640,425	$16,472,363	$1,497,200	$1,588,500	$2,860,804	$3,024,636
0.125 mgd	$12,978,225	$12,978,225	$900,100	$973,400	$2,031,601	$2,104,901

[1] Sum of O&M and annualized cost at 6 percent over 20 years.

As indicated by the values in the table, there is very little cost difference (capital or O&M) to treat Brine A or Brine B, indicating that the costs for BC are largely determined by the flow rate rather than the feed water quality.

4.4.3 Crystallizer

Scenarios using a crystallizer to further concentrate the brine streams were used in conjunction with a brine concentrator. For that reason, it is impractical to price the crystallizer as a stand-alone unit. The cost estimates are provided by HPD-Veolia Water Solutions for a combined BC/XLZR system for 0.125- to 1-mgd brine capacity and summarized in table 4.8. The same assumptions in section 4.4.2 are applied. Brine concentration following crystallizer treatment includes a built-in landfill cost of $50 per ton for the remaining solids.

Table 4.8 Conceptual Cost Estimates and Projected Revenue Base for the Recommended Brine Treatment Systems (1 mgd to 0.125 mgd) for BC/XLZR System

Capacity	Total Capital Cost ($)		O&M Cost ($/yr)		Annual Costs[1] ($/yr)	
	Brine A	Brine B	Brine A	Brine B	Brine A	Brine B
1 mgd	$38,171,250	$41,107,500	$5,197,050	$7,046,400	$8,524,994	$10,630,339
0.5 mgd	$26,426,250	$28,383,750	$2,834,400	$3,775,450	$5,138,361	$6,250,075
0.25 mgd	$18,987,750	$21,532,500	$1,624,950	$2,160,050	$3,280,389	$4,037,351
0.125 mgd	$15,660,000	$16,638,750	$963,100	$1,280,750	$2,328,410	$2,731,392

[1] Sum of O&M and annualized cost at 6 percent over 20 years.

4.4.4 Evaporation Ponds

A net evaporation rate of 70 inches per year and a TDS of 263,600 mg/L were assumed for the evaporation pond cost estimation. The assumed water depth for the pond was 18 inches, with a minimum freeboard of 2 feet. Excavation costs of

$5 per cubic yard, land costs of $4,000 per acre, and an installed liner cost of $2 per square foot were assumed for the model. A 20-percent required land contingency is built into the model for sizing the pond. This cost estimate assumes that salts are hauled to landfill disposal following evaporation.

4.5 Process Combination Alternatives

Fourteen process combination alternatives and two base cases are presented below. Projections are based in part on the results of pilot study conducted onsite and on work done by the various subcontractors. Primary brine, secondary RO, EDR, and seeded RO tests were conducted onsite. Primary and secondary brine samples were sent to the University of Nevada, Reno, for FO and MD experiments; the corresponding water quality data for the primary and secondary brine streams were sent to HPD-Veolia Water Solutions for modeling of brine concentration and crystallization unit processes. Comprehensive water quality data for primary brine and seeded RO concentrate were also sent to Geo-Processors for a desktop prefeasibility study. The results of these desktop modeling, bench-scale and pilot testing were used to propose 14 treatment trains that would support the minimization (or elimination) of brine concentrate to the SARI line. Conceptual line diagrams showing the proposed treatment trains are provided in figure 4.14 and discussed in detail below. See Appendix E for detailed cost calculations.

4.5.1 Alternative 1: Primary RO/Chemical Softening/Secondary RO/Brine Concentrator/SARI Discharge

This treatment train includes the following components:

4.5.1.1 Pretreatment

Primary brine (Brine A) is fed to a softening plant where chemicals are dosed to encourage precipitation of sparingly soluble salts. Hydrated lime is dosed to raise the pH to a level where calcium, magnesium, or both settles out as their respective salts, calcium carbonate and magnesium hydroxide. When needed, soda ash is added to provide sufficient alkalinity to encourage the removal of the more dominant calcium ions. An anionic polymer is added to aid in densification and settling of the flocs. This is especially essential when a significant fraction of the floc is comprised of the lighter magnesium hydroxide flocs. Silica adsorbs to calcium carbonate and magnesium carbonate flocs and is removed through co-precipitation.

The solids settle out leaving a clarified water in which the concentration of scaling precursors in the water has been reduced by as much as 90 percent. At 2.0 NTUs, the turbidity of the clarified water is too high for membrane processes,

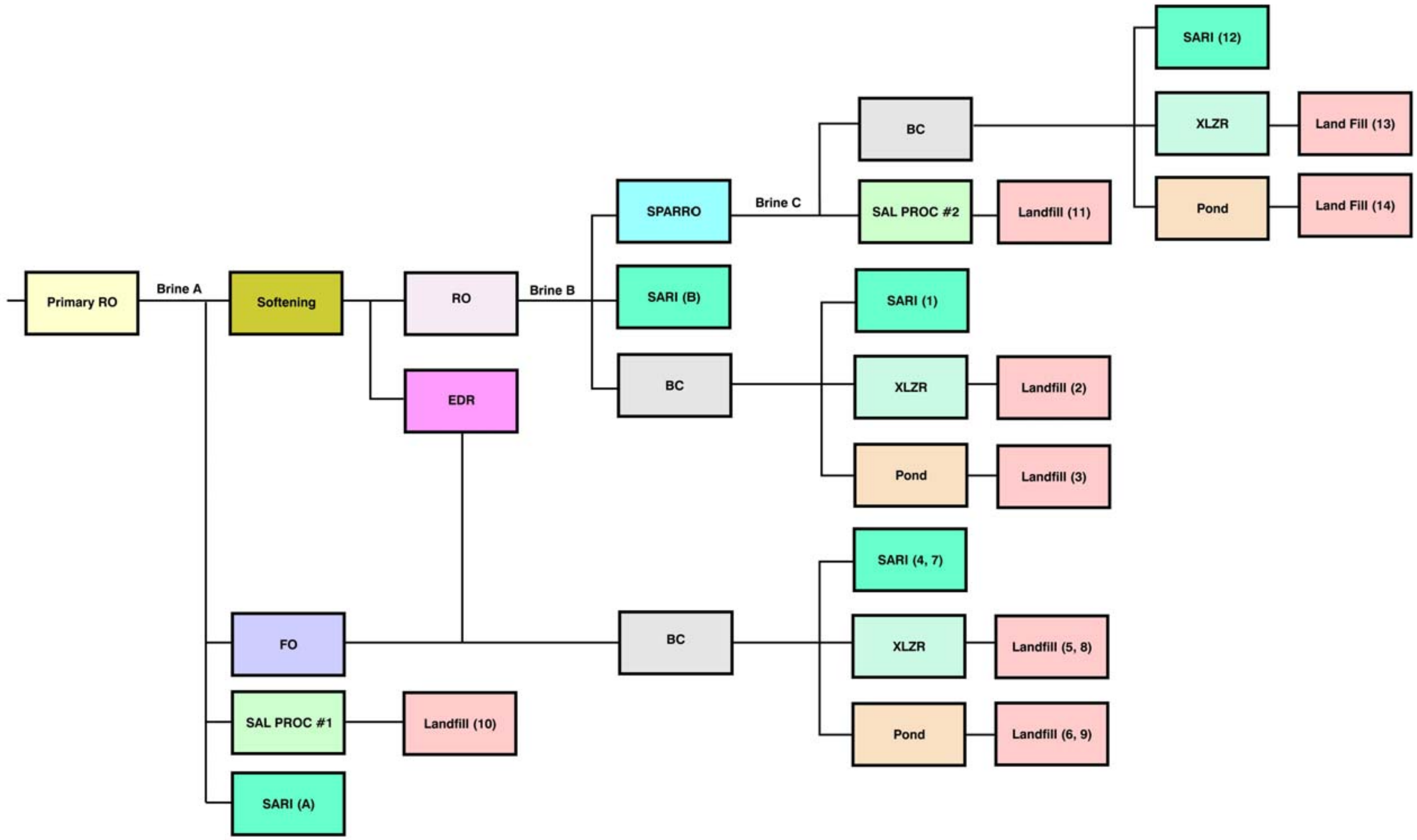


Figure 4.14 Treatment alternatives for Brines A, B, and C.

especially the more sensitive RO membranes. The clarified water is fed to a multimedia filter where appropriate turbidity and SDI_{15} values are achieved.

The water is further conditioned by dosing acid and an appropriate antiscalant. The pH of the water leaving the filter is approximately 9.0. Acid is added to lower the pH to between 7.0 and 8.0. Although the chemical softening has lowered the concentration of scaling precursors, as the water concentrates along the succeeding RO membranes, the remaining salts exceed their solubility limit and can settle out of solution causing irreparable scaling of the membranes. The salts of concern are SiO_2, barium sulfate ($BaSO_4$), $CaCO_3$, $Mg(OH)_2$, and calcium fluoride (CaF_2). By controlling pH, the salts are maintained in soluble form and removed in the concentrate stream. Antiscalant is used to further inhibit scale formation.

4.5.1.2 Secondary RO

The conditioned stream of water is fed to a second stage of RO membranes, which operate at between 75- and 85-percent recovery. Assuming that the primary RO continues to recover 70 percent of brackish water, the combined recovery for both membrane processes now ranges from 92.5 to 95.5 percent. This level of recovery is possible since the pretreatment will have reduced Ca, Mg, and Si levels to below that of the influent ground water.

The permeate stream from the secondary RO has a maximum TDS of 300 mg/L and will combine with the 100-mg/L TDS permeate stream from the primary RO. The concentrate stream (Brine B) has a TDS ranging from 18,000 to 30,000 mg/L, depending on level of recovery. At the upper limit of RO recovery, TDS is nearly as high as ambient Pacific Ocean water (approximately 34,000 mg/L).

4.5.1.3 Brine Concentrator

Since between 75 to 85 percent of the primary brine have been recovered for use, the secondary brine stream is now significantly reduced. This stream is augmented with the solid byproduct of chemical softening, and both are fed to a thermal vapor compression brine concentration unit. Here, the volume is reduced by an additional 90 percent. If the primary brine stream was 1 mgd, the secondary brine will be 0.15 mgd (at maximum recovery); and the effluent from the concentrator will be only 0.015 mgd. These estimates do not include the water lost as backwash waste. Resulting TDS of the effluent can be as high as 250,000 mg/L.

4.5.1.4 SARI Discharge

The effluent from the brine concentrator is discharged into the SARI line along with backwash waste. The water discharged to the SARI is now reduced by 97 percent over the volume discharged without additional treatment.

4.5.2 Alternative 2: Primary RO/Chemical Softening/Secondary RO/Brine Concentrator/Crystallizer/Solids Disposal

This treatment train is similar to the Alternative 1, which includes primary RO/Chemical Softening/Secondary RO/Brine Concentrator/SARI Discharge, except the discharge to SARI line is replaced by Crystallizer/Solids Disposal.

4.5.2.1 Crystallizer

The effluent from the brine concentrator is not discharged to the SARI line but is instead sent to a crystallizer for further processing. With a TDS of 250,000 mg/L, the water entering the crystallizer is supersaturated with respect to all component salts. Therefore, they are relatively easily extracted. The process achieves a ZLD outcome, and the byproduct is commercial grade salts, which can be sold to offset operating costs.

4.5.2.2 Solids Disposal

The solids are trucked from the site. The solids may either be sold to use in some manufacturing or agricultural process or landfilled. Under the Comprehensive Environmental Response, Compensation, and Liability Act, the solids are classified as type I or II and can be landfilled in municipal facilities. In the current cost model, a landfill disposal cost of $50 per ton is assumed for trucking off remaining salts.

4.5.3 Alternative 3: Primary RO/Chemical Softening/Secondary RO/Brine Concentrator/Evaporation Pond/Solids Disposal

This option is similar to Alternative 1, which includes Primary RO/Chemical Softening/Secondary RO/Brine Concentrator/SARI Discharge, except the discharge to SARI is replaced by Evaporation Pond.

4.5.3.1 Evaporation Pond

No mechanical processing of secondary brine is done, and concentrate disposal is achieved solely through the use of evaporation ponds. Ponds are sized to accommodate full secondary RO concentrate flow, as much as 7.5 percent of the primary brine flow.

4.5.3.2 Solids Disposal

The solids are trucked from site, as discussed in Alternative 2.

4.5.4 Alternative 4: Primary RO/Chemical Softening/EDR/Brine Concentrator/SARI Discharge

Similar to Alternative 1, which includes Primary RO/Chemical Softening/Secondary RO/Brine Concentrator/SARI Discharge, except for the secondary RO is replaced by EDR.

4.5.4.1 Electrodialysis Reversal

EDR membranes are used in place of RO membranes to achieve recovery ahead of brine concentration and SARI discharge. Since the EDR membranes are not as sensitive to solids and silica as RO membranes, it is possible to do less chemical treatment of the primary brine, and the EDR feed will comprise of combined streams of softened and raw brine. The EDR will be operated at the same recovery as the RO and, therefore, hydraulic loading for the downstream processes will be unchanged.

4.5.5 Alternative 5: Primary RO/Chemical Softening/EDR/Brine Concentrator/Crystallizer/Solids Disposal

Similar to Alternative 2, which includes Primary RO/Chemical Softening/Secondary RO/Brine Concentrator/Crystallizer/Solids Disposal, except that EDR is used in place of RO for recovery ahead of brine concentration.

4.5.6 Alternative 6: Primary RO/Chemical Softening/EDR/Brine Concentrator/Evaporation Pond/Solids Disposal

Similar to Alternative 3, which includes Primary RO/Chemical Softening/Secondary RO/Brine Concentrator/Evaporation Pond/Solids Disposal, except that EDR is used in place of RO for recovery ahead of thermal brine concentration and evaporation ponds.

4.5.7 Alternative 7: Primary RO/Forward Osmosis/Brine Concentrator/SARI Discharge

This alternative is another variant of Alternative 1, which includes Primary RO/Chemical Softening/Secondary RO/Brine Concentrator/SARI Discharge. The difference is that there is no chemical pretreatment and that all the softening is provided through an FO unit.

4.5.7.1 Forward Osmosis

The brine stream from the primary RO plants is used as feed to the FO plant. This stream varies in concentration according to well usage. The FO unit is sized for the maximum primary brine TDS of 7,500 mg/L. A 50,000-mg/L NaCl solution is used as the draw solution and is kept constant through the use of an ancillary RO plant. The FO recovers as much as 86 percent of the primary brine bringing overall recovery of brackish well water to 95.8 percent. This is similar to the 95.5-percent maximum achieved from using RO as secondary membranes. The advantage of the FO is that it achieves similar recovery as secondary RO without using chemicals. In this process combination, FO does not require chemical softening pretreatment. The diluted DS from FO will need to undergo an additional secondary RO treatment to recover the water from primary RO brine.

4.5.7.2 Brine Concentrator

Fourteen percent of the primary RO brine cannot be extracted by the draw solution and is sent to a brine concentrator for volume reduction. The concentrator recovers 90 percent of the water, and the final effluent is similar in volume to that from Alternative 1.

4.5.7.3 SARI Discharge

Discharge volume to the SARI line is now substantially reduced. The volume of brine is less than 2 percent of the initial RO brine.

4.5.8 Alternative 8: Primary RO/Forward Osmosis/Brine Concentrator/Crystallizer/Solids Disposal

Similar to Alternative 7, which includes Primary RO/Forward Osmosis/Brine Concentrator/SARI Discharge, except that the effluent from the brine concentrator undergoes further processing to achieve a ZLD outcome.

4.5.8.1 Crystallizer

Concentrator effluent is processed to produce commercial grade pure salts and a variety of mixed salts which can be used as a revenue-generating byproduct.

4.5.8.2 Solids Disposal

The resulting salts are trucked from the site. Pure salts are used in manufacturing processes, and mixed salts are landfilled.

4.5.9 Alternative 9: Primary RO/Forward Osmosis/Brine Concentrator/Evaporation Pond

Similar to Alternative 7, which includes Primary RO/Forward Osmosis/Brine Concentrator/SARI Discharge, except that the final brine stream is discharged to evaporation ponds and landfilled.

4.5.10 Alternative 10: Primary RO/LOW TDS SAL-PROC #1/Solids Disposal

This treatment option includes the following:

4.5.10.1 Primary RO

Brackish ground water is concentrated in the primary RO plants. The plants are operated at 70-percent recovery. The remaining water is about 30 percent of the inflow to the plant and is supersaturated with respect to numerous salts.

4.5.10.2 SAL-PROC#1

One mgd of Brine A will undergo mixing with lime to harvest 5.53 tons of carbonated ($Mg(OH)_2$). Subsequently, the supernatant from the $Mg(OH)_2$ recovery step passes through an RO process at 60-percent recovery for further concentration; and the brine (0.6 mgd) will undergo a second salt recovery system to harvest PCC slurry. The PCC slurry would then be dewatered and processed via a dryer or sold as filter cake product as commercial-grade filler for paper and PVC manufacturing. It is estimated that 14 tons per day of PCC would be produced from treatment of 1 mgd of brine from the primary RO. The supernatant from the PCC recovery system is treated by a second RO process at 60-percent recovery, and the resulting brine (0.16 mgd) is then subjected to volume reduction to 2.3 percent of the original brine feed volume using a brine concentrator. The total recovery from the two RO units would be approximately 84 percent of the inflow (1 mgd).

The final stage in the SAL-PROC system is concentrate minimization. This is achieved through using a thermo-mechanical crystallizer to produce dry-mixed salts for landfill disposal.

4.5.10.3 Solids Disposal

An estimated 9,000 tons per 1 mgd of mixed salts will be produced, and the landfill disposal will be calculated at $50 per ton.

4.5.11 Alternative 11: Primary RO/Chemical Softening/Secondary RO/Seeded RO/SAL-PROC#2/Solids Disposal/SARI Discharge

This treatment option is similar to Alternative 10, which includes Primary RO/Low TDS SAL-PROC #1/Solids Disposal, except for the following:

4.5.11.1 Pretreatment

Primary brine (Brine A) is fed to a softening plant where chemicals are dosed to encourage precipitation of sparingly soluble salts. Hydrated lime is dosed to raise the pH to a level where calcium, magnesium, or both settle out as their respective salts, calcium carbonate, and magnesium hydroxide. When needed, soda ash is added to provide sufficient alkalinity to encourage the removal of the more dominant calcium ions. An anionic polymer is added to aid in densification and settling of the flocs. This is especially essential when a significant fraction of the floc is comprised of the lighter magnesium hydroxide flocs. Silica adsorbs to calcium carbonate and magnesium carbonate flocs and is removed through co-precipitation.

The solids settle out leaving clarified water in which the concentration of scaling precursors water has been reduced by as much as 90 percent. At 2.0 NTUs, the turbidity of the clarified water is too high for membrane processes, especially the more sensitive RO membranes. The clarified water is fed to a multimedia filter where appropriate turbidity and SDI_{15} values are achieved.

The water is further conditioned by dosing acid and an appropriate antiscalant. The pH of the water leaving the filter is approximately 9.0. Acid is added to lower the pH to between 7.0 and 8.0. Although the chemical softening has lowered the concentration of scaling precursors, as the water concentrates along the succeeding RO membranes, the remaining salts exceed their solubility limit and can settle out of the solution, causing irreparable scaling of the membranes. The salts of concern are SiO_2, $BaSO_4$, $CaCO_3$, $Mg(OH)_2$ and CaF_2. By controlling pH, the salts are maintained in soluble form and removed in the concentrate stream. Antiscalant is used to further inhibit scale formation.

4.5.11.2 Secondary RO

The conditioned stream of water is fed to the second stage of RO membranes which operate at between 75- and 85-percent recovery. Assuming that the primary RO continues to recover 70 percent of brackish water, the combined recovery for both membrane processes now ranges from 92.5 to 95.5 percent. This level of recovery is possible since the pretreatment will have reduced Ca, Mg, and Si levels to below that of the influent ground water.

The permeate stream from the secondary RO has a maximum TDS of 300 mg/L and would combine with the 100-mg/L TDS permeate stream from the primary RO. The concentrate stream (Brine B) has a TDS ranging from 18,000 to 30,000 mg/L, depending on the level of recovery. At the upper limit of RO recovery, TDS is nearly as high as ambient Pacific Ocean water (approximately 34,000 mg/L).

4.5.11.3 Seeded RO

The brine from the secondary RO system is not sent immediately for brine concentration and minimization. Instead, it is subjected to a tertiary membrane process—seeded RO system. Concentrations of calcium and sulfate ions in the secondary brine are 1,453 and 2,202 mg/L, respectively. These ion concentrations in the brine exceed the saturation index of calcium sulfate and will precipitate on the recirculating gypsum seeds in the seeded RO membranes. The seeded RO system operates at an optimum recovery of 75 percent. Therefore, the concentrate from this tertiary membrane (Brine C) is now approximately 6.25 percent of the original primary RO brine flow.

4.5.11.4 SAL-PROC#2

Brine C from the seeded RO unit is channeled toward a SAL-PROC™ treatment to maximize water recovery. This stage involves the same process steps as SAL-

PROC#1 (Alternative 12) with the exception that there is one less RO unit in the process (i.e., the supernatant from the $Mg(OH)_2$ settling pond is immediately dosed with soda ash to harvest PCC and then sent to a single RO unit). This RO operates at 60-percent recovery, and the entire treatment option recovers 97.5 percent of water from the influent primary brine. This brings overall plant recovery up to 99 percent.

The spent water, now about 2.5 percent of the original inflow to the treatment train, (which would be devoid of deleterious dissolved elements) would be further concentrated in a brine concentrator to reduce the volume by 9/10 to just 0.25 percent of original (Brine A) feed volume (0.0025 mgd concentrate per 1 mgd Brine A).

The final stage in the SAL-PROC™ system is brine precipitation. This is achieved through using a thermo-mechanical crystallizer to produce dry-mixed salts for landfill disposal.

4.5.11.5 Solids Disposal

An estimated 2,060 tons per annum of mixed salts would be produced per 1 mgd of primary brine, requiring 1 acre of established landfill space.

4.5.12 Alternative 12: Primary RO/Chemical Softening/Secondary RO/Seeded RO/Brine Concentrator/SARI Disposal

This treatment option is similar to Alternative 11, which includes Primary RO/Chemical Softening/Secondary RO/SPARRO/SAL-PROC#2/Solids Disposal/SARI Discharge, except that Brine C from the SPARRO process is sent directly to the brine concentrator.

4.5.12.1 Brine Concentrator

Since no additional chemical and RO softening is done upstream of the brine concentrator, all reject from the secondary RO enters the brine concentrator. The feed volume to the concentrator is 0.0625 mgd per 1 mgd of primary RO brine (2.5 times the concentrator feed in Alternative 13). This brine is reduced by 90 percent, and the effluent is discharged via the SARI line.

4.5.12.1 SARI Disposal

Volume of saline waste to the SARI line is 0.00625 mgd per 1 mgd of primary RO brine treated.

4.5.13 Alternative 13: Primary RO/Chemical Softening/Secondary RO/Seeded RO/Brine Concentrator/Crystallizer/Solids Disposal

This treatment option is similar to Alternative 12, which includes Primary RO/Chemical Softening/Secondary RO/Seeded RO/Brine Concentrator/SARI Disposal, except the discharge to SARI is replaced by Crystallizer/Solids Disposal.

4.5.13.1 Crystallizer

The concentrator effluent is sent to a thermo-mechanical crystallizer to produce dry-mixed salts for landfill disposal.

4.5.13.2 Solids Disposal

Although solids may be purified more to produce pure salable salts, this is not considered in this treatment option. Instead, the salts are landfilled.

4.5.14 Alternative 14: Primary RO/Chemical Softening/Secondary RO/Seeded RO/Brine Concentrator/Evaporation Pond/Solids Disposal

Similar to Alternative 13, which includes Primary RO/Chemical Softening/Secondary RO/Seeded RO/Brine Concentrator/Crystallizer/Solids Disposal, except the Crystallizer/Solids Disposal option is replaced by Evaporation Pond.

4.5.14.1 Evaporation Pond

The concentrate from the brine concentrator is sent to a concentrator pond for solar-powered crystallization. A mixed salt is produced which is periodically removed and trucked off to a landfill.

Recovery for this treatment option is 94 percent, and overall process recovery is 98 percent.

4.5.15 Base Case A: Primary RO/SARI Disposal

This considers the cost for disposing primary RO brine (Brine A) from the Menifee desalter directly to the SARI line. A flow capacity of 1 mgd is assumed. A total capital cost of $8,100,000 is required, and an additional disposal cost of $365,000 is needed per year. Only 70-percent water recovery will be obtained by this treatment scheme.

4.5.16 Base Case B: Primary RO/Softening/Secondary RO/SARI Disposal

One mgd of Brine A is softened to reduce hardness of the water so secondary RO processing would be possible. The overall process recovery achieved for an additional RO train would increase from 70 to 93 percent, leaving 0.3 mgd of untreated brine (Brine B).

4.6 Cost Summary

As seen from the previous sections in this chapter, it is apparent that the various treatment processes discussed can be divided into three major categories—pilot testing, bench-scale experimenting, and computer simulations. Brine A from the primary Menifee desalter underwent a pilot-scale softening unit before secondary treatment by either RO or EDR. Water analysis from the secondary RO brine stream and EDR brine is similar in composition and is designated as Brine B. This secondary brine stream is introduced as feed to SPARRO, and the concentrate stream is termed Brine C.

Samples of Brine A and Brine B were sent to the University of Nevada, Reno, for FO and MD bench-scale experiments to test the viability of these two processes for brine minimization. Membrane distillation of both Brine A and Brine B required extremely high-energy demands for low water flux (described in section 4.3.1) and was removed from further evaluation. Forward osmosis, a relatively new technology, was able to recover 86 percent of Brine A and 51 percent of Brine B. The only drawback to this process is the absence of a commercial membrane especially suited to this technology. To achieve the same water recovery level as RO using cellulose triacetate membranes, a higher membrane surface area would be required for FO. Recycling the brine feed solution for FO also meant that the membrane areas are susceptible to scaling, and more experiments would need to be conducted to test antiscalant effects in FO. The University of Nevada, Reno, then sized the process to treat a 1-mgd brine influent and computed the energy requirements and equipment sizing for such a system. The resulting information was incorporated into Carollo's project cost estimates. Although it has been shown that FO has the capability as an alternative method for water recovery, the process is still at its infancy when compared to RO and EDR.

Water quality data compiled for Brines A, B, and C were sent to companies that specialize in brine minimization technologies such as SAL-PROC™, brine concentration, and crystallization. Based on the three brine compositions, equipment and process parameters were modeled for the specific water compositions at various flow capacities (0.25 to 1 mgd). The vendor price quotes were modified to reflect Carollo's project cost estimates. Modeled recoveries and anticipated amounts of solids generated were fed into Carollo's in-house model for evaporation pond sizing, in addition to solid disposal/landfill costs estimated from prior projects.

Table 4.9 summarizes the overall costs for all 14 treatment alternatives. Costs do not include the primary desalting step and are for brine treatment and disposal only. Potential revenues generated from sale of recovered water and/or salt byproducts are not included in the total costs. The purpose for table 4.9 is to account for capital and annual O&M costs only. Two base cases are included for comparison. The first base case assumes that primary RO brine (Brine A) stream does not undergo further treatment and is discharged into the SARI line at 1-mgd capacity. The total capital cost in this case would be $8,100,000; and the annualized capital cost would be $1,071,195. The second base case considers softening and secondary RO treatment of the primary brine and discharges the remaining 0.3 mgd of secondary RO brine (Brine B) stream into the SARI line (refer to Base Case B in table 4.9). For this study, it is assumed that the excess SARI capacity is not sold when calculating the costs for brine disposal to the SARI line.

From table 4.9, it is apparent that the most expensive treatment alternatives for total cost per annum are SAL-PROC™ processes (Options 10 and 11). SAL-PROC #1, which treats Brine A only at 1 mgd, can reach upward of $8.7 million per year. If Brine A were softened prior to secondary RO and seeded RO treatment, a smaller volume (0.15 mgd of Brine C) can be processed by SAL-PROC#2, and the overall cost will lessen by $2 million.

The second-most expensive alternatives are process combinations (Options 7 to 9) that utilize FO technology ($5.9 to $6.3 million per year). It is challenging to price this technology due to the absence of pilot-scale demonstrations and suitable commercial membranes to make this treatment scheme more economically enticing. The high costs associated with using FO for brine treatment are attributed to its high energy demand and the amount of membrane modules required to achieve approximately 70- to 75-percent recovery.

Treatment alternatives that consider RO and EDR interchangeably (Options 1 to 6), in addition to brine concentrators, are between $5.5 to $5.9 million per year. Brine output generated by concentrators can be directed to a crystallizer or evaporation pond, where the remaining solids will be disposed to landfill. Bypassing thermal evaporation and discharging directly to SARI will lower the treatment costs by $300,000 to $400,000 per year. If SARI disposal is omitted to approach a true ZLD process, Option 6 will be the cheaper alternative, owing to its slightly lower O&M cost between EDR and RO. In regions where there is

Table 4.9 Cost Comparison of Treatment and Disposal Alternatives

	Process Combinations		Total Capital Cost ($)	Annualized Capital Cost ($/annum)	O&M ($/annum)	Total Cost ($/annum)
A	SARI	=	**$8,100,000**			**$1,071,195**
B	Softening + 2°RO + SARI	=	**$14,745,375**	**$1,285,569**	**$978,165**	**$2,263,734**
1	Softening + 2°RO + BC + SARI	=	**$32,085,114**	**$2,798,326**	**$2,741,738**	**$5,540,064**
2	Softening + 2°RO + BC + XLZR-Disposal	=	**$29,151,685**	**$2,541,577**	**$3,413,113**	**$5,954,690**
3	Softening + 2°RO + BC + Evap Pond	=	**$25,563,491**	**$2,229,742**	**$3,638,828**	**$5,868,570**
4	Softening + EDR + BC + SARI	=	**$32,558,033**	**$2,839,558**	**$2,844,123**	**$5,683,681**
5	Softening + EDR + BC + XLZR-Disposal	=	**$29,080,162**	**$2,535,341**	**$3,410,019**	**$5,945,360**
6	Softening + EDR + BC + Evap Pond-Disposal	=	**$25,773,347**	**$2,248,038**	**$3,591,698**	**$5,839,736**
7	FO + BC + SARI	=	**$33,760,072**	**$2,944,357**	**$2,983,877**	**$5,928,234**
8	FO + BC + XLZR-Disposal	=	**$30,826,643**	**$2,687,607**	**$3,655,252**	**$6,342,859**
9	FO + BC + Evap Pond-Disposal	=	**$27,238,449**	**$2,375,772**	**$3,880,967**	**$6,256,740**
10	SAL-PROC#1-Disposal	=	**$25,474,179**	**$2,220,955**	**$6,551,732**	**$8,772,687**
11	Softening + 2°RO + Seeded RO + SAL-PROC#2-Disposal	=	**$27,826,445**	**$2,430,876**	**$4,343,651**	**$6,774,528**
12	Softening + 2°RO + Seeded RO + BC + SARI	=	**$35,279,829**	**$3,081,696**	**$2,314,582**	**$5,396,278**
13	Softening + 2°RO + Seeded RO + BC + XLZR-Disposal	=	**$30,621,940**	**$2,674,600**	**$2,669,835**	**$5,344,435**
14	Softening + 2°RO + Seeded RO + BC + Evap Pond-Disposal	=	**$27,969,018**	**$2,444,306**	**$2,763,125**	**$5,207,431**

[1] Total cost per annum = annualized capital cost + O&M cost.
[2] Costs do not include primary desalting step.

insufficient land to build an evaporation pond (up to a maximum of 12 acres), brine concentration followed by crystallizers would be the only economical choice as a ZLD alternative.

It is interesting to note that in Options 12 to 14, using seeded RO to recover water from Brine B results in the lowest total costs among the 14 (plus 2 base cases) alternatives considered. Although current pilot tests have demonstrated that this technology is capable of recovering secondary RO brine at about 60 to 65 percent, there have not been long-term (over 1 year) studies of the robustness of the membranes' longevity against the abrasiveness of the seed gypsum crystals scouring the membrane surface. The membranes used for the seeded RO pilot testing for Brine B are tubular NF membranes, as there are currently no commercially available RO membranes dedicated for seeded RO usage.

Figure 4.15 summarizes the total capital cost for each process alternative. A depiction of the O&M, annualized capital, and total annual costs is presented in figure 4.16.

4.7 Conclusions

It is projected that the volume of brine discharge will increase when EMWD builds a third RO treatment facility in the future, and brine disposal to the SARI line will result in a significant increase in the annual operating cost. Due to the concentrating effect on components such as heavy metals and arsenic, it is conceivable that EMWD may be required to pay increased costs for ocean discharge in the future. In this report, alternative processes (including ZLD) such as RO, EDR, FO, MD, and seeded RO were studied by either bench-scale experiments or pilot plant testing. Subsequent brine minimization and elimination techniques, which use brine concentrators, crystallizers, evaporation ponds, and SAL-PROC™, were incorporated into the treatment processes, and the process economics were calculated through desktop modeling. A cost analysis model based on individual treatment modules was developed for each process alternative, and a total of 14 process trains in addition to 2 base cases were evaluated. This cost model will not only benefit EMWD but will also benefit other local, State, regional, and national agencies facing brine disposal issues from inland desalination facilities.

The least expensive alternative evaluated was primary RO+softening+EDR+ BC+Evaporation Pond-Disposal (Option 6) at a total annual cost of $5,839,736. It is interesting to note that for the volume and quality of the brine considered in this study, there was only a 5- to 7-percent difference in total annual cost between all treatment trains that result in BC/Evaporation Pond-Disposal and BC/XLZR/ Landfill options. The treatment costs for secondary RO and EDR are nearly

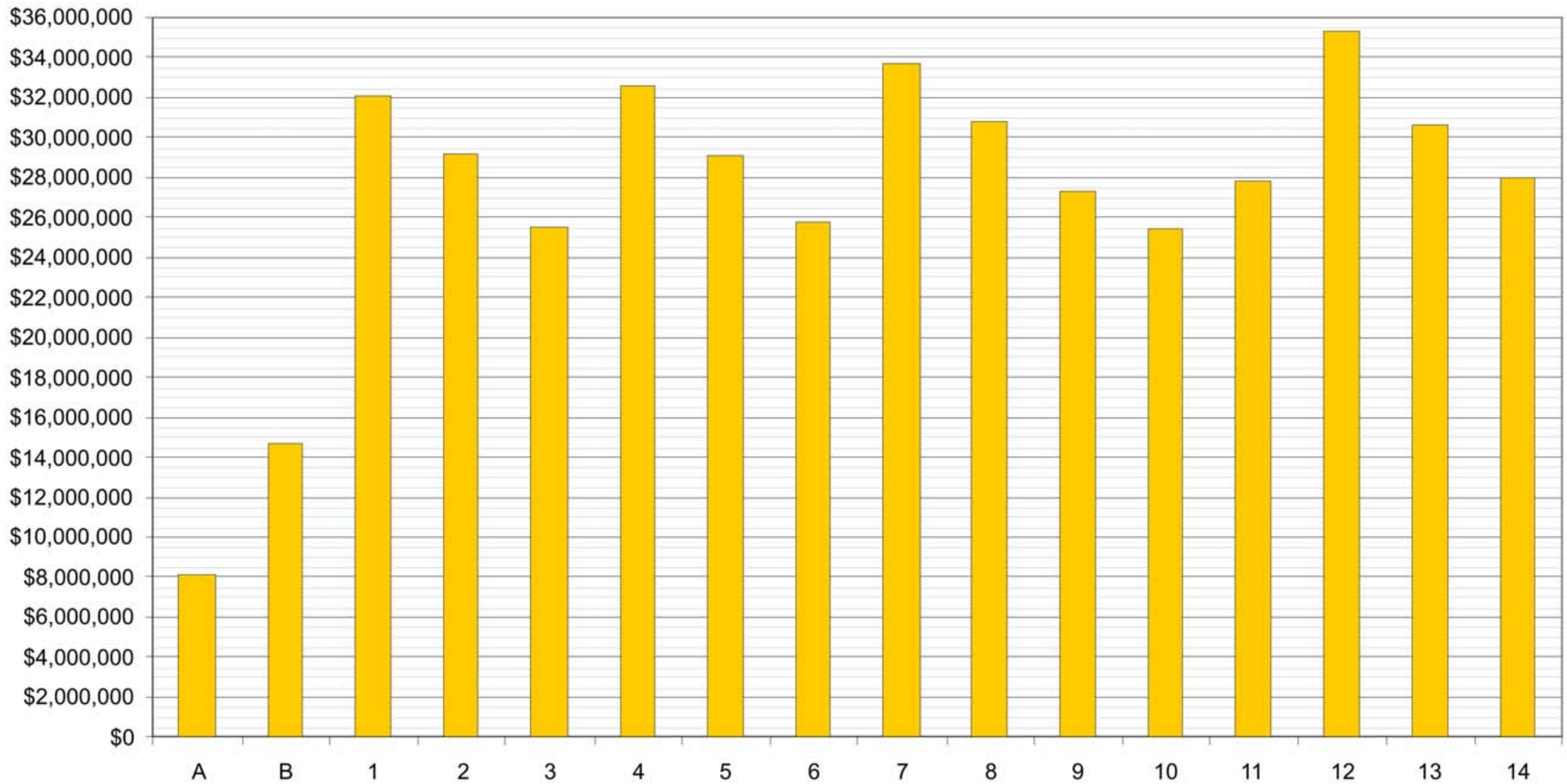


Figure 4.15 Total capital costs for various treatment alternatives.

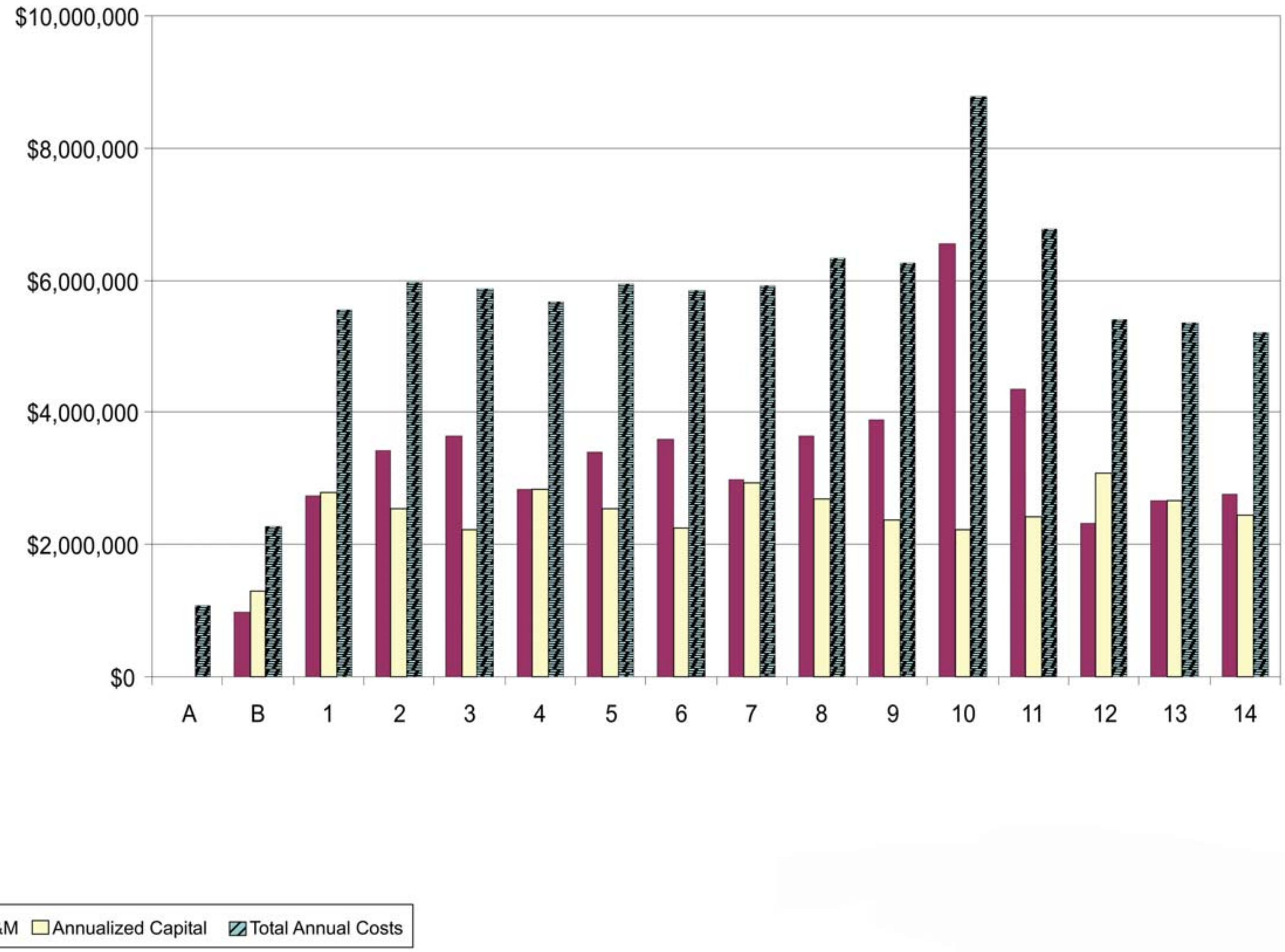


Figure 4.16 Total annual costs for various treatment alternatives.

equivalent, at $5,868,570 and $5,839,736, respectively, for BC/Evaporation Pond-Disposal and at $5,954,690 and $5,945,360 for BC/XLZR/Landfill, respectively. For inland communities where access to the SARI line is not a viable option, brine minimization by brine concentrators and further crystallization prior to land filling are comparable to thermal evaporation ponds. Although the capital costs of BC/XLZR are more expensive than evaporation ponds, O&M costs for the BC/XLZR alternatives are slightly cheaper ($200,000). In the long run, it most likely could be a "greener" solution than the upkeep of an evaporation pond—which can occupy as much as 12 acres of land for the proposed treatment alternatives. The total cost calculated were a sum of the amortized capital annual costs plus O&M costs and did not deduct the revenue generated from recovered water through additional brine minimization. As mentioned in section 4.7, the excess SARI capacity was not sold in the cost calculations. Therefore, the actual total annual costs may be less costly than what is presented in table 4.9.

Although emerging technologies such as MD, FO, and SPARRO are evaluated in this study, there remains further development before these processes can be commercially available at more competitive pricing in the future.

Appendix A

UNIVERSITY OF NEVADA, RENO REPORT ON VEDCMD AND FO

Final Report

To

Carollo Engineers

Direct Contact Membrane Distillation and Forward Osmosis for the Concentration of Primary and Secondary RO Brine

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1. BACKGROUND

The demand for potable water is continuously increasing while water resources are being depleted, contaminated, or adversely affected by drought. The demand is usually met through the use of surface water or groundwater. Surface water is a viable resource in many regions, but areas with limited fresh water are exploring the use of non-conventional sources to meet their demands.

Seawater desalination has been used to produce potable water since the middle of the 20th century [1]. Large-scale land-based desalination facilities are common along coast lines [2]. Brackish water desalination is becoming recognized for inland communities with limited fresh water and seawater sources [3]. Brackish water salinity concentrations range from 1,000 to 8,000 mg/L TDS, compared to 35,000 mg/L of ocean water, and therefore, brackish water can be desalinated at substantially lower cost than seawater [1].

Current technologies for brackish water treatment comprise three well-established processes. These include chemical softening, pressure-driven membrane processes, and electrodialysis [2]. Chemical precipitation with coagulants such as lime or caustic soda can reduce hardness associated with calcium and magnesium ions present as carbonates, bicarbonates, chlorides, and sulfates [4]. Reverse osmosis (RO) and nanofiltration (NF) have been successful with seawater desalination; and the development of high performance low pressure RO and NF membranes have enabled the use of these processes in brackish water desalination. RO and NF membrane processes can remove hardness ions and monovalent ions such as sodium and chloride and can produce water of the highest purity [5]. Electrodialysis has been used for brackish water desalination since the 1960s [6]. The ion selective membranes used in ED are durable and their integrity is not easily compromised by feed streams with high scaling potential.

Although well-established, these processes are limited in their ability to meet the environmental and economic standards of today. Chemical treatment results in large volumes of sludge that require dewatering and transport to solid waste disposal facilities. Despite the technological advances of RO, NF, and ED, they are still inclined to high capital cost, are energy intensive, and membrane replacement costs are high and frequent. Fouling, and especially scaling, are high in RO, NF, and ED of brackish water and therefore, total water recovery is limited in order to prolong the life of the membranes. Large volumes of brine associated with limited water recovery may not be a concern for desalination facilities near a shoreline, but brine disposal is a considerable challenge for inland desalination facilities.

For these reasons, advanced processes are sought that can desalinate brackish water at higher recoveries without the associated problems experienced in currently used processes. Brine concentrators, crystallizers, and other zero liquid discharge (ZLD) systems are proven effective for volume minimization but are energy intensive and complex [2, 7, 8]. Less energy-intensive systems such as membrane distillation (MD) and forward osmosis (FO) can potentially minimize brine volume at lower energy expenditure and with less complexity [9, 10].

In the current study, direct contact MD (DCMD) and FO are being investigated as alternative processes to enhance water recovery from brackish water – or more specifically, from brines generated during brackish water desalination. DCMD is a non-pressure-driven separation process involving the transport of mass and heat through a hydrophobic, microporous membrane [11]. The driving force for mass transfer is the vapor pressure difference across the membrane. In VEDCMD, warmer feed water is in contact with the feed side of the membrane and a cooler water stream, under vacuum, is in direct contact with the opposite side of the membrane. The temperature difference between the streams and the vacuum applied on the permeate side of the membrane induce the vapor pressure gradient for mass transfer. The vacuum configuration of DCMD has been shown to increase flux by up to 85% over the traditional configuration [9]. FO is a non-pressure-driven separation process involving the diffusion of water through a semipermeable membrane. Water diffuses from a stream of low solute concentration to a draw solution (DS) stream of high osmotic pressure. The driving force for mass transport is the difference in osmotic pressure between the DS and feed solution. FO is shown to effectively concentrate a variety of feed streams, including municipal wastewater, landfill leachate, grey water, fruit juices, and beverages [10]. Studies have shown that both DCMD and FO can efficiently treat and highly concentrate a wide range of feed solutions.

2. INTRODUCTION

The West San Jacinto Groundwater Management Plan (WSJGMP) was adopted by the Eastern Municipal Water District (EMWD) in 1995. The goal of the plan is the implementation of a long-term salinity management program. Objectives of the program include groundwater monitoring, protection of adjacent sub-basins from salinity intrusion, development of salinity offsets to support extensive water recycling plans, and use of brackish groundwater as a local water supply. The Perris Basin Desalination Program (PBDP) will construct three desalination facilities to treat 15.3 MGD of brackish groundwater. From this, 5.1 MGD of brine will be produced and require disposal. Approximately 48,000 tons of salt will be removed from the basin annually.

The high cost of transporting brine from the desalination facilities and the unavailability of disposal capacity in the existing Santa Ana Regional Interceptor (SARI) necessitates an investigation into zero liquid discharge (ZLD) systems. Potential systems are being tested on two brines. The first brine being tested is concentrate from the primary RO desalter; it contains approximately 8,000 mg/L TDS, and is referred to as Brine A. The second brine is from two treatment steps used to further concentrate Brine A. In the first step, Brine A is softened by a caustic softening process and in the second step, the softened Brine A is treated in a secondary RO desalter. The concentrate from the secondary RO desalter, which is approximately 24,000 mg/L TDS, is referred to as Brine B. The five processes being evaluated for the ZLD system include:

- Recovery of salt for beneficial use using the SAL-PROC™ process by Geo-Processors (Sydney, Australia).
- Seeded RO treatment of Brine A.
- Brine concentrator treatment of Brine B with or without crystallization.
- Vacuum enhanced direct contact membrane distillation for the concentration of Brines A and B.
- Forward osmosis for the concentration of Brines A and B.

The scope of services performed by the Membrane Research Group at the University of Nevada, Reno include bench-scale testing and evaluation of vacuum enhanced DCMD (VEDCMD) and FO for the concentration of Brines A and B from the EMWD desalting facility.

The main objective of the current work is to determine the technical viability of using VEDCMD and FO processes to concentrate Brines A and B. This includes determining process performance, assessing maximum sustainable water recoveries, and estimating costs for each of the processes.

3. MATERIAL AND METHODS

3.1. Solution Chemistries

3.1.1. Feed Solutions

Two feed solutions were tested. Brine A was the primary RO concentrate produced at the EMWD groundwater desalting facilities (Menifee and Perris Desalters, Perris, CA); its initial TDS concentration was approximately 7,500 mg/L. Brine B was RO concentrate produced from a caustic/soda ash softening of Brine A; its initial TDS concentration was approximately 20,000 mg/L.

3.1.2. FO Draw Solution

ACS grade NaCl (Fisher Scientific, Pittsburgh, PA) was used to prepare the concentrated DS for the FO investigation. NaCl is an ideal DS because it has a high osmotic pressure at relatively low concentrations and also a high saturation concentration so it will not precipitate and scale the membrane [5].

3.1.3. Scale Inhibitors/Antiscalants

Antiscalants were used to investigate the effects of scale inhibition on water recovery. Pretreat Plus Y2K and Pretreat Plus 0400 are proprietary formulas developed by King Lee Technologies (San Diego, CA) to control membrane scaling. Pretreat Plus Y2K is used by EMWD to inhibit the formation of reactive silica and calcium carbonate scale on the RO membranes. Pretreat Plus 0400 was engineered specifically to inhibit calcium sulfate scale formation.

3.1.4. Cleaning Solution

A cleaning solution made of disodium ethylenediaminetetraacetic acid (Na_2EDTA) and sodium hydroxide (NaOH) was used to remove sulfate scale from the membrane feed surfaces. The pH of the solution was 11.8; the molar concentrations were 0.029 M Na_2EDTA and 0.058 M NaOH.

3.2. Membranes

Four membranes were used in this study. Three hydrophobic, microporous membranes were used for the MD study and one hydrophilic semi-permeable membrane was used for the FO study.

3.2.1. MD Membranes

Three types of MD membranes were tested, a capillary polypropylene membrane (Figure 1) (MD020-CP-2N, Microdyn, Germany), a flat sheet polypropylene membrane (GE polypropylene membrane, GE Osmonics, Minnetonka, MN) and a flat sheet PTFE membrane (GE PTFE (Teflon®) Laminated Membrane, GE Osmonics, Minnetonka, MN) both having 0.22

micron pore size. The capillary membrane is designed in a tube-and-shell configuration to allow for tangential flow on both the feed and permeate sides of the capillaries. The MD020-CP-2N module contains 40 capillaries with a total surface area of 0.1 m^2. Previous studies [9] have shown that the polypropylene membrane has very high salt rejection and good membrane integrity, but relatively low water flux; the PTFE membrane has high flux and high salt rejection. The flat sheet MD membranes were tested in a modified SEPA cell (Figure 1) (Osmonics, Minnetonka, MN), which has symmetric channels on both sides of the membrane to allow for cross flow on both the feed and permeate sides of the membrane. The active membrane surface area is 139 cm^2.

3.2.2. FO Membranes

One flat-sheet cellulose triacetate (CTA) FO membrane was tested in the FO experiments. The CTA membrane was specifically developed for FO applications and was acquired from Hydration Technologies Inc. (Albany, OR). In previous investigations, this membrane outperformed other semi permeable membranes (i.e., RO membranes) used for FO [12]. The CTA membrane was tested in the same modified SEPA cell used in the MD experiments.

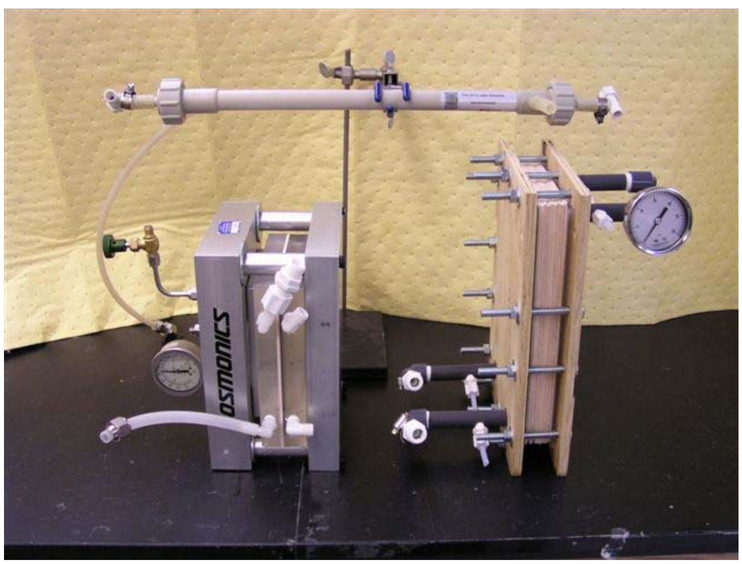


Figure 1. Bench scale test units. Modified SEPA cell (left), Microdyn capillary membrane (top), and a plate-and-frame bench-scale module (right).

3.3. Apparatuses and Procedures

MD and FO bench-scale experiments were conducted to investigate the capability of each process to concentrate brine from desalinated brackish water. Bench-scale configurations were different for the MD and FO setups but the experimental procedures were similar. Experiments were terminated when substantial flux decline occurred.

3.3.1. MD Test Unit and Procedures

The performance of the VEDCMD process was evaluated under various operating conditions using a bench-scale membrane test unit. A flow diagram of the bench-scale apparatus used in the VEDCMD investigation is illustrated in Figure 2. Feed solution at either 40 or 60ºC was recirculated on the feed side of the membrane at 1.5 LPM using a variable speed gear pump (Micropump Series GB, Cole-Parmer, Vernon Hills, IL). DI water at 20ºC was recirculated countercurrently on the support side of the membrane at 1.5 LPM using a constant speed gear pump (Procon, Murfreesboro, TN). DI water flow rate was controlled by bypassing some of the water back to the permeate tank. Both the feed and permeate flow rates were monitored using rotameters (Series FR4000, Key Instruments, Trevose, PA). Feed water temperature was controlled and maintained using a hot bath (Model 280, Precision Scientific, Winchester, VA) and permeate water temperature was controlled and maintained with a heat exchanger connected to a chiller (Model M-33, Thermo-Electron, Newington, NH). Feed and permeate temperatures were monitored using dual channel thermocouple thermometers (DigiSense DualLogR, Cole-Parmer, Vernon Hills, IL) located at their respective membrane unit inlet and outlet in order to investigate heat transfer and calculate the vapor pressure driving force. For each feed temperature, two different permeate pressures (vacuums) were tested – 360 mmHg absolute (deeper vacuum) and 660 mmHg absolute (lower vacuum). Permeate pressure was controlled using a back-pressure valve and monitored with vacuum pressure gauges at the membrane inlet and outlet. As water evaporated through the membrane, the concentration of the feed stream slowly increased and the water level in the permeate reservoir slowly increased. Excess water overflowed the permeate reservoir and was continuously collected and weighed on an analytical balance (PB3002-S DeltaRange, Mettler Toledo, Inc., Columbus, OH). The change in weight of the collection tank was recorded and used to calculate water flux and recovery. A conductivity meter (Jenway 4320, Jenway Ltd., U.K.) was used to monitor permeate conductivity in order to calculate salt rejection. MD experiments were stopped when substantial flux decline was observed.

3.3.2. MD Cleaning

Each flat sheet membrane was severely scaled during concentration experiments conducted at feed and permeate temperatures of 60 and 20°C, respectively. 4 L of DI water were flushed through the feed side of the system to remove loose precipitates. 2 L of an EDTA cleaning solution (section 3.3.4) were recycled at 2 LPM on the feed side for 30 minutes; the system was then stopped and the membrane was soaked in the cleaning solution for 1.5 hours. Subsequently, the feed channel was flushed with 8 L of DI water at 2.5 LPM.

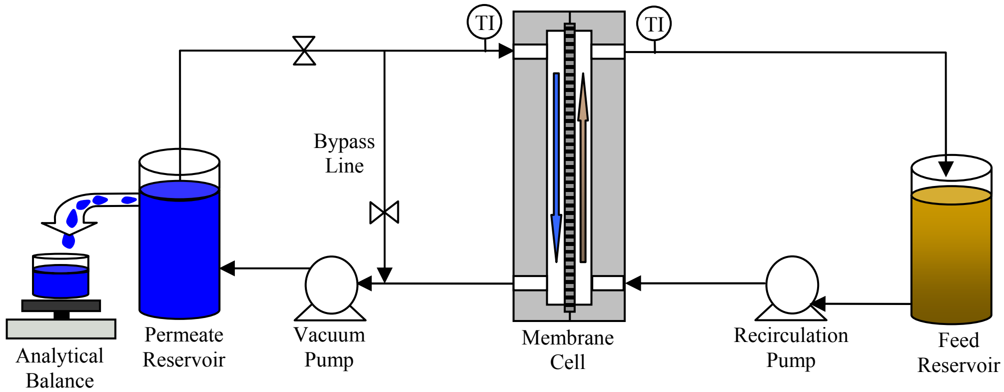


Figure 2. A schematic drawing of bench-scale MD apparatus.

3.3.3. FO Test Units and Procedures

The performance of the FO process was evaluated using the bench-scale membrane test unit coupled with a special pilot-scale RO system (Figure 3). The pilot-scale RO system supplied the FO system with DS at constant concentration. The RO system utilized an array of six 2.5" spiral wound membrane elements (Dow Filmtech). The concentrate stream from the RO system was recirculated through the DS tank to maintain a constant, highly concentrated DS. The conductivity of the DS was monitored using a conductivity meter (YSI Co. Model 30, Yellow Springs, OH) and minor adjustments were made when necessary. The feed solution was recirculated at 1.5 LPM on the feed side of the FO membrane and a concentrated DS of 50±3 g/l NaCl was recirculated countercurrently at 1.5 LPM on the support side of the membrane. Flowrates were controlled using variable speed gear pumps (Cole-Parmer, Vernon Hills, IL) and monitored with rotameters (Series FR4000, Key Instruments, Trevose, PA). The experiments were conducted at a constant temperature of 23±2^{o}C. The DS temperature was controlled using a heat exchanger fed by a chiller (Isotemp 1013S, Fisher Scientific, Pittsburgh, PA) and the feed was allowed to reach similar temperature by heat transfer through the membrane. Temperatures were monitored at the feed and permeate inlets using a dual channel thermocouple thermometer (DigiSense DualLogR, Cole-Parmer, Vernon Hills, IL). The water level in the feed reservoir declined and the TDS concentration of the feed stream increased as water diffused through the membrane. The change of weight of the feed reservoir was recorded using an analytical balance (PB5001, Mettler Toledo, Inc., Columbus, OH) and the data was used to calculate water flux, recovery, and feed concentration. The feed solution was recirculated on the feed side until water flux was substantially reduced.

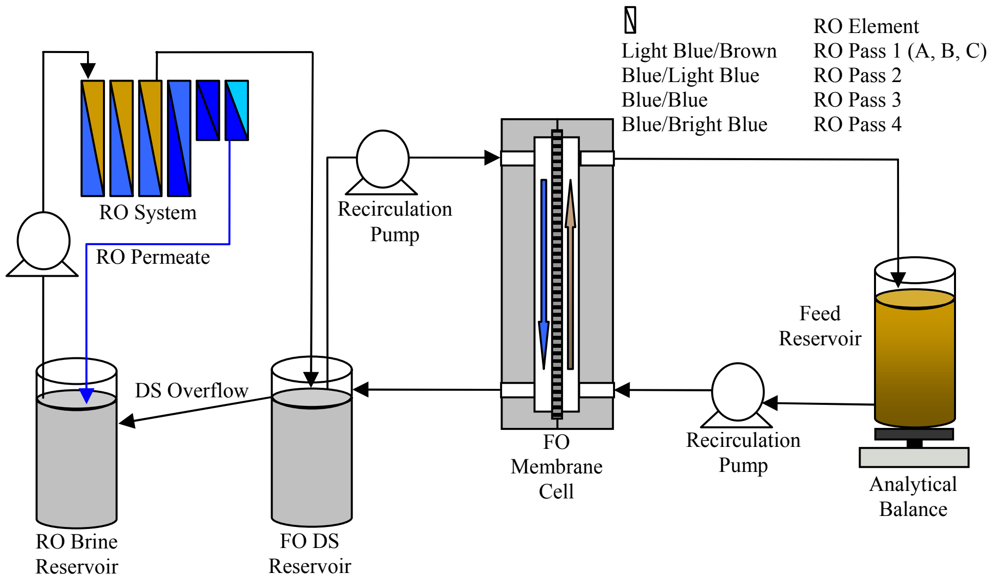


Figure 3. Schematic of the combined bench-scale FO and pilot-scale RO apparatus.

3.3.4. FO Cleaning

A series of four FO batch experiments were conducted to investigate the effectiveness of membrane cleaning. FO batch experiments were performed in a test unit similar to that shown in Figure 3, but without the RO system providing the highly concentrated DS. Instead a saturated NaCl solution was used as the DS in the experiments. A virgin CTA membrane was used for the four experiments. Brine A was used as the feed in each experiment. A chiller and heat exchanger were used to maintain the DS temperature at 21±2°C. As water crossed the membrane, the feed level decreased and the concentration increased; the DS was maintained at saturation concentration for the duration of the experiments.

3.3.4.1 Chemical Cleaning

The procedure was similar to the one used for the chemical cleaning of the MD membranes (Section 3.3.2).

3.3.4.2 Osmotic Backwash

An osmotic backwash was tested on the chemically cleaned CTA membrane after the membrane was scaled again. Prior to the osmotic backwash, 4 L of DI water were flushed through the feed and DS sides of the system to remove loose precipitates and to rinse NaCl from the system. For the osmotic backwash, DI water was recirculated for 20 minutes at 1.5 LPM on the DS side of the membrane and a 100 g/L NaCl solution was recirculated simultaneously on the feed side of the membrane. After the osmotic backwash, both sides of the system were thoroughly rinsed by flushing with 8 L of DI water.

3.3.5. FO Batch Experiment

The procedures used for the FO batch experiments were similar to those used in the FO cleaning experiments. Brine A was recirculated on the feed side of the membrane at 1.5 LPM while DS was recirculated countercurrently at 1.5 LPM on the support side of the membrane. 1 L of DS was prepared in a 4 L Erlenmeyer flask at a concentration of 70 g/L NaCl. As water diffused through the membrane, the level in the DS flask increased and the DS was diluted. Experiments were conducted until the DS was diluted to approximately 20 g/L NaCl.

3.3.6. Analysis

3.3.6.1 SEM Imaging

Scanning electron microscopy (SEM) was used to determine the major constituents of the scale that formed on the membrane surface. After an experiment was completed, the used membrane was not cleaned or rinsed before being removed from the modified SEPA cell and placed in a desiccator to dry. 1 cm coupons were cut from the dried membranes and positioned in an SEM to obtain either an edge or a surface view.

3.3.6.2 Water Quality

Brines A and B and the DS from the FO batch experiment were analyzed using Standard Methods. Standard Method (SM) 2320 B was used to analyze for alkalinity as $CaCO_3$, hydroxide, carbonate, and, bicarbonate. Environmental Protection Agency (EPA) Method 200.7 was used to analyze for calcium, magnesium, and silica. EPA Method 300.0 was used to analyze for chloride and sulfate. The DS from the FO batch experiment was analyzed to investigate the salt rejection of the CTA membrane. The permeate conductivity from the VEDCMD experiments was measured continuously to monitor reverse salt transport through the membrane (i.e., transport from the DS to the permeate) [5].

3.4. Process Modeling

3.4.1. FO/RO Modeling

A model was developed to estimate the energy demand and major operating costs required to treat 1 MGD of brackish water RO brine at a total water recovery of 70%. The model consisted of an FO model used in conjunction with an RO model. The FO model was an Excel-based model developed in the laboratory and the RO model was Reverse Osmosis System Analysis (ROSA) available from DOW-Filmtec (Midland, MI).

The FO model (Figure 4) was developed as a vessel containing two membrane elements connected in series. The number of vessels used in parallel was determined using the amount of brackish water to be processed and the expected water recovery. Each FO membrane element was divided into twenty area elements where the flows, concentrations, and osmotic pressures of the feed and DS streams were calculated based on values obtained in the previous area element. Six input parameters were required to initiate the model: 1) membrane permeability coefficient (determined experimentally to be $8.68 \cdot 10^{-8}$ m/sec-bar), 2) membrane surface area (20 times the area of one area element in one module), 3) DS inlet flowrate, 4) DS inlet concentration, 5) feed inlet flowrate, and 6) feed inlet concentration. The model does not account for concentration polarization, temperatures, and scaling effects. Currently, these are incorporated in the water permeability coefficient.

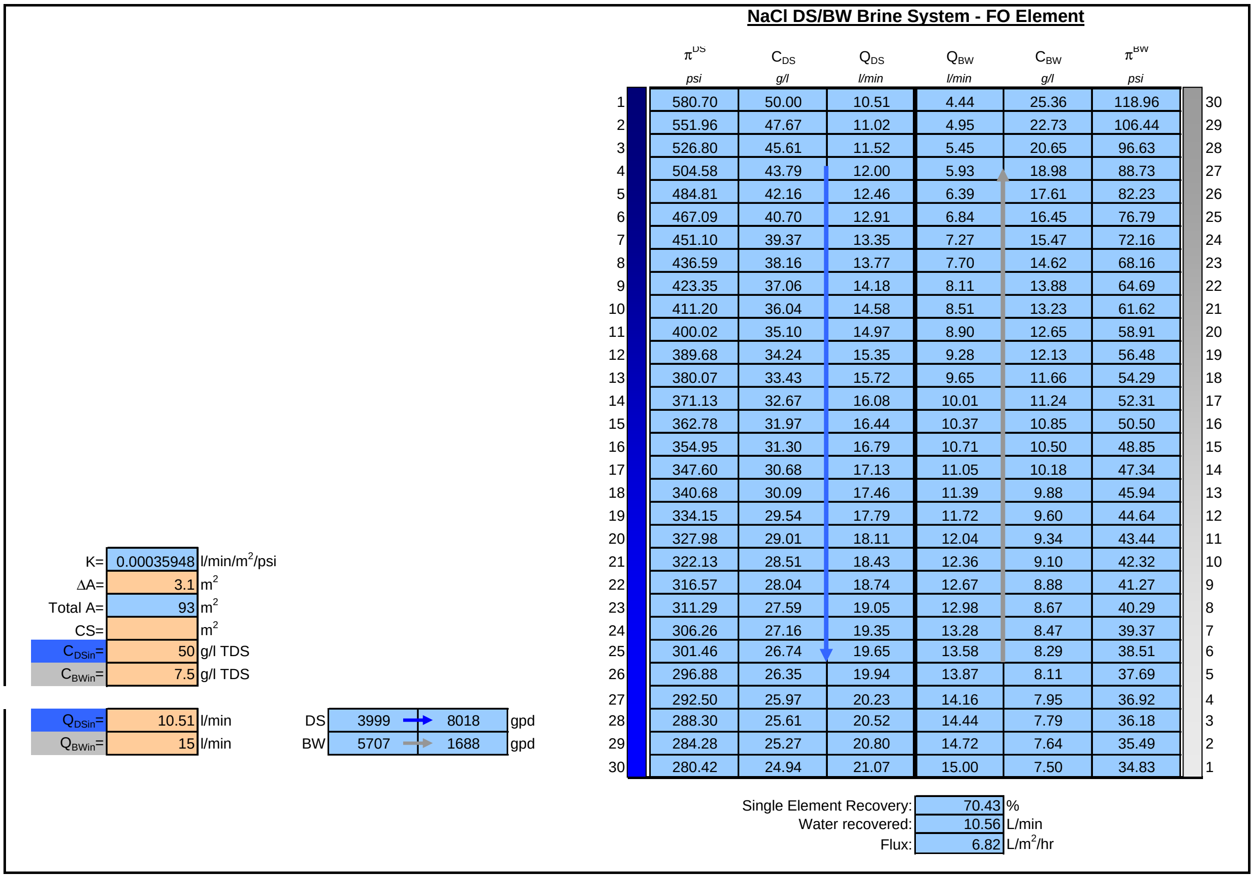
The membrane area was arbitrarily chosen to initiate the model and then adjusted until a predetermined recovery was achieved. The DS inlet concentrations was chosen to be 50 g/L NaCl and the feed concentration was 7.5 g/L TDS for Brine A and 20 g/L TDS for Brine B. The DS and feed inlet flowrates were chosen arbitrarily and then adjusted to achieve the predetermined recovery. The model used an iterative numerical method to calculate the module outputs, including outlet feed and DS concentrations and flowrates and module permeation rate.

The outlet DS concentration and flowrate were used as inputs into ROSA. A one-stage membrane array was used. The feed concentration was obtained from the FO model and the permeate stream was the final product water. The RO system recovery was determined based on the requirement to produce a reject stream concentrated to 50 g/L NaCl. The SW30HR LE-400i (Filmtec) spiral wound RO membrane element was selected for the model. The RO system was scaled up to produce 1 MGD of product water at a total recovery of 68%.

The specific energy and power consumption were calculated by ROSA for the RO system. Because FO is a non-pressure driven membrane process, the power and energy were calculated based on the horsepower of specified pumps and the required amount of water to be pumped. 1.5 horsepower pumps (Stock# 2P006, Grainger Industrial Supply, Lake Forest, IL) were used as a reference for the calculations. The specific energy required for the FO process was calculated using the ratio between the energy consumed when pumping 1.34 MGD of feedwater and the volume of water produced per day. Power consumption was calculated as the product of the specific energy and the volume produced. ***NOTE that the energy consumption of an FO/RO system for desalination of brine from brackish water desalination is comparable to the energy consumption for seawater desalination. Most of the energy demand in the FO/RO system is due to the high pressure RO pump required to desalinate the NaCl draw solution from approximately 15,000 ppm to approximately 50,000 ppm and an additional small demand is due to the feed brackish water recirculation pumps.***

3.4.2. MD Modeling

A mathematical model was developed to describe the pilot-scale VEDCMD process operated with Microdyn membrane elements. The Excel-based spreadsheet (Figure 5 and 6) uses a numerical model that considers both physical and thermodynamical forces and describes the effects of both vacuum and flow velocity on the flux of water vapors across the membrane. To do this, the basic governing equations for DCMD were considered and are incorporated with the governing equations of heat transfer in heat exchangers. Like in the FO modeling, each MD membrane element was divided into area elements, this time into 28. Multiple parameters are required to initiate the model. These include membrane element characteristics (e.g., number of fibers, fiber diameter and length, pressure vessel size, and membrane pore size), hydraulic characteristics (including feed and cold water flowrates, concentrations, and temperatures), and thermodynamic characteristics of the fluids (density, viscosity, enthalpy, and dimensionless numbers). The model considers several membrane modules in series with feed reheating stages between membranes and once-through flow of cold water on the permeate side of each membrane element.



	π^{DS} (psi)	C_{DS} (g/l)	Q_{DS} (l/min)	Q_{BW} (l/min)	C_{BW} (g/l)	π^{BW} (psi)	
1	580.70	50.00	10.51	4.44	25.36	118.96	30
2	551.96	47.67	11.02	4.95	22.73	106.44	29
3	526.80	45.61	11.52	5.45	20.65	96.63	28
4	504.58	43.79	12.00	5.93	18.98	88.73	27
5	484.81	42.16	12.46	6.39	17.61	82.23	26
6	467.09	40.70	12.91	6.84	16.45	76.79	25
7	451.10	39.37	13.35	7.27	15.47	72.16	24
8	436.59	38.16	13.77	7.70	14.62	68.16	23
9	423.35	37.06	14.18	8.11	13.88	64.69	22
10	411.20	36.04	14.58	8.51	13.23	61.62	21
11	400.02	35.10	14.97	8.90	12.65	58.91	20
12	389.68	34.24	15.35	9.28	12.13	56.48	19
13	380.07	33.43	15.72	9.65	11.66	54.29	18
14	371.13	32.67	16.08	10.01	11.24	52.31	17
15	362.78	31.97	16.44	10.37	10.85	50.50	16
16	354.95	31.30	16.79	10.71	10.50	48.85	15
17	347.60	30.68	17.13	11.05	10.18	47.34	14
18	340.68	30.09	17.46	11.39	9.88	45.94	13
19	334.15	29.54	17.79	11.72	9.60	44.64	12
20	327.98	29.01	18.11	12.04	9.34	43.44	11
21	322.13	28.51	18.43	12.36	9.10	42.32	10
22	316.57	28.04	18.74	12.67	8.88	41.27	9
23	311.29	27.59	19.05	12.98	8.67	40.29	8
24	306.26	27.16	19.35	13.28	8.47	39.37	7
25	301.46	26.74	19.65	13.58	8.29	38.51	6
26	296.88	26.35	19.94	13.87	8.11	37.69	5
27	292.50	25.97	20.23	14.16	7.95	36.92	4
28	288.30	25.61	20.52	14.44	7.79	36.18	3
29	284.28	25.27	20.80	14.72	7.64	35.49	2
30	280.42	24.94	21.07	15.00	7.50	34.83	1

Figure 4. Interface of the FO model for one membrane element.

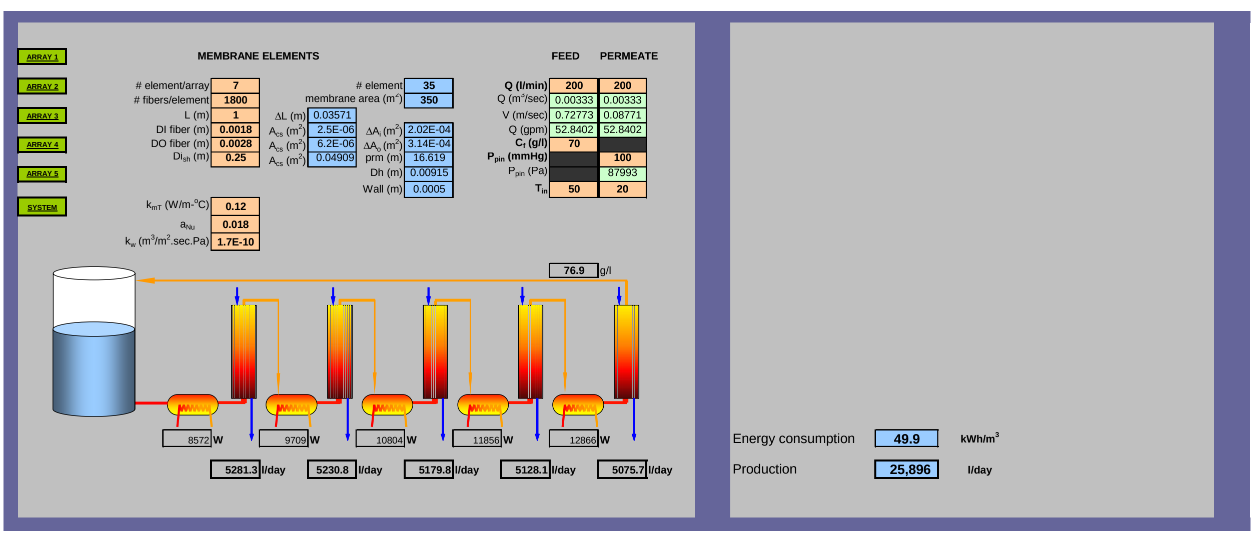


Figure 5. Interface of the VEDCMD modeling for five membrane element.

ARRAY #1

HOME | | ARRAY 2 | ARRAY 3 | ARRAY 4 | ARRAY 5

E--> (W)

	Nu	Re	Pr	μ, kg/m.s	k_{wT}, W/m.	ρ, kg/m³	h, W/m².C	v, m/sec	Q, m³/sec	k_w (m³/m².sec.P	P_v(Pa)	C (g/l)	λ_{fg} (J/kg)	Ui (W/m².°C)
1	2.9	358	3.5856	0.0006	0.6447	1058	1044.2	0.104	3.333E-03	1.4660E-10	12352	70	2382757	6.6812E-01
2	2.9	355	3.6163	0.0006	0.6442	1058.2	1038.9	0.1039	3.330E-03	1.4657E-10	12050	70.061	2383954	6.6810E-01
3	2.9	352	3.6468	0.0006	0.6436	1058.5	1033.8	0.1038	3.328E-03	1.4654E-10	11761	70.121	2385125	6.6809E-01
4	2.9	349	3.6771	0.0006	0.6431	1058.8	1028.7	0.1037	3.325E-03	1.4652E-10	11483	70.179	2386272	6.6807E-01
5	2.9	346	3.7071	0.0006	0.6425	1059	1023.8	0.1036	3.322E-03	1.4649E-10	11216	70.236	2387394	6.6806E-01
6	2.9	344	3.737	0.0006	0.642	1059.3	1018.9	0.1035	3.319E-03	1.4647E-10	10960	70.293	2388494	6.6804E-01
7	2.8	341	3.7666	0.0006	0.6415	1059.5	1014.1	0.1034	3.317E-03	1.4645E-10	10714	70.348	2389572	6.6803E-01
8	2.8	338	3.7961	0.0006	0.641	1059.7	1009.5	0.1034	3.314E-03	1.4642E-10	10477	70.402	2390628	6.6801E-01
9	2.8	336	3.8254	0.0006	0.6405	1060	1004.9	0.1033	3.312E-03	1.4640E-10	10249	70.455	2391663	6.6800E-01
10	2.8	333	3.8545	0.0006	0.64	1060.2	1000.4	0.1032	3.309E-03	1.4638E-10	10030	70.507	2392679	6.6798E-01
11	2.8	331	3.8835	0.0006	0.6395	1060.4	996.0	0.1031	3.307E-03	1.4636E-10	9818.1	70.558	2393675	6.6797E-01
12	2.8	328	3.9123	0.0006	0.639	1060.6	991.7	0.1031	3.305E-03	1.4634E-10	9614.2	70.608	2394653	6.6795E-01
13	2.8	326	3.941	0.0006	0.6385	1060.8	987.4	0.103	3.302E-03	1.4632E-10	9417.4	70.658	2395613	6.6794E-01
14	2.8	324	3.9696	0.0006	0.638	1061	983.2	0.1029	3.300E-03	1.4630E-10	9227.6	70.706	2396556	6.6792E-01
15	2.8	322	3.998	0.0006	0.6376	1061.2	979.1	0.1029	3.298E-03	1.4629E-10	9044.3	70.754	2397482	6.6791E-01
16	2.8	319	4.0264	0.0006	0.6371	1061.4	975.1	0.1028	3.296E-03	1.4627E-10	8867.2	70.8	2398391	6.6789E-01
17	2.7	317	4.0546	0.0006	0.6366	1061.6	971.1	0.1027	3.294E-03	1.4625E-10	8696.1	70.847	2399285	6.6788E-01
18	2.7	315	4.0826	0.0006	0.6362	1061.8	967.2	0.1027	3.291E-03	1.4624E-10	8530.6	70.892	2400163	6.6786E-01
19	2.7	313	4.1106	0.0006	0.6357	1062	963.3	0.1026	3.289E-03	1.4622E-10	8370.5	70.936	2401027	6.6785E-01
20	2.7	311	4.1385	0.0006	0.6353	1062.1	959.5	0.1025	3.287E-03	1.4621E-10	8215.5	70.98	2401877	6.6784E-01
21	2.7	309	4.1663	0.0006	0.6349	1062.3	955.8	0.1025	3.285E-03	1.4619E-10	8065.5	71.023	2402712	6.6782E-01
22	2.7	307	4.194	0.0006	0.6344	1062.5	952.1	0.1024	3.283E-03	1.4618E-10	7920.1	71.066	2403534	6.6781E-01
23	2.7	305	4.2216	0.0006	0.634	1062.6	948.5	0.1023	3.281E-03	1.4617E-10	7779.3	71.108	2404343	6.6779E-01
24	2.7	303	4.2491	0.0006	0.6336	1062.8	944.9	0.1023	3.280E-03	1.4615E-10	7642.8	71.149	2405139	6.6778E-01
25	2.7	301	4.2765	0.0006	0.6331	1063	941.4	0.1022	3.278E-03	1.4614E-10	7510.3	71.189	2405923	6.6777E-01
26	2.7	299	4.3039	0.0007	0.6327	1063.1	937.9	0.1022	3.276E-03	1.4613E-10	7381.9	71.229	2406695	6.6775E-01
27	2.7	297	4.3312	0.0007	0.6323	1063.3	934.5	0.1021	3.274E-03	1.4612E-10	7257.2	71.269	2407455	6.6774E-01
28	2.7	296	4.3584	0.0007	0.6319	1063.4	931.1	0.1021	3.272E-03	1.4611E-10	7136.1	71.308	2408204	6.6773E-01

	hfv	k	T_{in} (°C)	T_{out} (°C)	P_{Tot}(Pa)	C (g/l)	P_v(Pa)	Q, m³/sec	v, m/sec	h, W/m².C	ρ, kg/m³	k_{wT}, W/m.	μ, kg/m.sec	Pr	Re	Nu
1	6.9E+03	0.002616	50	30.6	98		4397.724	0.00339446	0.0128	139.9	994.89	0.6197	7.84E-04	5.2742	148.14	2.1
2	6.9E+03	0.002615	49.5	30.1	98.1		4276.931	0.00339156	0.0127	138.99	995.03	0.619	7.92E-04	5.3359	146.51	2.1
3	6.8E+03	0.002615	49.0	29.6	98.1		4158.948	0.00338873	0.0127	138.08	995.17	0.6183	8.00E-04	5.3989	144.89	2.0
4	6.6E+03	0.002615	48.5	29.2	98.2		4046.199	0.00338595	0.0127	137.19	995.3	0.6175	8.09E-04	5.4619	143.30	2.0
5	6.5E+03	0.002615	48.1	28.7	98.3		3938.371	0.00338324	0.0127	136.31	995.43	0.6168	8.17E-04	5.5249	141.76	2.0
6	6.3E+03	0.002615	47.6	28.2	98.4		3835.169	0.00338058	0.0127	135.46	995.55	0.6161	8.25E-04	5.5879	140.24	2.0
7	6.2E+03	0.002615	47.2	27.8	98.4		3736.325	0.00337798	0.0127	134.61	995.67	0.6154	8.33E-04	5.6509	138.76	2.0
8	6.1E+03	0.002615	46.7	27.3	98.5		3641.587	0.00337544	0.0127	133.79	995.79	0.6148	8.42E-04	5.7139	137.32	2.0
9	6.0E+03	0.002615	46.3	26.9	98.6		3550.723	0.00337294	0.0127	132.98	995.9	0.6141	8.50E-04	5.7768	135.91	2.0
10	5.9E+03	0.002615	45.9	26.5	98.7		3463.516	0.0033705	0.0127	132.18	996.01	0.6134	8.58E-04	5.8397	134.52	2.0
11	5.7E+03	0.002615	45.5	26.1	98.7		3379.765	0.0033681	0.0127	131.4	996.11	0.6128	8.66E-04	5.9026	133.17	2.0
12	5.6E+03	0.002615	45.0	25.7	98.8		3299.282	0.00336575	0.0127	130.63	996.21	0.6122	8.74E-04	5.9655	131.85	2.0
13	5.5E+03	0.002615	44.6	25.3	98.9		3221.893	0.00336344	0.0126	129.88	996.31	0.6115	8.83E-04	6.0283	130.55	1.9
14	5.4E+03	0.002615	44.3	24.9	99.0		3147.434	0.00336118	0.0126	129.14	996.41	0.6109	8.91E-04	6.0911	129.28	1.9
15	5.3E+03	0.002615	43.9	24.5	99.0		3075.752	0.00335895	0.0126	128.41	996.5	0.6103	8.99E-04	6.1538	128.04	1.9
16	5.2E+03	0.002614	43.5	24.1	99.1		3006.707	0.00335677	0.0126	127.7	996.59	0.6097	9.07E-04	6.2165	126.83	1.9
17	5.1E+03	0.002614	43.1	23.7	99.2		2940.163	0.00335463	0.0126	126.99	996.68	0.6091	9.15E-04	6.2792	125.64	1.9
18	5.1E+03	0.002614	42.7	23.4	99.3		2875.997	3.353E-03	0.0126	126.3	996.77	0.6085	9.23E-04	6.3418	124.47	1.9
19	5.0E+03	0.002614	42.4	23.0	99.3		2814.091	0.00335046	0.0126	125.62	996.85	0.6079	9.31E-04	6.4043	123.32	1.9
20	4.9E+03	0.002614	42.0	22.6	99.4		2754.335	0.00334843	0.0126	124.95	996.93	0.6073	9.39E-04	6.4668	122.20	1.9
21	4.8E+03	0.002614	41.7	22.3	99.5		2696.627	0.00334643	0.0126	124.29	997.01	0.6068	9.47E-04	6.5293	121.10	1.9
22	4.7E+03	0.002614	41.3	22.0	99.6		2640.87	0.00334447	0.0126	123.64	997.09	0.6062	9.55E-04	6.5916	120.03	1.9
23	4.6E+03	0.002613	41.0	21.6	99.6		2586.972	0.00334253	0.0126	123.01	997.16	0.6056	9.63E-04	6.6539	118.97	1.9
24	4.6E+03	0.002613	40.7	21.3	99.7		2534.848	0.00334063	0.0126	122.38	997.23	0.6051	9.71E-04	6.7162	117.93	1.9
25	4.5E+03	0.002613	40.3	21.0	99.8		2484.416	0.00333877	0.0126	121.76	997.3	0.6045	9.79E-04	6.7784	116.92	1.8
26	4.4E+03	0.002613	40.0	20.6	99.9		2435.602	0.00333693	0.0125	121.15	997.37	0.604	9.87E-04	6.8405	115.92	1.8
27	4.4E+03	0.002613	39.7	20.3	99.9		2388.334	0.00333512	0.0125	120.55	997.44	0.6035	9.95E-04	6.9025	114.94	1.8
28	4.3E+03	0.002612	39.4	20.0	100		2342.543	0.00333333	0.0125	119.96	997.5	0.603	1.00E-03	6.9645	114	1.8
			Tf_{out} ,°C	Tp_{in} ,°C	P_{Tot}(Pa)	C (g/l)	P_v(Pa)	Q, m³/sec	v, m/sec	h, W/m².	ρ, kg/m	k_{wT}, W/r	μ, kg/m.se	Pr	Re	Nu

220.052 l/hr

8572.377 W

220.0522

Figure 6. The VEDCMD model for one membrane element.

4. RESULTS AND DISCUSSION

4.1. VEDCMD

Membrane scaling at elevated feed concentrations severely reduced the water flux through the Microdyn capillary membrane. Water flux was only partially recovered following a membrane cleaning. Subsequent experiments were performed using the flat-sheet PTFE membrane and PP membrane.

VEDCMD experiments were performed using Brines A and B as feed solutions. Water flux versus feed concentration is illustrated in Figure 6. Initial water flux was substantially greater in experiments conducted at $\Delta T = 40°C$ than those conducted at $\Delta T = 20°C$. This is because a larger ΔT generates a greater vapor pressure gradient across the membrane and results in increased water flux. Water flux was further improved by decreasing the permeate pressure from 660 mmHg to 360 mmHg. By decreasing the permeate pressure, the partial vapor pressure gradient across the membrane is increased – resulting in higher water flux.

For Brine A (Figure 6a), a relatively constant water flux was observed during the initial stage of the concentration experiments. Between 10,000 mg/L and 13,000 mg/L TDS, a sharp flux decline was observed. Because previous VEDCMD experiments conducted with two different feed solutions (NaCl and sea salt) demonstrated a gradual (9%) water flux decline over a feed concentration range of 0 g/L to 75 g/L [9], it is likely that the flux decline in the current investigation was due to scale formation on the membrane surface formed when sparingly soluble salts precipitated out of solution. The decline continued until the water flux was below 5 $L/m^2/hr$ and the experiment was terminated.

In the experiments conducted with Brine B (Figure 6b), with the exception of one of the experiments conducted at the lower pressure gradient across the membrane (square black symbol), flux decline started after the feed was only minimally concentrated. The higher water flux associated with the increased pressure gradient, likely increased concentration polarization at the membrane surface and resulted in earlier supersaturation and onset of water flux decline. The results suggest that feed concentration has minimal effect on water flux in VEDCMD processes but eventually, scale formation at higher feed concentrations is detrimental. Based on previous VEDCMD experiments (which had higher feed concentrations), water flux was expected to be relatively constant and approximately 90% total water recovery was anticipated [9]. However, because of membrane scaling, 62% and 81% were the highest total water recoveries obtained for Brines A and B, respectively.

When considering the full process recovery (EMWD RO (70% for Brine A and 88.8% for Brine B) plus VEDCMD) the numbers are more encouraging. Using Equation (1):

$$R_{tot} = R_{RO} + (1 - R_{RO})R_{VEDCMD} \tag{1}$$

the overall system recovery for Brine A would be 88.6% and the overall system recovery for Brine B would be 97.9%.

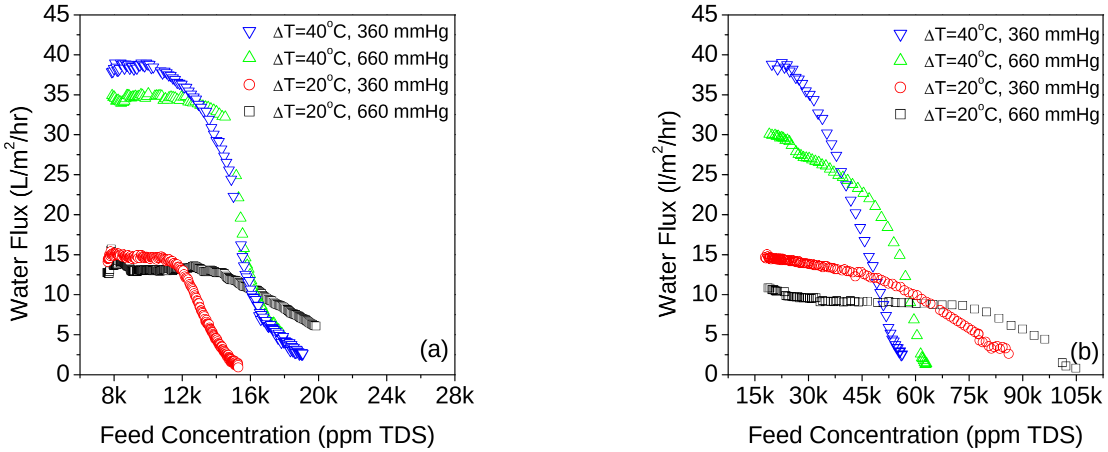


Figure 6. Water flux versus feed concentration during VEDCMD of (a) Brine A and (b) Brine B. All experiments were performed using the flat-sheet PTFE membrane.

Water fluxes during VEDCMD of Brines A and B are presented as a comparison of two feed temperatures in Figure 7. Initial water flux was very similar for Brines A and B at each temperature and pressure. This suggests that indeed, feed concentration has a minimal effect on water flux in VEDCMD. Total water recovery achieved in the experiments conducted at ΔT = 40°C ranged from 60-65% and at ΔT = 20°C from 60-81%.

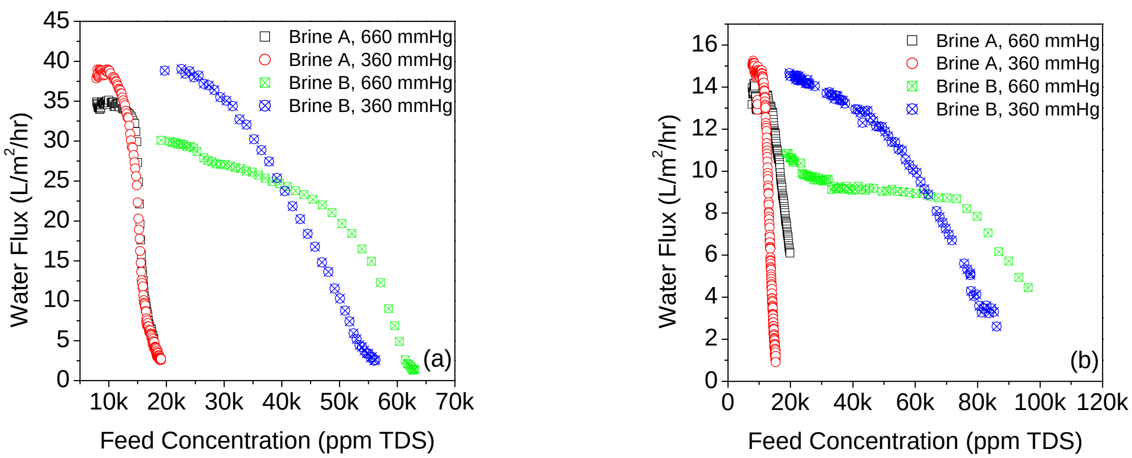


Figure 7. Comparison of Brine A and Brine B results at (a) ΔT = 40°C and (b) ΔT=20°C.

4.1.1. VEDCMD cleaning

The performance of VEDCMD with new flat sheet membranes and after membrane cleanings was compared. Water flux and recovery as a function of time for the polypropylene membrane and PTFE membranes are shown in Figure 8a and b, respectively. Brine A was used as feed in both sets of experiments. EDTA cleaning solution was used to clean the membranes. The feed temperature was maintained at 60±1°C during the experiments with the polypropylene membrane and the permeate temperature for all experiments was maintained at 20±0.3°C. As can be seen in Figure 8a, the performance of the polypropylene membrane was similar with a new membrane and after membrane cleaning, implying that membrane scaling in VEDCMD is almost completely reversible. As can be seen in Figure 8b, water flux through the PTFE membrane declined at a slower rate than the polypropylene membrane (Figure 8a) but the membrane performance was not recovered after cleaning.

The feed temperature used during the PTFE membrane experiments was 58±0.3°C. Based on previous experience, the flux through the PTFE membrane was expected to be closer to 40 L/m^2/hr, but was most likely lower because of operation at slightly lower feed temperature. Flux through the polypropylene membrane declined at a faster rate but the membrane performance was almost completely recovered following the cleaning procedure. This suggests that scale deposited on the surface of the membrane but did not clog the pores. This was also confirmed by the high salt rejection – greater than 99.9% rejection was observed in all experiments. Past studies [13, 14] have demonstrated that crystallization at the pore entrance causes pore flooding and consequently, lowers salt rejection. The cleaned PTFE membrane rejected salts similar to the new one; however, the sharp flux decline after cleaning suggests that scaling may be associated with the membrane material.

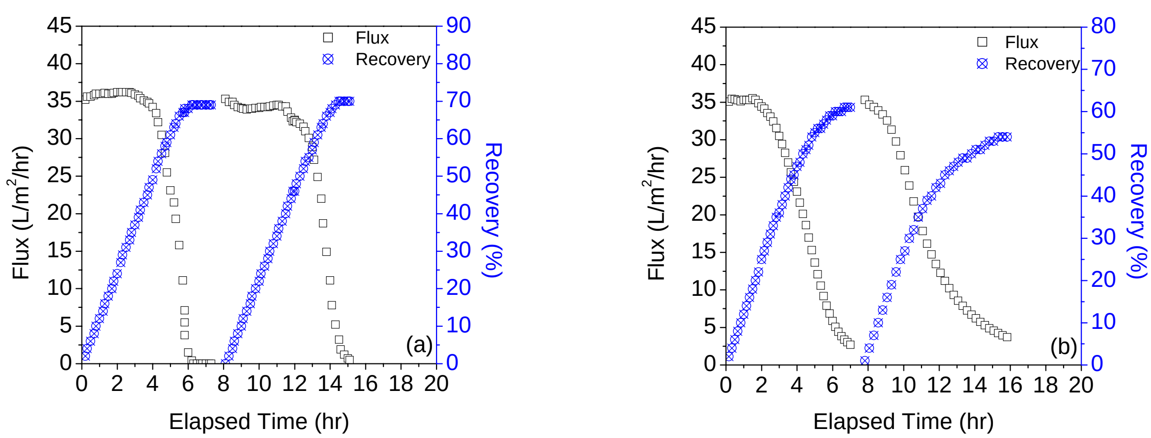


Figure 8. Comparison of VEDCMD cleaning results; (a) flat sheet polypropylene membrane and (b) flat sheet PTFE membrane.

4.2. FO

All FO experiments were carried out using the flat-sheet CTA membrane. Water flux versus feed concentration for Brines A and B are illustrated in Figure 9. The DS was maintained at a constant concentration of 50,000±3,000 mg/L using the pilot-scale RO system. The osmotic pressure difference between the feed stream and the DS stream was large enough to maintain a driving force capable of producing a relatively high flux. The initial water flux was comparable to those achieved using VEDCMD at $\Delta T = 20°C$. Water flux in the Brine B experiment was substantially lower than Brine A at the same feed TDS concentration. Thus, although Brines A and B have the same TDS concentration, they do not have the same chemical compositions and they will result in different flux declines.

Water flux declined continuously in both experiments. As water crossed the membrane, the TDS concentration, and thus osmotic pressure, of the feed stream increased; the DS concentration was maintained at constant concentration by the RO system. Therefore, the osmotic pressure gradient across the membrane was reduced as Brines A and B were concentrated. ***When reading the graph in Figure 9, it is important to realize that the initial flux decline is not due to fouling or scaling (as commonly seen in pressure-driven fouling results) but mainly due to decreased driving force.***

Scale did form and deposit on the surface of the membrane during the experiments. Results for Brine A in Figure 9 indicate a sudden change in the rate of flux decline at a feed concentration of approximately 25,000 ppm TDS. Further changes in the rate of flux decline/increase were repeatedly observed and are not well understood. It is believed that due to scale inhibitor residues in Brine B, no noticeable change in flux rate was observed. The water recoveries for Brines A and B were 86% and 51%, respectively. Brine B had a greater osmotic pressure than Brine A and consequently, water flux declined to an unacceptable level at a relatively low water recovery.

When considering the full process recovery (EMWD RO plus VEDCMD), the overall system recovery for Brine A would be 96% and the overall system recovery for Brine B would be 95%.

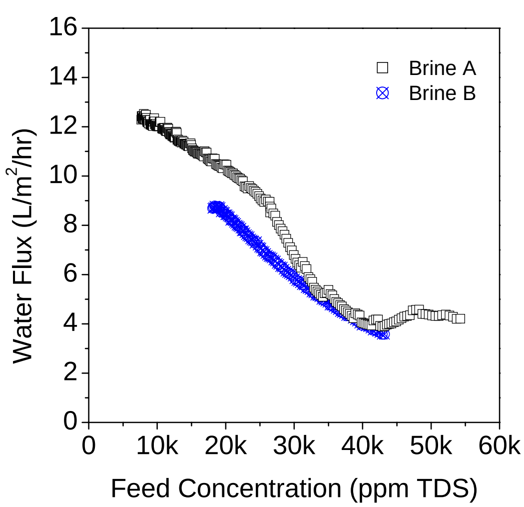


Figure 9. Water flux versus feed concentration during FO of Brines A and B.

4.2.1. FO cleaning

The performance of the FO membrane when it was new and after it had been cleaned is compared in Figure 10a. Four concentration experiments were conducted using Brine A as the feed. EDTA cleaning solution was used to clean the CTA membrane after the first concentration experiment. Osmotic backwashing was used to clean the membrane after the second and after the third concentration experiments. The high water fluxes during the cleaning experiments were due to the high DS concentration. ***It is important to note again that large fraction of the flux decline seen in Figure 10 is due to decline in driving force and <u>not</u> due scaling***.

Membrane performance was recovered following chemical cleaning. This suggests that EDTA was effective at removing calcium sulfate scaling from the CTA membrane and from the FO system. The third experiment revealed that flux was recovered after the first osmotic backwashing. However, upon initiation of the third experiment, scale was visible in the feed lines; flakes of scale were flushed from the system and settled at the bottom of the feed reservoir. This suggests that scale was not completely removed from the system by flushing with DI water and osmotic backwashing. The more rapid flux decline is likely due to a higher apparent osmotic pressure near the feed surface of the membrane caused by scale that remained in the feed spacer. The fourth experiment revealed substantially lower flux than in the previous experiments but still followed a similar trend. This suggests that osmotic backwashing was not effective in cleaning the membrane. However, it is possible that scale trapped in the feed spacer blocked areas of the membrane and reduced the effective surface area of the membrane.

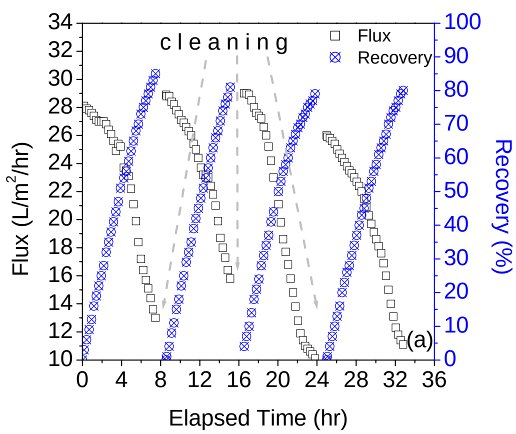


Figure 10. Four FO concentration experiments were performed. EDTA chemical cleaning was performed after the first and osmotic backwashing was performed after the second and third concentration experiments.

4.2.2. FO batch experiment

An FO batch experiment was performed to investigate the rejection of ions by the CTA membrane. A 70 g/L NaCl DS was prepared and was diluted to 20 g/L as 2.5 L of water crossed the membrane. The results of the water quality analysis of the DS are shown in Table 1. The results suggest that some ions crossed the membrane. High TDS in the DS increased the reporting limits of many ions and exact values were not obtained. These results suggest good ion rejection but not 100%. The water quality of Brines A and B are given in Section 4.3.2.

Table 1. Ions from Brine A that were found in the DS.

TEST	METHOD	RESULTS (mg/L)	REPORTING LIMIT (mg/L)
Alkalinity as $CaCO_3$	SM 2320 B	< 20	20
Hydroxide	SM 2320 B	< 7	7
Carbonate	SM 2320 B	< 12	12
Bicarbonate	SM 2320 B	< 25	25
Calcium	EPA 200.7	18	5
Magnesium	EPA 200.7	5	5
Silica	EPA 200.7	14	1
Sulfate	EPA 300.0	< 500	500

4.2.3. FO with scale inhibitor

Results from experiments conducted with the scale inhibitor are compared with results from an experiment conducted without scale inhibitor (Figure 11). Flux decline due to the change in osmotic pressure of the feed solution is the same for both experiments. However, the sharp decline in water flux associated with membrane scaling is not observed in the experiment when scale inhibitor is used. This suggests that scale inhibitors can be used to enhance the total water recovery of the FO process.

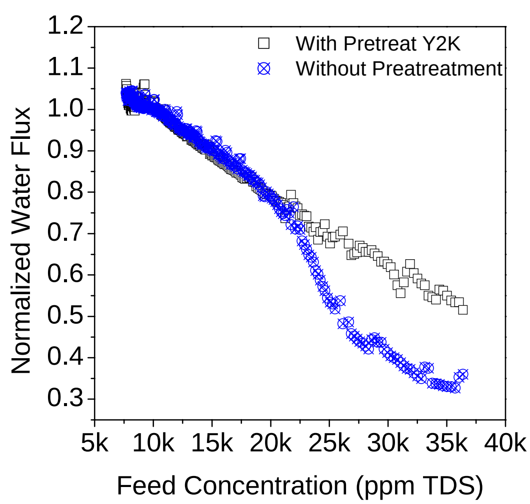


Figure 11. FO Performance with and without antiscalant.

4.3. Analysis

4.3.1. SEM imaging

In experiments with Brine A, calcium sulfate scale formed on the feed surface (Figure 12) and on the mesh spacer (Figure 13) of the PTFE MD membrane. Scale also formed in the tubing, feed reservoir, pump (moving and stationary parts), and heat exchanger, as well as on the membrane. In experiments conducted with Brine B, scale was not visible on any components of the system – implying that residual scale inhibitor in the feed solution may have played a role in delaying membrane scaling. However, scale did form and was visible on the feed surface (Figure 14) and mesh spacer (Figure 15). Chromatographs of the scale layer on the PTFE membrane are shown in **Figure 16**. These indicate that the precipitate is primarily calcium sulfate.

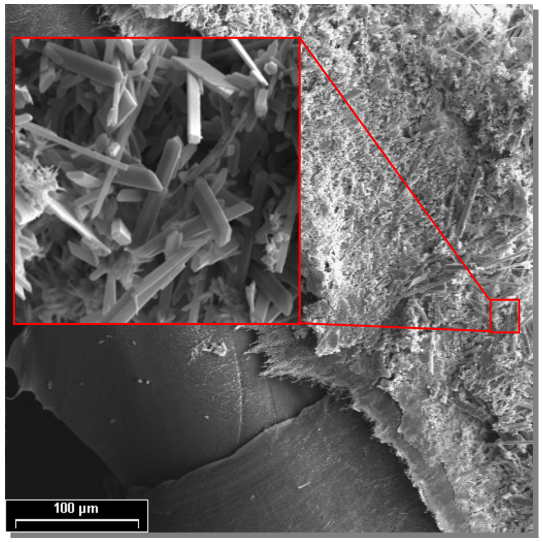


Figure 12. SEM image of calcium sulfate scale on the PTFE membrane used during the Brine A VEDCMD experiments.

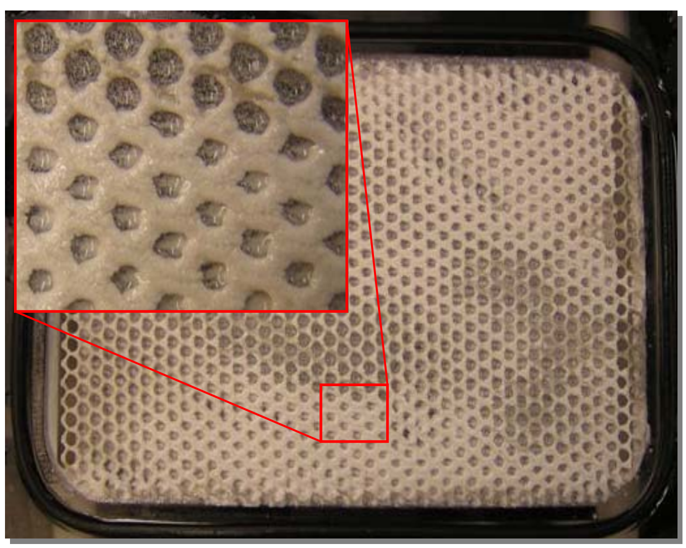


Figure 13. Photograph showing the severity of scale on the spacers used during the Brine A VEDCMD experiments.

Figure 14. Photograph of the scale formed on the membrane during the Brine B VEDCMD experiments.

Figure 15. Photograph showing the scale on the spacer used during Brine B VEDCMD experiments.

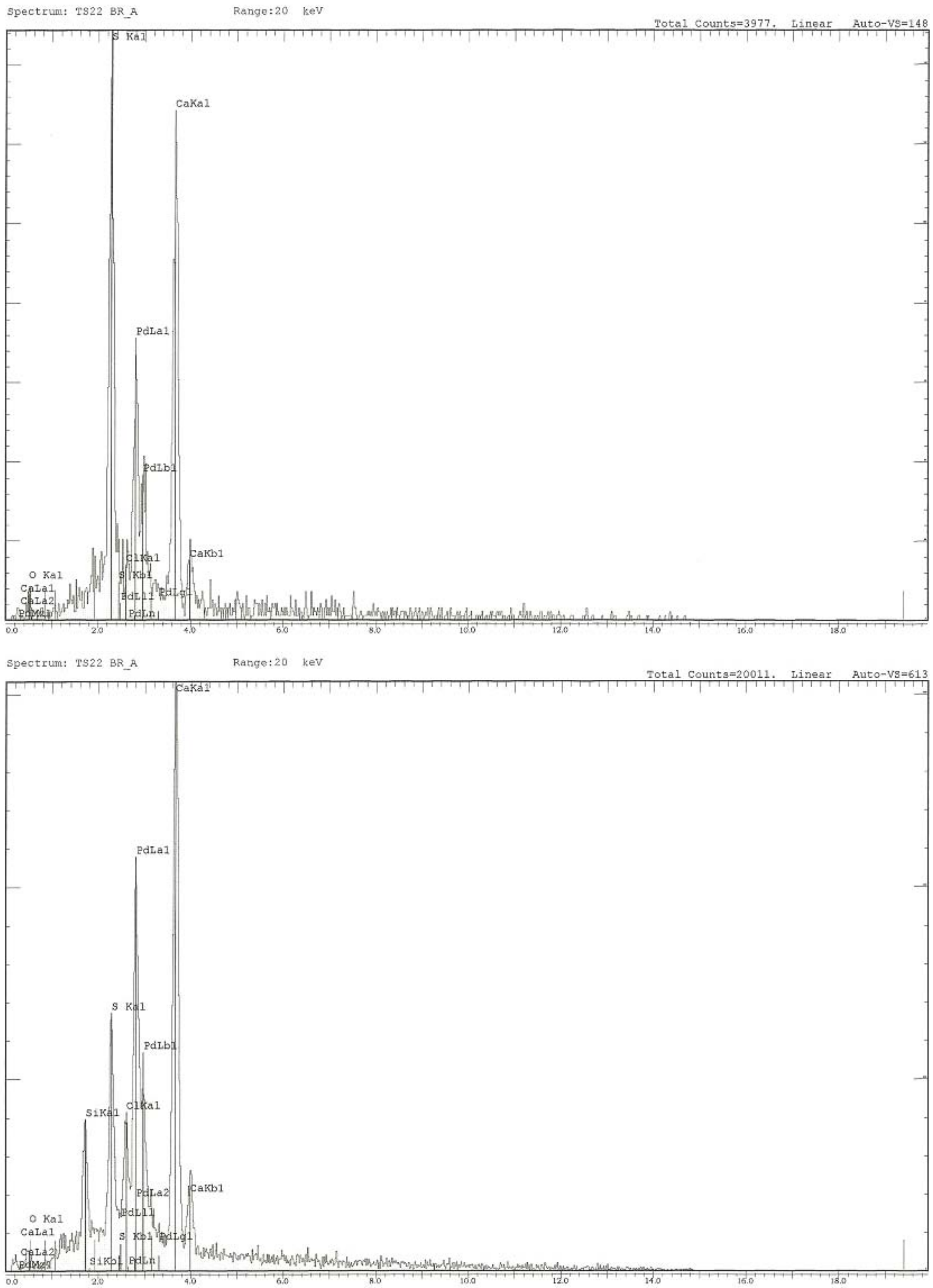


Figure 16. Chromatographs of the PTFE membrane surface at two spots.

Calcium sulfate scale also formed on the mesh spacer and on the feed surface of the CTA membrane during FO experiments.

Figure 17 **is an SEM image of the calcium sulfate scale on the CTA membrane surface. The scale on the feed spacer is shown in**

Figure 18.

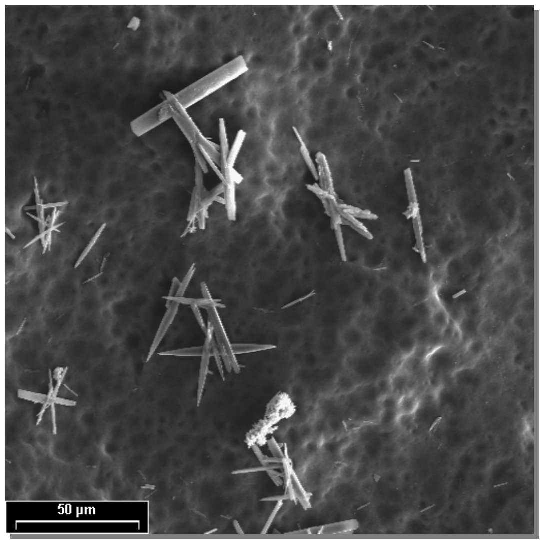


Figure 17. SEM image of calcium sulfate scale on the CTA membrane during Brine A FO experiments.

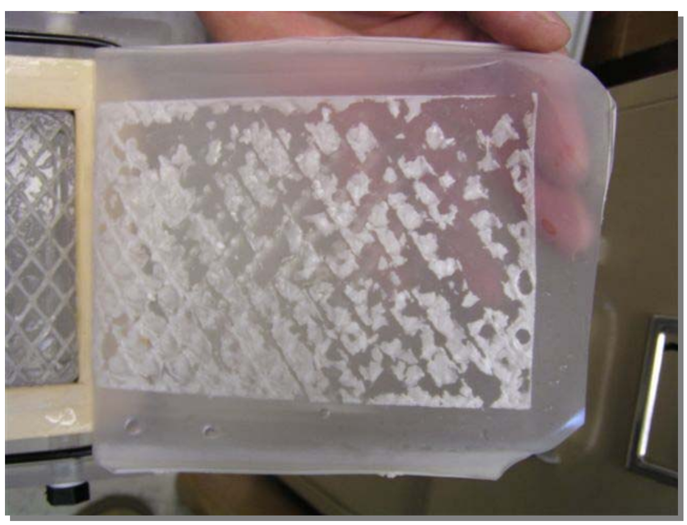


Figure 18. Photograph showing the loose scale on the CTA membrane and the feed spacer used during Brine A FO experiments.

4.3.2. Water quality

The major ions found in Brines A and B are shown in **Table 2**. Ion concentrations in the feed solutions are substantially larger than those found in the DS. This suggests that the CTA membrane exhibits good ion rejection (see Table 1).

Table 2. Ions found in Brines A and B.

TEST	METHOD	BRINE A RESULTS (mg/L)	BRINE B RESULTS (mg/L)	REPORTING LIMIT (mg/L)
Alkalinity	SM 2320 B	470	142	20
Hydroxide	SM 2320 B	< 7	< 7	7
Carbonate	SM 2320 B	< 12	< 12	12
Bicarbonate	SM 2320 B	573	164	25
Calcium	EPA 200.7	1200	870	25
Magnesium	EPA 200.7	370	410	500
Silica	EPA 200.7	135	77	25
Sodium	EPA 200.7	585	4600	25
Chloride	EPA 300.0	2800	8400	1
Sulfate	EPA 300.0	1300	1800	500

4.3.3. Process modeling

4.3.3.1 FO/RO

Pilot system modeling was performed to estimate the size, specific energy and power consumptions, and other operating cost components required for producing 1 MGD of drinking water from brackish water RO brine. These parameters were estimated for a total water recovery of 75% in the FO subsystem. Therefore, a membrane array was designed with ROSA that produced 1 MGD of purified water. **Figure 19** and Figure 20 illustrate the ROSA design and performance using a 16,000 ppm NaCl feed solution. Results in Figure 20a indicate that the specific energy for this design is 10 kWh/kgal (2.64 kWh/m^3) and the power requirement is close to 420 kW. The product water is highly purified (88 mg/l TDS); the DS produced by the RO system is approximately 50,000 ppm NaCl.

Using data from a published survey on the cost of desalination (for a 1,000 m^3/day RO plant) by Ettouney et al. [1], the reported energy cost was on the order of \$0.5/m^3, the chemical cost was approximately \$0.11/m^3, the unit membrane replacement cost was \$0.20/m^3, and labor cost was on average \$0.045/m^3. More current cost estimates generated by the ROSA program are provided in Figure 22. It is also important to note the further savings can be realized by incorporating an energy recovery device into the RO system.

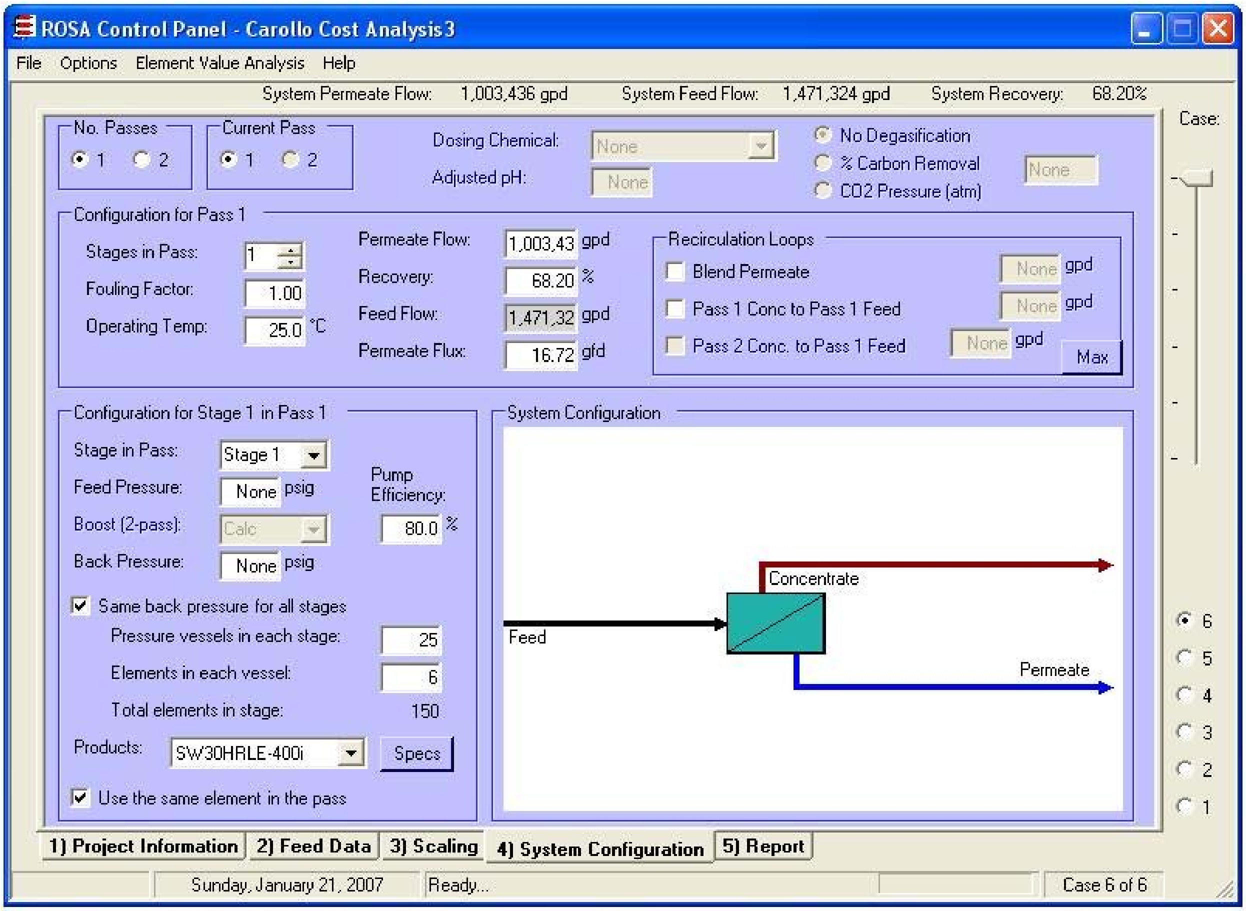


Figure 19. ROSA system configuration for RO desalination of 16,000 ppm NaCl feed solution. Permeate flow is 1 MGD at 70% water recovery.

Reverse Osmosis System Analysis for FILMTEC™ Membranes ROSA v6.1.3 ConfigDB u238786_51
Project: Carollo Cost Analysis3 Case: 6
, 1/21/2007

Project Information: RO Subsystem for DS Concentration and Water Recovery

System Details

Feed Flow to Stage 1	1471324.00 gpd	Pass 1 Permeate Flow	1003436.04 gpd	Osmotic Pressure:	
Raw Water Flow to System	1471324.00 gpd	Pass 1 Recovery	68.20 %	Feed	181.27 psig
Feed Pressure	752.97 psig	Feed Temperature	25.0 C	Concentrate	581.32 psig
Fouling Factor	1.00	Feed TDS	16000.01 mg/l	Average	381.30 psig
Chem. Dose	None	Number of Elements	150	Average NDP	360.04 psig
Total Active Area	60000.00 ft²	Average Pass 1 Flux	16.72 gfd	Power	418.40 kW
Water Classification: RO Permeate SDI < 1				Specific Energy	10.01 kWh/kgal

Stage	Element	#PV	#Ele	Feed Flow (gpd)	Feed Press (psig)	Recirc Flow (gpd)	Conc Flow (gpd)	Conc Press (psig)	Perm Flow (gpd)	Avg Flux (gfd)	Perm Press (psig)	Boost Press (psig)	Perm TDS (mg/l)
1	SW30HRLE-400i	25	6	1471324.00	747.97	0.00	467887.96	734.70	1003436.04	16.72	0.00	0.00	88.10

Pass Streams (mg/l as Ion)					
Name	Feed	Adjusted Feed	Concentrate	Permeate	
			Stage 1	Stage 1	Total
NH4	0.00	0.00	0.00	0.00	0.00
K	0.00	0.00	0.00	0.00	0.00
Na	6294.00	6294.00	19717.83	34.65	34.65
Mg	0.00	0.00	0.00	0.00	0.00
Ca	0.00	0.00	0.00	0.00	0.00
Sr	0.00	0.00	0.00	0.00	0.00
Ba	0.00	0.00	0.00	0.00	0.00
CO3	0.00	0.00	0.00	0.00	0.00
HCO3	0.00	0.00	0.00	0.00	0.00
NO3	0.00	0.00	0.00	0.00	0.00
Cl	9706.00	9706.00	30406.97	53.44	53.44
F	0.00	0.00	0.00	0.00	0.00
SO4	0.00	0.00	0.00	0.00	0.00
SiO2	0.00	0.00	0.00	0.00	0.00
Boron	0.00	0.00	0.00	0.00	0.00
CO2	0.00	0.00	0.00	0.00	0.00
TDS	16000.01	16000.01	50124.80	88.10	88.10
pH	N/A	N/A	N/A	N/A	N/A

Figure 20a. ROSA detailed report.

Reverse Osmosis System Analysis for FILMTEC™ Membranes — ROSA v6.1.3 ConfigDB u238786_51
Project: Carollo Cost Analysis3 — Case: 6
,
1/21/2007

Design Warnings

-None-

Solubility Warnings

-None-

Stage Details

Stage 1 Element	Recovery	Perm Flow (gpd)	Perm TDS (mg/l)	Feed Flow (gpd)	Feed TDS (mg/l)	Feed Press (psig)
1	0.18	10599.67	34.36	58852.96	16000.01	747.97
2	0.19	9241.23	47.89	48253.29	19507.14	743.94
3	0.20	7628.43	70.31	39012.06	24116.67	740.91
4	0.19	5852.35	109.13	31383.63	29961.63	738.67
5	0.16	4128.39	178.45	25531.28	36804.50	737.00
6	0.13	2687.37	304.24	21402.89	43869.27	735.73

Figure 21b. ROSA detailed report.

In FO, as with other membrane processes, the amount of recovered water depends largely on the surface area of the membrane. Therefore, the model was iterated to obtain the most efficient amount of membrane area that would produce approximately 75% water recovery from either Brine A or B. The FO membranes require the majority of the membrane area of the system because flux is lower in FO than in RO.

Ultimately, for the FO model, 352 FO vessels (704 elements) were arranged in parallel. Each element is 50 m^2 in membrane surface area. DS flowrate into each lead FO element is 3.5 L/min and Brine A flowrate into the second FO element is 10 L/min (Figure 23). Fourteen 1.5-hp centrifugal pumps were chosen to deliver feed water at 70 gpm each at less than 10 psi. Each pump will deliver water to approximately 25 FO vessels. The estimated conditions in each FO element are illustrated in Figure 23 for Brine A.

The power required to pump 1.34 MGD of water into the FO system by the 14 pumps is 8.7 kW assuming 620 W per pump. The specific energy requirement for the FO system is therefore calculated by dividing 8.7 kW by 158 m^3/hr water production rate, to give approximately 0.055 kWh/m^3. The results from the ROSA model indicate that 2.64 kWh/m^3 are required to produce 1 MGD using 150 RO membrane elements. The specific energy of the combined system is 2.7 kWh/m^3, with the FO system requiring approximately 2% of the energy. The majority of the specific energy is required by the RO membranes due to the high pressure operation.

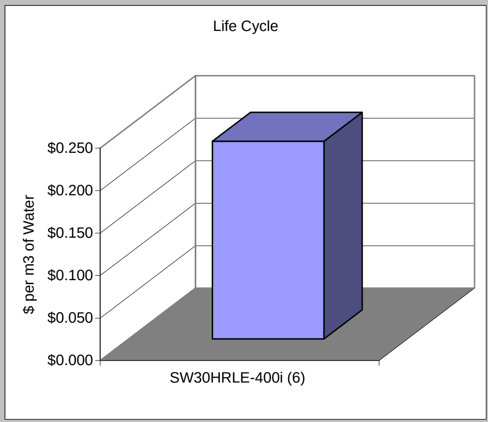
Similar results have been obtained for the FO/RO desalination modeling of Brine B, yet, scaling and concentration polarization effects are more significant in this kind of operation and further research needs to be done to establish a reliable model.

EVA (Element Value Analysis) for FILMTEC™ Products

FilmTec Corporation - A wholly owned subsidiary of The Dow Chemical Company

Yellow box = Input from ROSA
Green box = Input from the User
Blue font = Metric units
Red font = English units

	Carollo Cost Analysis3 Case 6
Product	**SW30HRLE-400i (6)**
System water production (m3/h)	**158.27**
System water production (GPM)	**696.83**
System recovery (%)	**68.2**
Pass 1 feed pressure (bar)	**51.9**
Pass 1 feed pressure (psi)	**753**
Pass 1 concentrate pressure (bar)	**50.7**
Pass 1 concentrate pressure (psi)	**735**
Pass 1 recovery (%)	**68.2**
Pass 1 energy recovery efficiency (%)	**0**
CAPEX	
PV	**25**
Total System Elements	**150**
$ per element	**$550**
$ per PV	**1000**
Element $	$82,500
PV $	$25,000
Capital	$107,500
Capital/(m3)	$0.0078
Capital/(kgal)	$0.0294
OPEX	
Years	**10**
Interest %	**10%**
Power	
Pumping power (kW)	**418.4**
Pump specific energy (kWh/m3)	**2.643650**
Pump specific energy (kWh/kgal)	**10.01**
Brine energy recovery, kWh/m3	0.00
Brine energy recovery, kWh/kgal	0.00
Net energy consumption; KWh/m3	**2.644**
Net energy consumption; KWh/kgal	**10.007**
Power cost ($/kWh)	**0.080**
Net energy cost ($/m3)	0.2115
Net energy cost ($/kgal)	0.80
Net energy cost ($/yr)	$293,217
Energy expense NPV ($)	$1,801,694
Energy expense ($/m3)	$0.2114920
Energy expense ($/kgal)	$0.8005844
Membrane Cost	
Replacement Rate	**13%**
Replacement Price $ per element	**$600**
Replacement Cost $	$11,700
Replacement Membrane NPV $	$71,891
Replacement memb expense ($/m3)	$0.0084
Replacement memb expense ($/kgal)	$0.0319
OPEX $ NPV	$1,873,585
OPEX per m3	$0.220
OPEX per kgal	$0.833
Life Cycle Cost NPV ($)	$1,981,085
Life Cycle Cost ($/m3)	**$0.233**
Life Cycle Cost ($/kgal)	**$0.880**



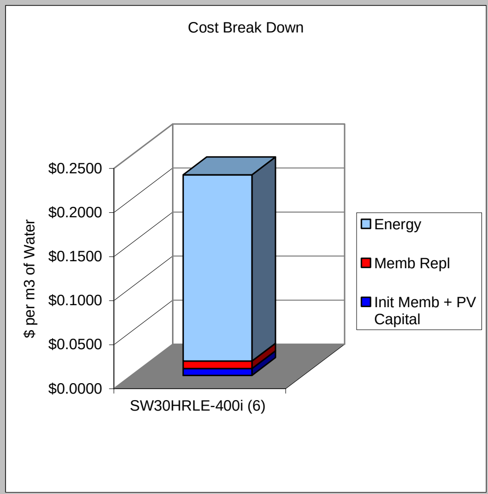


Figure 22. ROSA cost estimates for treatment of Brine A.

K=	0.00035948	l/min/m^2/psi
ΔA=	2.5	m^2
Total A=	50	m^2
C_{DSin}=	50	g/l TDS
Q_{DSin}=	3.5	l/min
C_{BWin}=	7.5	g/l TDS
Q_{BWin}=	10	l/min

NaCl DS/BW Brine System - Element No. 1

	π^{DS} psi	C_{DS} g/l	Q_{DS} l/min	Q_{BW} l/min	C_{BW} g/l	π^{BW} psi	
1	580.70	50.00	3.50	2.52	29.73	139.80	20
2	518.29	44.92	3.90	2.92	25.70	120.55	19
3	472.45	41.14	4.25	3.28	22.89	107.23	18
4	436.95	38.19	4.58	3.60	20.81	97.35	17
5	408.41	35.81	4.89	3.91	19.18	89.67	16
6	384.81	33.83	5.17	4.20	17.87	83.49	15
7	364.88	32.14	5.44	4.47	16.79	78.38	14
8	347.75	30.69	5.70	4.72	15.87	74.07	13
9	332.82	29.42	5.95	4.97	15.09	70.37	12
10	319.66	28.30	6.18	5.21	14.41	67.16	11
11	307.94	27.30	6.41	5.43	13.80	64.33	10
12	297.42	26.40	6.63	5.65	13.27	61.82	9
13	287.91	25.58	6.84	5.86	12.79	59.57	8
14	279.25	24.84	7.05	6.07	12.36	57.55	7
15	271.33	24.15	7.25	6.27	11.96	55.70	6
16	264.04	23.52	7.44	6.46	11.61	54.02	5
17	257.32	22.94	7.63	6.65	11.28	52.48	4
18	251.07	22.40	7.81	6.83	10.97	51.06	3
19	245.26	21.90	7.99	7.01	10.69	49.74	2
20	239.84	21.43	8.17	7.19	10.43	48.52	1

NaCl DS/BW Brine System - Element No. 2

	π^{DS} psi	C_{DS} g/l	Q_{DS} l/min	Q_{BW} l/min	C_{BW} g/l	π^{BW} psi	
1	240.12	21.45	8.16	7.19	10.43	48.54	20
2	235.02	21.01	8.33	7.36	10.19	47.40	19
3	230.23	20.59	8.50	7.53	9.96	46.33	18
4	225.72	20.20	8.67	7.69	9.75	45.33	17
5	221.47	19.83	8.83	7.86	9.55	44.39	16
6	217.45	19.48	8.99	8.02	9.36	43.50	15
7	213.64	19.14	9.14	8.17	9.18	42.67	14
8	210.02	18.83	9.30	8.33	9.01	41.88	13
9	206.58	18.53	9.45	8.48	8.85	41.13	12
10	203.30	18.24	9.60	8.62	8.70	40.41	11
11	200.17	17.96	9.74	8.77	8.55	39.74	10
12	197.19	17.70	9.89	8.92	8.41	39.09	9
13	194.33	17.45	10.03	9.06	8.28	38.47	8
14	191.59	17.21	10.17	9.20	8.15	37.88	7
15	188.97	16.98	10.31	9.34	8.03	37.32	6
16	186.45	16.76	10.44	9.47	7.92	36.78	5
17	184.02	16.55	10.58	9.61	7.81	36.27	4
18	181.69	16.34	10.71	9.74	7.70	35.77	3
19	179.45	16.14	10.84	9.87	7.60	35.29	2
20	177.29	15.95	10.97	10.00	7.50	34.83	1

Total Vessel Recovery:	74.78	%
Total Water recovered:	7.48	L/min
Total Water recovered:	2845	gpd
Average Flux:	4.49	L/m^2/hr

Figure 23. FO modeling of Brine A.

4.3.3.2 VEDCMD Modeling

Although MD processes have the substantial advantage of low impact of feed water salinity on membrane performance, there are no established large-scale applications of MD for seawater or brackish water desalination and no membrane companies that make membranes specifically for MD. Membranes that are currently used for MD are merely microfiltration membranes that have specific pore size and are made of hydrophobic materials.

For the purpose of estimating the cost of brackish water brine desalination, an off-the-shelf membrane from Microdyn (MD 150 CP 2N) (retail cost of \$5,000-\$7,000) that is 10 m^2 in membrane area and has 1,800 polypropylene capillary membranes installed in a tube-and-shell configuration was used. Under the conditions investigated in the current study, the Microdyn membrane can generate 2-4 L/m^2-hr at 40°C/20°C feed/permeate temperatures (approximately 750 L/day per membrane element) and 6-8 L/m^2-hr at 60°C/20°C feed/permeate temperatures (approximately 1,600 L/day per membrane element). In order to achieve this performance, the flowrate of both the feed and permeate streams into each membrane element must be on the order of 250 L/min. This implies that the single-element recovery under 40°C/20°C conditions is approximately 0.2% and under 60°C/20°C conditions the expected recovery with the Microdyn element is 0.4%.

Assuming that energy to heat the water is free, and further assuming 70% water recovery from Brine A, approximately 31,000 Microdyn elements would be required to produce 0.7 MGD fresh water from feed water at 40°C; approximately 15,000 Microdyn elements would be required to produce 0.7 MGD fresh water from feed water at 60°C.

In each case, cold water (at approximately 20°C) would have to be pumped to the permeate side of the membrane at a rate of 200 L/min in every membrane element, which converts to total pumping of 6,200 m^3/min (!) at a pressure of approximately 10 psig. This flowrate can be reduced if several membrane elements can be connected in series and have cooling stages between elements to maintain temperatures of approximately 20°C in the inlet of each element. On the feed side, assuming that 25 membrane elements can be connected in series (with re-heating stages in between), feed water at a rate of 310 m^3/min would have to be distributed between 1,240 vessels at a pressure of approximately 10 psig.

With these figures and comparing to the performance of FO for desalination of brines from brackish water desalination, MD is not a viable process for brackish water desalination. The drawbacks of the high energy required for the relatively low water flux and the very low single-module water recovery outweigh the advantages of very high salt rejection, chemically stability, ability to operate under very low pressure, and high life expectancy. Membranes designed specifically for MD applications would improve the outlook of MD for brine desalination. Additionally, if waste heat can be utilized to heat and re-heat the feed water (feed water looses heat during the process and requires re-heating to achieve high water recovery), MD would become a more viable alternative.

5. Conclusions

Bench-scale evaluations of VEDCMD and FO demonstrated that both were capable of concentrating a variety of brackish water RO brines. The range of recoveries achieved using VEDCMD to treat Brines A and B was 60-65%, with the higher recoveries obtained from the concentration of Brine B. The highest recovery obtained was from the concentration of Brine A using FO – 81% total water recovery was obtained in the experiments.

Similar water fluxes were obtained during VEDCMD at ΔT = 20°C and FO experiments. VEDCMD at ΔT = 40°C demonstrates a substantial increase in water flux over the others; however, the increased transport of scalants to the membrane surface caused the water flux to decline earlier. In all experiments, scale formed on the feed surface of the membranes substantially reducing water flux. Water flux was able to be recovered after cleaning of the CTA membrane, however, the cleaning method for the PTFE membrane was damaging to the active layer of the membrane.

Results from the FO and RO modeling indicate that RO system requires the majority of the specific energy and power requirements and the FO system requires the majority of the membrane area. Treating 1 MGD of brackish water RO brine at 70% recovery will result in 90-97% total recovery of brackish water and substantially reduce waste distribution loads.

6. References

[1] H.M. Ettouney, H.T. El-Dessouky, R.S. Faibish and P.J. Gowin, Evaluating the economics of desalination, Chemical Engineering Progress 98 (2002) 32-39.

[2] J. Glater and Y. Cohen, Brine disposal from land-based membrane desalination plants: a critical assessment, 2003, University of California, Los Angeles, Los Angeles.

[3] G.W. Miller, Integrated concepts in water reuse: managing global water needs, Desalination 187 (2006) 65-75.

[4] E. Yildiz, A. Nuhoglu, B. Keskinler, G. Akay and B. Farizoglu, Water softening in a crossflow membrane reactor, Desalination 159 (2003) 139-152.

[5] M. Mulder, Basic principles of membrane technology, 2nd, Kluwer Academic Publishers, Dordrecht, The Netherlands, 1997.

[6] B. Pilat, Practice of water desalination by electrodialysis, Desalination 139 385-392.

[7] P. Johannsen, R. Karlapudi and G. Reinhold, High pressure reverse osmosis for wastewater minimization and zero liquid discharge applications, Desalination 199 (2006) 84-85.

[8] R.Y. Ning, A. Tarquin, M. Trzcinski and G. Patwardhan, Recovery optimization of RO concentrate from desert wells, Desalination 201 (2006) 315-322.

[9] T.Y. Cath, V.D. Adams and A.E. Childress, Experimental study of desalination using direct contact membrane distillation: a new approach to flux enhancement, Journal of Membrane Science 228 (2004) 5-16.

[10] T.Y. Cath, V.D. Adams and A.E. Childress, Membrane contactor processes for wastewater reclamation in space. II. Combined direct osmosis, osmotic distillation, and membrane distillation for treatment of metabolic wastewater, Journal of Membrane Science 257 (2005) 111-119.

[11] K.W. Lawson and D.R. Lloyd, Review: membrane distillation, Journal of Membrane Science 124 (1997) 1-25.

[12] T.Y. Cath, S. Gormly, E.G. Beaudry, M.T. Flynn, V.D. Adams and A.E. Childress, Membrane contactor processes for wastewater reclamation in space: Part I. Direct osmosis concentration as pretreatment for reverse osmosis., Journal of Membrane Science 257 (2005) 85-98.

[13] A.C.M. Franken, J.A.M. Nolten and M.H.V. Mulder, Wetting criteria for the applicability of membrane distillation, Journal of Membrane Science 33 (1987) 315-328.

[14] H. Reulen, C.A. Smolders and G.F. Versteeg, Determination of mass transfer rates in wetted and non wetted microporous membranes, Chemical Engineering Science 48 (1993) 2093-2102.

Appendix B

GEO-PROCESSORS USA, INC. REPORT ON SAL-PTOC

(Draft 1 for Review/Geo/010707)

Eastern Municipal Water District

Evaluation of the SAL-PROC™ Process for Byproducts Recovery from RO Brines and Zero Liquid Discharge

January 2007

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Evaluation of the SAL-PROC™ Process for Byproducts Recovery from RO Brines and Zero Liquid Discharge

January 2007

1.0 EXECUTIVE SUMMARY

1.1 Overview and Study Objectives

The Eastern Municipal Water District (EMWD) as part of its Desalination Program is evaluating several water purification alternatives including pilot trials of RO membrane technology. The reject from RO desalination is a brine, which requires safe and cost effective disposal. Several brine reduction strategies are under consideration one being a Zero Liquid Discharge (ZLD) process developed by Geo-Processors that allows the recovery of dissolved salts as commercial byproducts in a process technology known as SAL-PROC™. This process selectively or sequentially removes dissolved salts from a saline stream such as RO brine as useful products and returns a large percentage of the flow stream for potable water use. It is a proprietary process that is generally applied in conjunction with RO membrane and other volume reduction technologies to minimize or eliminate the brine disposal needs.

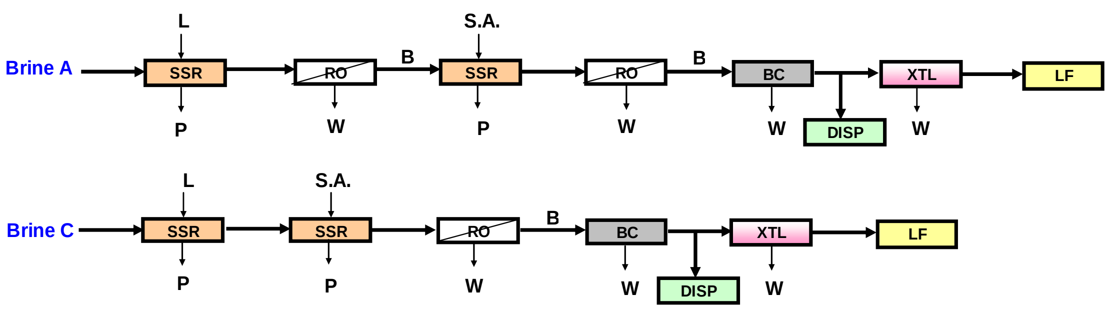
This report provides the results of a desktop pre-feasibility study performed by Geo-Processors USA, Inc. (Geo) on behalf of EMWD to assess the applicability and economic viability of the SAL-PROC™ process for selective recovery of byproducts and zero discharge. The recovered salts with beneficial uses become products which could generate a revenue stream to offset the treatment costs. One such product is Precipitated Calcium Carbonate (PCC), which is used as a premium coating and filling agent in paper manufacturing industry thus commanding high market demand and prices.

1.2 Methodology

The study was based on water quality parameters of two reference RO brine streams supplied by EMWD. It was assumed that brine flow rate would be 1 MGD and preference was given to a brine ZLD system with minimum space requirements and maximum flexibility in the design for future scale up.

The study was conducted by inputting the nominated concentrate flow and loadings into a desk top modeling software developed by Geo to identify a number of technically feasible ZLD process systems. Each of the process systems defined by the SAL-PROC™ model are comprised of two subsystems including one or more selective salt recovery steps that are linked with RO desalination, thermo-mechanical brine concentration and crystallization steps. The desk top modeling exercise enabled selection of the most appropriate ZLD process schemes as recommended options for the two reference brine streams. Figure 1 shows the process steps of the selected ZLD treatment systems and a variance in which the reject from brine concentrator is transferred and disposed at a commercial disposal facility (such as a POWT or a brine line) without further treatment. The selected ZLD systems utilize multiple reaction steps using lime and soda ash reagents to produce carbonated magnesium, PCC and a mixed salt. The systems also recover a large percentage of the flow as a potable water. A crystallizer is proposed in preference to crystallization ponds to minimize space requirements.

Figure 1. Schematic of brine treatment systems subjected to techno-economic evaluations.



SSR	Selective Salt Recovery
RO	Reverse Osmosis
BC	Brine Concentrator
XTL	Crystallizer
LF	Landfill
DISP	Commercial Disposal (Brine line or POWT)
L	Lime
S.A.	Soda Ash
W	Water (High Purity)
P	Salt Product

1.3 Conceptual Cost Estimates and Revenue Potential

Given the scope of study, Class V estimates of the Capex and Opex with an accuracy in the range of -30% to +50% were prepared to assist in economic analysis of the proposed brine treatment options. A summary listing of the cost estimates and revenue generation potential from sale of byproducts are given in Table 1.

Table 1. Conceptual cost estimates and projected revenue base for the recommended brine treatment systems.

Option	Annual Revenue, $m	Opex, $ m	Capex, $ m	Annual Cost*, $ m
Brine A - ZLD	3.29	2.93	16.72	4.51
Brine A - Partial Treatment & Commercial Disposal	3.29	3.79	14.06	5.12
Brine B - ZLD	5.26	5.10	28.31	7.77
Brine B - Partial Treatment & Commercial Disposal	5.26	8.17	15.17	9.60

** Sum of Opex and annualized cost at 7% over 20 years*

The operating and maintenance costs (Opex) were determined to be higher for options involving partial treatment; this reflects the high disposal cost of reject from brine concentrators. The capital costs (Capex) were influenced by the range of equipment and the salt load of brine stream. Using a 20 year life for capital and a 7% rate of return, the capital costs were annualized to a range between $4.51m and $5.12m for the less saline brine stream (6.5 g/L TDS), and $7.77m and $9.6m for the more saline brine (22.2 g/L TDS).

As indicated in the table above, the estimated revenue potential from the sale of commercial grade salts and fresh water is quite significant, with the most probable revenue base for ZLD options estimated to be $3.3 m for Brine A and $5.3 m for Brine C.

1.4 Benefit-to-Cost Ratio

Comparing potential income benefits to cost, in a ratio, provides a glimpse into the economics of the SAL-PROC™ based effluent treatment systems in terms of the potential for the offset of treatment costs by the revenue generated from the sale of byproducts. Such an analysis obviously excludes the consideration of other benefits from the waste treatment process to the overall project.

The benefit-to-cost ratio analysis was used to determine if the proposed brine treatment systems have the potential to either make money (ratio > 1), pay for themselves (ratio = 1) or be a cost (ratio < 1). The benefit-to-cost ratio is sensitive to the product income and the capital and operating costs. Table below summarizes the results of a simple sensitivity analysis conducted on the basis that all cost estimates were Class V and had an accuracy of -30% and + 50 %.

(Brine A)

Parameter	ZLD Option			Partial Treatment with Commercial Disposal		
	Best Case	Most Probable Case	Worst Case	Best Case	Most Probable Case	Worst Case
Benefit (Annual Revenue), $m	4.95	3.29	2.3	4.95	3.29	2.3
Annual Cost, $ m	3.16	4.51	6.77	3.58	5.12	7.68
Benefit-to-Cost Ratio	1.57	0.73	0.34	1.38	0.64	0.3
Net Cost, $ m	1.79	-1.22	-4.47	1.37	-1.83	-5.38

(Brine C)

Parameter	ZLD Option			Partial Treatment with Commercial Disposal		
	Best Case	Most Probable Case	Worst Case	Best Case	Most Probable Case	Worst Case
Benefit (Annual Revenue), $m	7.89	5.26	3.68	7.89	5.26	3.68
Annual Cost, $ m	5.44	7.77	11.66	6.72	9.6	14.4
Benefit-to-Cost Ratio	1.45	0.65	0.32	1.17	0.55	0.26
Net Cost, $ m	2.45	-2.51	-7.98	1.17	-4.34	-10.72

The possible range of benefit-to-cost ratio is moderately wide, indicating sensitivity to costs used and the need for additional work to improve accuracy. The results also confirm that Geo's ZLD options offer potentially cost-effective methods for sustainable management of the EMWD's brine streams.

1.5 Conclusions and Recommendations

From this pre-feasibility study it is concluded that:

(A) SAL-PROC™ process provides a technically feasible option for both dramatically reducing salt of the brine streams by recovery for beneficial use /sale. Ultimately technical feasibility would need to be complemented by confirmation of the process performance through pilot testing on the actual RO brine streams.

(B) The ZLD options based on SAL-PROC™ process operated in linkage with RO and other conventional volume reduction methods are potentially attractive economic options for brine disposal because of the possible income and additional capacity benefits. Although this study was not intended to prove the economic viability, the analysis clearly points to the potential for substantial income and additional capacity cost benefits from the proposed integrated process systems.

The following recommendations are offered:

- Pilot study using the EMWD brines and the ZLD processing option 2 to verify the results from the desk top modeling, assess the system performance and optimize treatment efficiencies.
- Compare the economics of SAL-PROC™ based ZLD processing options with the economics of the next best alternative for the disposal of EMWD brines.
- An integrated RO-Brine treatment facility to gain efficiencies.
- Further evaluation of beneficial uses of the mixed salts – including possible downstream processing for value adding.
- Marketing analysis development – confirm the local and regional interests.

2.0 INTRODCUTION

Desalination is a key element of an Integrated Water Resources Plan being implemented by the Eastern Municipal Water District (EMWD). As part of EMWD's Desalination Program the district has instigated a comprehensive assessment of various water purification alternatives including pilot trials using RO membrane technology. A byproduct from RO is a reject stream that is commonly known as RO brine or concentrate, which needs safe disposal. In view of the projected volume of brine that will be generated from the proposed desalination facilities at EMWD and expected limitations with its discharge to existing and planned brine lines, the identification and implementation of cost-effective and sustainable brine disposal methods has become an important component of EMWD's Desalination Program. Accordingly, EMWD in collaboration with and funding support from DWR and USBR has embarked on a progressive evaluation of several brine reduction strategies. Some of the brine management options currently under consideration involve prior concentration of brine before discharging it into a brine-line. This option although by definition is not a Zero Liquid Discharge (ZLD) process it allows significant volume reduction and hence a less requirement for brine-line allocation. Another option is the application of SAL-PROC™ process as a ZLD process or for regulated discharge. SAL-PROC™ is a proprietary process technology developed by Geo-Processors which enables integrated water and salt recovery for value adding. The technology has been extensively tested and licensed; it is available for commercial application in the States and elsewhere through licensing arrangements.

As part of the EMWD/USBR component of brine concentration studies, Geo-Processors, Inc. (Geo) was engaged in 2006 to perform a desktop pre-feasibility study of the application of SAL-PROC™ process for recovery of useful byproducts and zero discharge. For techno-economic evaluations water quality parameters for two reference brine samples generated from recent RO membrane pilot trials by EMWD were used. It is assumed that these samples represent the quality of brine streams expected from future commercial RO facilities and will be consistent in quality and supply. These brine samples are referred throughout this study as "Brine A" and "Brine C". They show major differences in their salinity levels and chemical make-up and are accordingly assumed unrelated to each other.

This report details the outcomes of the pre-feasibility study performed by Geo and provides approximation of important financial consideration that should be considered in decisions involving treatment of the EMWD brines in the recommended ZLD processing schemes that incorporate salt recovery steps. In view of limited information made available to Geo on the desalination component of the project, all components of the proposed processing schemes (including disposal of concentrated brine through the brine lines owned or leased by EMWD) are fully costed. Hence, the conceptual cost estimates provided in this report are indicative only and subject to change. Further, through further evaluations it may become evident that one or more components of the proposed brine processing schemes can be conveniently integrated with the overall water purification/recycling scheme to reduce the overall cost of the water production system. Therefore, it is recommended that this analysis be used only as an aid to evaluate water treatment alternatives.

2.1 Project Objectives and Scope of Work

The objectives of this desktop study were two fold:

(a) To evaluate the effectiveness of the SAL-PROC™ process to treat and recover valuable byproducts from two reference brine streams with significantly differences in their TDS salinity and chemical make-up; and,
(b) To assess economics of the recommended ZLD processing schemes for further consideration by EMWD as cost effective RO brine management options.

To meet the above objectives, during process design and techno-economical evaluations particular emphasis was directed towards the following:

- Minimum space requirements and maximum flexibility in the design and scale up of process components for locating the future ZLD processing plant in close proximity to the primary RO facility;
- Recovery of commercial grade salts from RO brines that offer immediate and near-term financial return and environmental benefits;
- Production of more finished water through the use of secondary RO and volume reduction processes in which SAL-PROC™ process improves RO membrane performance and increases water recovery rates;
- Effectiveness of combined salt recovery and volume reduction processes for reducing the salt load and the need for direct discharge of concentrated brine to existing and future brine-lines;
- Maximum recovery of useful products for accelerated payback to capital investment or to offset the operating costs;
- Reduction in operational and environmental footprint of the overall water production facility through integration of byproducts recovery, salt load reduction and volume minimization processes.

The scope of work defined by EMWD required evaluations based on desktop modeling only; it involved the following tasks, which were outlined in the terms of contract received from Carollo Engineers on February 16th, 2006:

(Task A. Technical Assessments)

- Using SAL-PROC™ process model assessment of various ZLD processing options and simulate and compare the performance of these options for selecting the most appropriate economic option compatible with local conditions
- Selection of a process design of a treatment train to be used to recover valuable byproducts.
- Preparation of preliminary mass balances including chemical usages, byproduct production rates and estimated power consumptions.

- Preparation of block flow and mass diagrams and tables to indicate each of the treatment trains and prepare written process description for each. Compile this information into the feasibility report (Task E).

(Task B. Product Market Review)

- Establish estimated product specifications and tonnages for each potential product from treatment of brine.
- Provide an overview of potential markets for each of the products and market demand for these products.
- Determine the approximate market price for each of the products to be used for cost-benefit analysis.

(Task C. Conceptual Design and Costing)

- Preparation of a basic conceptual design for the treatment train and prepare/define estimated plant footprint for each brine type for a plant capable of treating 1 MGD of each type of brine.

- Preparation of conceptual level capital and operating and maintenance cost estimates for treating 1 MGD of brine with the O&M to be annual costs including the estimates chemical, electrical, maintenance and labor costs.

(Task D. Benefit/Cost Analysis)

- Performance of a benefit/cost analysis based on the information generated as part of this study.
- Comment on the findings where applicable.

(Task E. Reporting)

Preparation and submission of a draft pre-feasibility, describing work undertaken in Tasks A through D for each of Brines A and C, and including all data and analytical results in the report. Report to also include an explanation of the technology, concluding remarks and recommendations. Following the review and update of the draft, a final report in Word format to be submitted to Carollo Engineers.

2.2 About Geo-Processors and SAL-PROC™ Technology

Geo is an environmental technology company with a business focus on development and supply of sustainable technology-based salinity control and water reclamation solutions for a range of industries including, oil/gas, energy generation, water production, mining/mineral processing, agricultural drainage and food processing. Company provides consulting services and technology-based solutions using its proprietary integrated water and byproducts recovery systems. Geo operates from two offices one in Los Angeles, California and another in Sydney, Australia to cater for its global projects. Geo's team of scientists, engineers and researchers are led by Dr. Aharon Arakel, an accomplished scientist/technologist with over 30 years industry, research and technology development experience. Geo's team is equipped with comprehensive field and laboratory testing facilities. These resources together with company's proven technologies are put to the task of addressing the needs of clients in the U.S. and elsewhere. Please refer to company's website (www.geo-processors.com) for further information on Geo's technologies and application areas.

In brief, **SAL-PROC™** represents one of the proprietary platform technologies developed by the company. It is probably the only commercially available technology for comprehensive treatment of saline and alkaline waters through sequential or selective recovery of commercial grade salts. The technology provides ZLD outcomes for most of the solutions offered to clients. Geo's environmental solutions provide two key outcomes for the water industry:

- Treatment of saline-alkaline wastewater to minimize storage and comply with regulatory requirements for effluent disposal, and
- Creating value through integration of water and byproducts recovery stages to offset the cost of waste minimization.

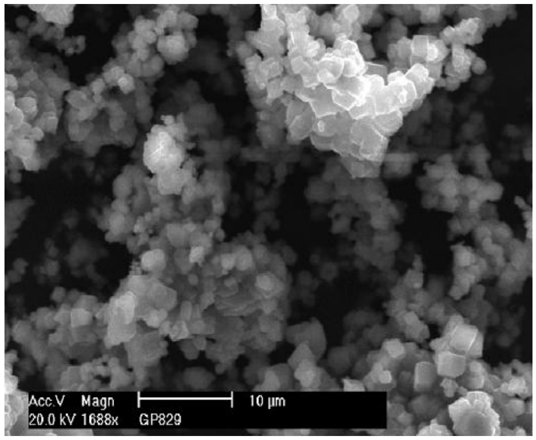
ROSP is an integrated processing scheme which combines reverse osmosis (RO) desalination and SAL-PROC™ processes for the recovery of freshwater and byproducts while achieving volume reduction targets. ROSP based processes are particularly suitable for ZLD processes where the make-up of feed water to desalination limits the rate of water recovery in the RO membrane or thermo-mechanical volume reduction steps (i.e., brine concentrators).

Another innovative technology developed by Geo is the Saline Effluent to Products Conversion (**SEPCON**) plants which use SAL-PROC™ technology. These plants can be fabricated in various sizes and configurations, either fixed or portable, to meet site-specific requirements.

Byproducts from SEPCON plants have been tested independently and confirmed having specifications comparable or better than the commercial grade products, thus offering potential for establishing a significant revenue base for either a return on capital investment or the offset of operation costs.

One byproduct of SAL-PROC™ with direct relevance to this project is Precipitated Calcium Carbonate (PCC). An example of a technical grade PCC produced in the SAL-PROC™ process is shown in Figure 1. This particular product was recovered from an alkaline produced water at a coalbed methane production facility in Queensland, Australia. In the case of alkaline feed waters, SAL-PROC™ may be used for not only the recovery of commercial grade PCC but also for improving the performance (ie, water recovery rate) and eliminating the need for acidification of the feed (and hence the release of CO_2 gas to air).

Figure 2. Scanning electron micrograph of a technical grade Precipitated Calcium Carbonate (PPC) product from SAL-PROC™ process showing the fine equi-granular nature of the PCC particles. The product was confirmed to be particularly suitable for manufacture of high grade paper products.



The following is a summary listing of selected major projects undertaken recently by Geo or are currently underway:

- ZLD process for management of saline effluent from Port Augusta City coal fired power station, South Australia, Australia. Detailed bench scale trials followed by large scale piloting of SAL-PROC™ technology for recovery of byproducts and ZLD process have since led to acquisition of a technology license by the Japanese client.

- Treatment of irrigation effluent to reduce the salt load and increase the life expectancy of a Salt Interception Scheme in Victoria, Australia. Project was funded by Murray-Darling Basin Commission and Goulburn-Murray Water Authority.

- ROSP technology application to management of saline discharge from dewatering bore fields in Cities of Wagga Wagga and Dubbo, New South Wales, Australia. A showcase urban salinity management and desalination based water re-use project, funded by the NSW and Australian Commonwealth governments.

- ZLD process based on ROSP process for treatment of produced water from a coal bed methane gas field, located in Queensland, Australia. The objective of the project was to remove the need for evaporation ponds and risks associated with the air release of CO2 gas from alkaline ground waters produced by a major gas company.

- Recovery of commercial salt from brines generated by Enhanced Oil Recovery (EOR) system from a major oil field in the Sultanate of Oman. The project initially under contract with Oman government and now the Occidental Oil, involves the design of salt recovery system with a capacity to treat up to 50,000 bbls/day of brine generated by EOR system currently being developed for this major oil field.

- Management of produced water from two major coal mining operations in South Africa and Australia owned by BHP Billiton. Feasibility study of the Australian site recently completed with preparations underway for pilot trials. South African project recently commenced.

- Several current projects in Australia and the U.S. involving feasibility study and piloting of SAL-PROC™ and ROSP technologies for application to integrated water and by-products recovery from RO brines of proposed oil field, food processing and municipal desalination projects.

3.0 METHODS AND ASSUMPTIONS

As per scope of the study the analysis reported in this document is based on desktop modeling only. It involved the definition, conceptual design and costing of the most appropriate brine processing options for EMWD's brines. Although no information was provided to Geo neither on the process steps nor the operational conditions of the primary RO it was assumed that the supply of brine to SAL-PROC™ process will be consistent in terms of water quality and the rate of supply. For the purpose of this desktop study a nominal 1 MGD flow rate was used for analysis.

3.1 Loadings and Chemical Composition

Table 2 presents the chemical make-up of the two RO brines (identified as "Brine A" and "Brine C"), received from Carollo Engineers for evaluation. Compositionally these brines are mildly alkaline-saline waters belonging to water type 2 in Geo's classification of the saline water types for treatment in the SAL-PROC™ process. Based on a 1 MGD flow rate the annual salt load of these brines are approximately 10,000 tpa for Brine A and 34,000 tpa for Brine C. These salt loads were considered in this study as one of the parameters for conceptual design and sizing of the SAL-PROC plants and follow-up cost estimations. The water analyses listed in Table 2 point to significant differences in elemental ionic ratios of major elements of the two brine streams. This together with the absence of background information on their origin comprised the reasons for preparing two separate cases in isolation from each other.

Table 2. Chemical composition of Brine A and Brine C received from EMWD for desktop techno-economic evaluations. All values are in mg/L except pH.

Parameter	Brine A	Brine C
Na	850	5,477
Ca	868	1,550
Mg	260	684
K	28.5	71
SO_4	1,215	3,338
Cl	2,250	10,599
HCO_3	850	316
Si	180	132
TDS	6,500	22,227
pH	6.61	-

3.2 Desk Top Modeling

The brine compositional data given in Table 2 were used for process simulation, using a computer modeling software developed by Geo for simulating treatment options under various TDS salinity and dissolved ionic (molar) ratios. Much of the treatment process is based on the proprietary SAL-PROC™ process developed by Geo.

The desk top modeling of Brines A and C generated three treatment options for each of the brine streams. The conceptual flow diagrams of the ZLD options for these brines are shown separately in Figure 3. As illustrated, all options start with one or two SAL-PROC™ process steps followed by treatment of the spent water in RO membrane unit(s) and then the reduction of the RO brine in a brine concentrator. These process steps form a unit operation identified as Subsystem A and then vary by either being treated in Subsystems B, C or D. As shown in Figure 3, Subsystem A for Brine A includes two secondary RO's (as against one RO in the case of Brine C); this enables maximum water recovery from a lower salinity feed water. An example of the conceptual flow diagram of linked operation of various subsystems in an integrated ZLD processing scheme is provided in Appendix A. In this case, the equipment arrangements are for Option 2 (Subsystems A+C) of Brine C. Equipment specifications for this specific example are listed in Appendix B.

The subsystems, schematically shown in Figure 3, are briefly described below.

Subsystem A

In the case of Brine A (Fig. 3A), it involves taking the brine stream through two "SAL-PROC" process trains that are linked alternatively with two brackish RO units; the reject brine from the second RO unit is then subjected to volume reduction using a brine concentrator. The first "SAL-PROC" process step involves mixing lime slurry with the Brine A. From the mixing process, the treated concentrate flows to a settling pond (approximately 0.3 acres). Periodically a beneficial product consisting of carbonated magnesium hydroxide [Mg(OH)2] will be harvested. The Mg(OH)2 product being a flocculating agent will also remove most of the dissolved Si and other metal ions from the brine that are detrimental to functioning of the RO membranes.

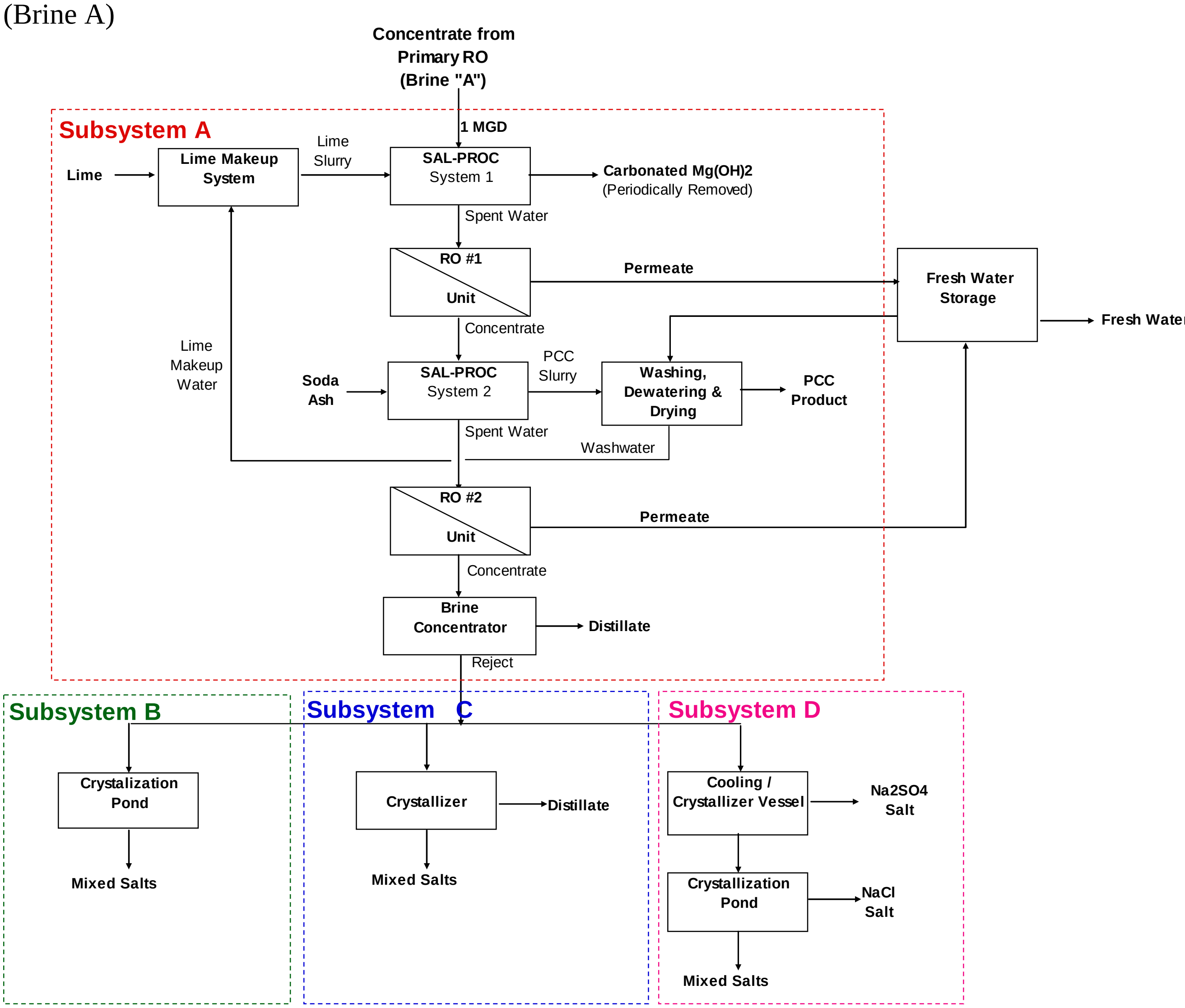
The supernatant from the Mg(OH)2 pond (also known as spent water) will be treated in the first brackish RO unit to recover permeate (at about 60% recovery) and the reject brine will flow into the second "SAL PROC" process for reaction with soda ash in one or more reactor tanks. The produced thin slurry from this process will be a Precipitated Calcium Carbonate (PCC) which will be thickened and then washed prior to dewatering in a filter press to produce a cake with 40% to 60% solids content. The filter cake may then be either processed via a dryer or sold as a filter cake product for use as a commercial-grade filler for paper and PVC manufacturing. It is estimated that 14 tons per day of PCC will be produced from treatment of 1 MGD of brine, type A.

The supernatant from the second "SAL-PROC" process will flow into a spent water tank where the water will be processed via a second RO unit. The RO unit will recover 60% of the water, thus the total freshwater recovery from the two RO units will be approximating 84% of the inflow (1 MGD). This level of water recovery is possible through the proposed staged separation of Mg ion (in the first SAL-PROC™ process) and Ca ion (in the second SAL-PROC™ process).

Byproducts from both stages will commercial grade. The remaining spent water, now about 16% of the original flow (which will be devoid of deleterious dissolved elements) will be further concentrated in a brine concentrator to reduce the volume to 0.023 MGD (which represents about 2.3% of the original feed volume) and then further treated in one of the three alternative subsystems shown in Figure 3 and described below. Apart from recovery of sodium salts as part of the process steps in Subsystem D, all these subsystems ultimately generate a mixed salt for landfill disposal. Although these mixed salts may be further treated or sold for downstream processing and salt recovery, for the purpose of this study no further treatment is assumed.

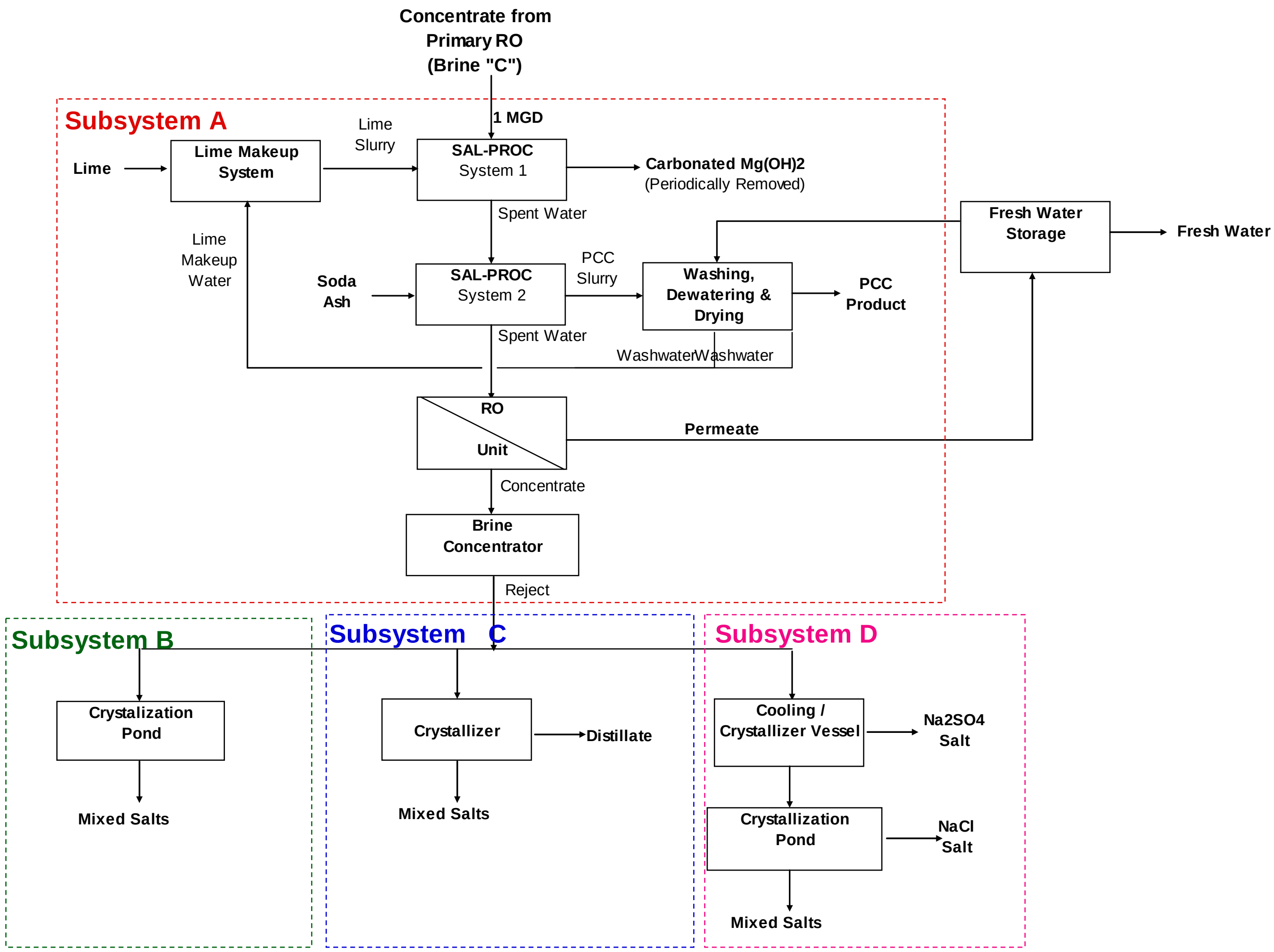
Figure 3. Conceptual flow diagrams of the ZLD options for Brines A and B.

(Brine A)



Option	Subsystems
1	A+B
2	A+C
3	A+D

(Brine C)



Option	Subsystems
1	A+B
2	A+C
3	A+D

In the case of more concentrated Brine C, as shown in Figure 3B, the Subsystem A involves the same process steps as Brine A with the exception that only one RO unit is deployed in this case for production of freshwater from spent water of the PCC production train.

The overall water recovery rate in this case is 67% of the base flow, and consequently the reject from brine concentrator (BC) is about 3.5 times larger in volume than the BC reject in the case of Brine A. It is suggested that as this reject is also characterized by elevated salt load, the selection of the follow-up subsystem for its disposal would be an important decision.

Overall, three alternatives (Subsystems B, C and D) are proposed for further treatment of the BC reject before final disposal; these are briefly described below. Although all are technically feasible, a decision on the preferred Subsystem will require consideration of several external factors, one important one being the short- and long-term capacity of the brine lines that may become available for the disposal of the BC reject stream. Note that it is assumed that all three subsystems for achieving ZLD will require a dedicated landfill for safe disposal of the mixed salts. For this reason the process design and cost analyses provided in the following sections of this report include a variance loosely named as “ Partial Treatment and Commercial Disposal”. This latter option denotes to a process train by which the BC reject from Subsystem A is directly transferred by a pipeline and disposed off at a commercial wastewater facility. Note also that the conceptual cost estimates for such disposal (whether via a POWT or a brine line) have all assumed commercial rates in order to enable a common base for first-order cost benefit comparisons.

Subsystem B

In both cases of Brines A and C the reject stream from BC unit will flow into a purpose-built crystallization pond to produce a mixed salt which is then harvested and transferred to a dedicated landfill. The mixed salt tonnages from the treatment of 1 MGD of Brine A and Brine C will be about 9000 tpa and 33,000 tpa, respectively. These tonnages will need about 4 acres and 17 acres of land for establishing landfills. An alternative disposal arrangement may involve allocating larger areas for the crystallization pond for subdividing it into several parts with the downstream parts progressively used as landfill site.

Subsystem C

In this subsystem dry salt production for landfill disposal is achieved by using a thermo-mechanical crystallizer in order to reduce the costs, risks and delays associated with the use of solar crystallization ponds.

Subsystems D

In this subsystem the BC reject is subjected to separate steps of cooling and evaporation to recover sodium sulfate [Na_2SO_4] and then sodium chloride [NaCl] salt with the end residual bittern converted to dry mixed salt for landfill disposal.

3.3 Selection of a Preferred Processing System

Considering the attributes of each of the ZLD processing options described above, with reference to project criteria and assumptions, Option 2 (Subsystems A+C) was selected as the best choice for both Brine A and Brine C stream for follow-up process design and economic evaluations. The variance involving the disposal of BC reject via a commercial facility was also included for first-order costing and comparison. This variance may become particularly attractive to EMWD if the use of expensive and energy intensive crystallizers can be eliminated through transfer and discharge of BC reject directly to an existing or future brine line.

It should be noted that although the selection of Option 2 was largely influenced by the inherited space limitations at EMWD, it also offers some added advantages that are unmatched by other options. While it is correct that the crystallizers are expensive to operate, they remain a better choice when the overall daily flow rates are considered and the costs involved with setting up and operating solar crystallizer ponds are factored in. Also, whereas the subsystem D is a common process in many salt harvesting operations it will need significant additional capital outlay to produces two low-value salt streams. This system will be more appropriate for large scale salt harvesting operations located close to their end markets for bulk supply.

Note also that in the case of conventional ZLD schemes up to 50% of the solids requiring landfill disposal are commonly generated from lime softening step. Although the SAL-PROC™ based ZLD systems will be devoid of such a high solids stream, considering the scope of this study no consideration is given to demonstrate this advantage offered by the proposed processing scheme.

3.4 Process Flow Diagram of the Selected Treatment Options

Using the conceptual flow diagrams for Subsystems A and C (Figure 2), two simplified process block flow diagrams (shown in Figure 3) were prepared for each of the Brines A and C to form the basis of follow up cost evaluations. These flow diagrams were then upgraded to include material balances and flow rates through each process step for treatment of 1 MGD brine using Brine A and Brine C water quality parameters (Figure 4). The options for commercial disposal of BC brine are also included in Figure 4 to give an indicative base for economic evaluation of both the ZLD and partial treatment/disposal options.

Note that the BC reject is essentially comprised of sodium sulfate, calcium sulfate and sodium chloride salts and thus amenable to further treatment for the separation and recovery of additional byproducts. However, in view of relatively small volumes involved and the low market value of the sodium sulfate salt ($75-$110/ton), the production of mixed salts for achieving ZLD or commercial disposal were considered more appropriate for the purpose of this study. These options for value adding should be investigated during a pilot study which is recommended as a next step activity. Also, the conceptual flow diagram and equipment arrangements presented in Appendix A show several thickening, filtration and chemical dosing components that could be possibly optimized. These types of improvements in process steps and equipment selection/arrangements shall also be investigated during the pilot study in order to reduce the footprint and cost of the overall treatment system.

Figure 4. Simplified process block flow diagrams of the ZLD processing schemes selected for Brines A and B.

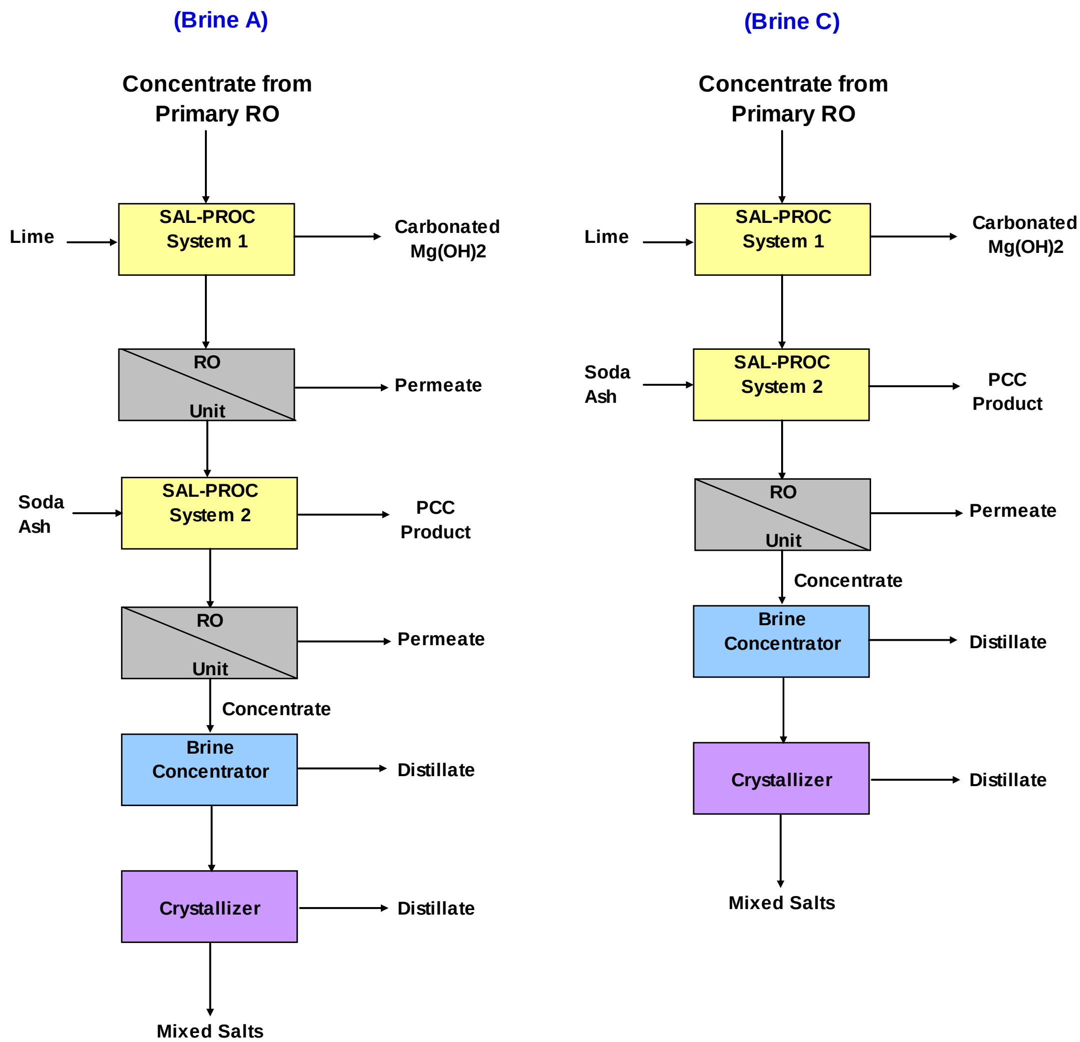
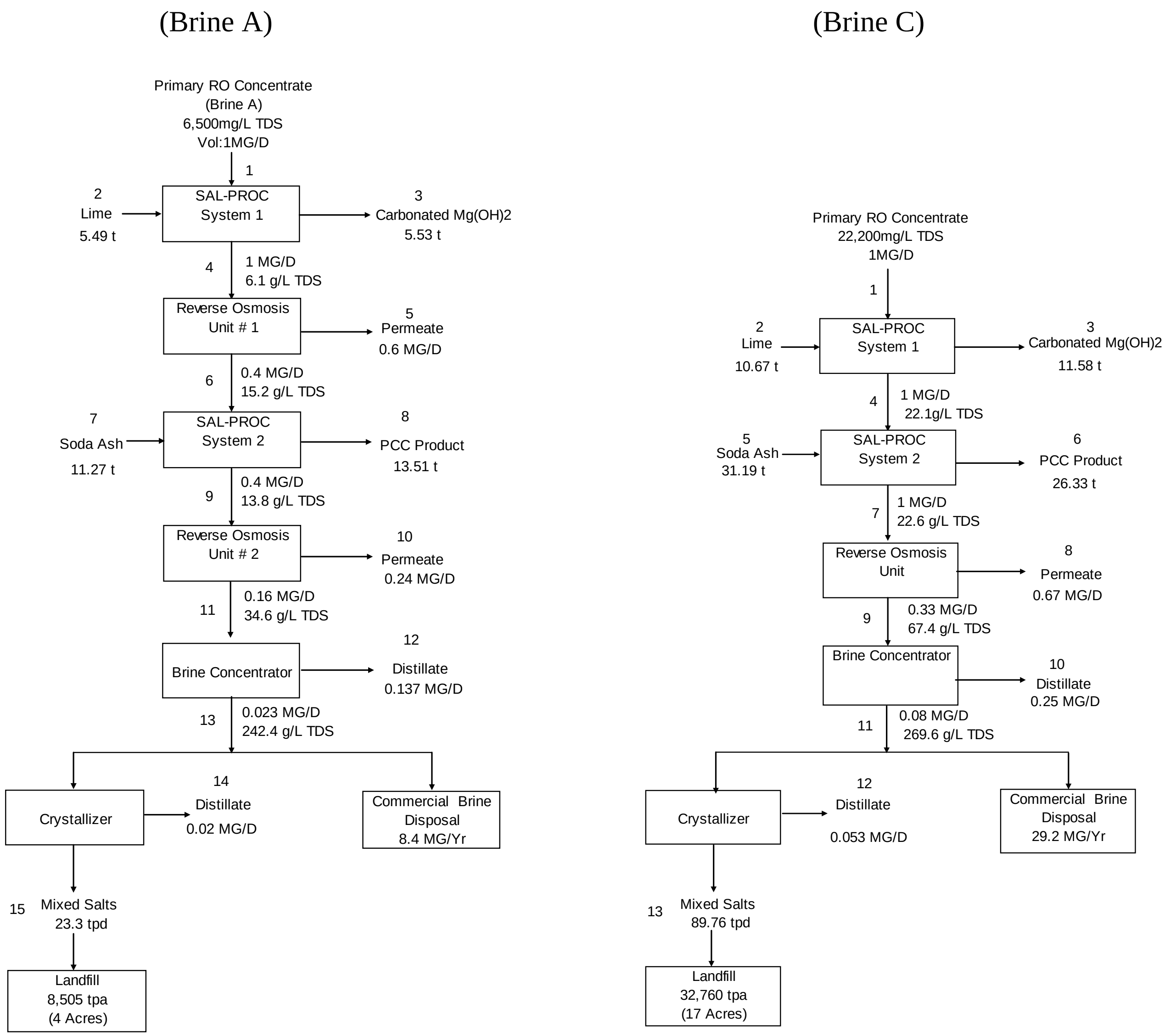


Figure 5. Process flow diagrams showing the material balances and flow rates for the selected ZLD options and variances for Brines A and C.

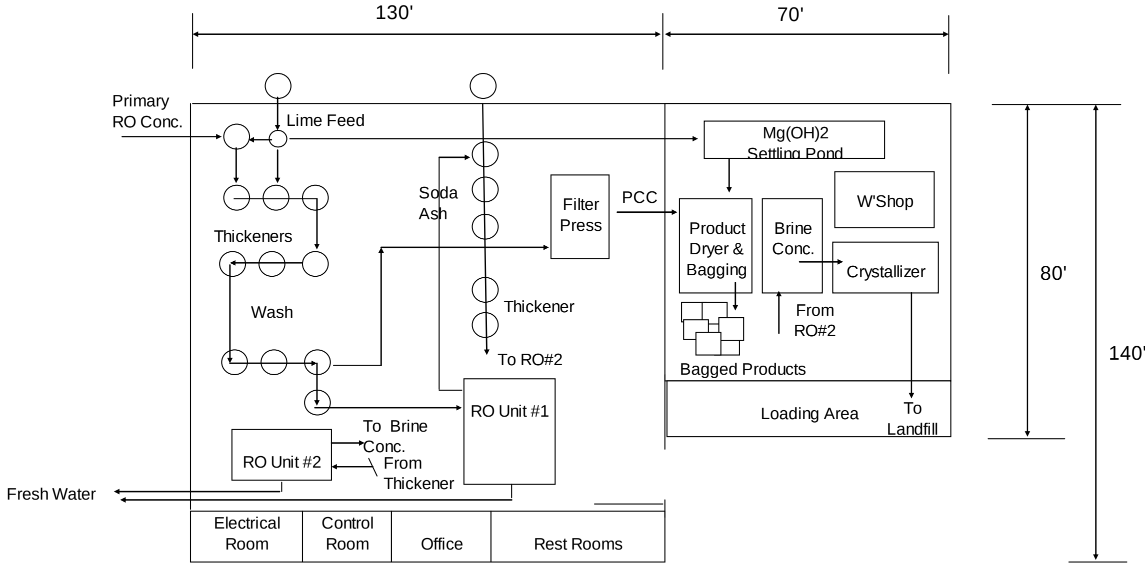


3.5 Conceptual Layout

Although the sites for either of Primary RO or brine treatment facility have not been selected, it is envisioned that that the brine treatment facility will occupy about 0.6 to 0.7 acre of land and be located at or near the EMWD's primary RO facility. This space requirement assumes 1 MGD brine flow through the brine treatment facility but excludes space for landfill ponds which will require 4 acres and 17 acres for disposal of mixed salts from treatment of Brine A and Brine C, respectively If a significant increase in flow rate is envisaged (thus requiring additional unit processes and chemical handling facilities), the brine treatment site may be located at an industrial area.

Figure 6 is the conceptual layout for ZLD processing of Brine A used to prepare the capital cost estimates. The layout is based on a number of inter-connected process equipment located in open space and pre-engineered buildings, filter press and product dryer/bagging facilities that are in enclosed space. The total area requirement is estimated to be 0.63 acre.

Figure 6. Conceptual layout of Brine A treatment equipment. Refer to Appendices A and B for equipment list and specifications.



4.0 MARKET ANALYSIS – OVERVIEW

As shown in Table 2, the EMWD brine streams are relatively elevated with respect to sulfate and calcium ions. Thus, during the process design consideration was given to identifying an appropriate ZLD processing scheme that satisfies the followings:

(a) improved water recovery rate for the RO units,
(b) high volume reduction rate for operation of the brine concentrators, and
(c) where possible production of high-value commercial grade product streams with known local market demand or use possibilities.

The Subsystem A was selected from a number of alternatives to achieve the above technical and product disposal objectives. This subsystem involves the removal of calcium ion to avoid saturation and subsequent precipitation of gypsum ($CaSO4.2H2O$), hemihydrate ($CaSO4.1/2H2O$) or glauberite ($CaSO4.Na2SO4.H2O$) salts in RO and BC steps of the process train. These salts are widely known as having detrimental effects on the performance of both RO membranes and brine concentrators. Furthermore, the selective recovery of the carbonated $Mg(OH)2$ in the first process step enables to remove most of the dissolved metal elements (including Si) by flocculation thus reducing the potential adverse impact of such metals on the performance of the RO membranes. Consequently, the selection of the proposed ZLD option was primarily for achieving these technical objectives (increased water recovery and reduced footprint) rather than solely for recovery of the highest value byproducts.

The byproducts from implementing the ZLD Option 2 are comprised of four streams namely fresh water, carbonated magnesium hydroxide, PCC and mixed salts; a brief overview of market demand for these products are given below.

4.1 Water

A key product from the brine treatment system is the recovery of fresh water. Based on information available to Geo the price for water is assumed $3.61 per 1,000 gallons. Using this price the potential revenue generated annually from the recovery of water from both brine streams in the recommended ZLD option will be around $1.3 m.

4.2 Precipitated Calcium Carbonate (PCC)

PCC is the main product stream from the proposed ZLD option in terms of both revenue generation potential and market demand.

Globally, over 75% of PCC produced is used by paper industry which has been the main factor behind the rapid growth in production of PCC since the mid-1980s. The use of PCC in paper is forecast to rise by an average of 5% a year and reach 7.2Mt by 2010. The development of satellite PCC plants by MTI Corporation has enabled the North American papermakers to move away from acid papermaking by providing access to high brightness calcium carbonate fillers.

Apart from paper industry PCC has many different uses within each industry but in order to sell PCC into a particular industry it must have specific traits for that specific industry. The use of PCC in plastics is closely linked with demand for PVC and hence building industry. The material helps increase impact strength for rigid polyvinyl chloride (PVC), bulk moulding compounds, phenolics, polyethylene and polypropylene materials.

Another major use of PCC is for paints to increase consistency and as a binder with pharmaceuticals and healthcare products. Other major uses are in special inks and sealants, and dyes.

The price of PCC varies greatly according to individual requirements, quality and the country of manufacture. For example, quality coated PCC imported from China is \$700-\$800/tonne, whereas the higher quality PCC imported from Japan is \$1200-1600/tonne. The PCC with average particle sizes less than 1.5 micron increases brightness and ink receptivity of paper and therefore command premium market prices. Based on available market demand and pricing structure for PCC with comparable specifications to that of PCC from the SAL-PROC™ process (see Table 3), for the purpose of this study a sales value of \$300/tonne is used in this analysis. Based on this pricing and assuming 100% sales achieved the potential revenue generated annually from recovery of PCC in the recommended ZLD option will be around \$1.48 m for Brine A and \$2.9 m for Brine C.

Table 3. Typical product specifications for technical grade PCC recovered previously in the SAL-PROC™ process.

Chemical Composition	
CaCO3%	98%
Magnesium Hydroxide %	0.58%
Iron and Aluminum Oxide %	0.30%
Silica Oxide %	0.90%
Sodium Chloride %	0.27%
pH (saturated solution)	11
Physical Features	
Inferred Particle Size Range (microns)	0.5 - 1 um
Inferred Average Particle Size (microns)	0.8 um
Particle Morphology	Discrete granular
Color	Bright white

4.3 Carbonated Magnesium Hydroxide

This product is used widely in the treatment of water and wastewater. It is used primarily for neutralization of acidic effluents and for the removal of dissolved heavy metals. The major advantage of magnesium hydroxide is that it is a pH buffer and treated wastewater will not exceed a pH value of 10 even if excess magnesium hydroxide is added. In contrast, the addition of lime, caustic soda or soda ash can raise the pH well over 12 causing a potential environmental violation. As indicated earlier, in the context of this project's objectives one major advantage from recovery of carbonated magnesium hydroxide relates to simultaneous removal of Si and other detrimental metals with magnesium hydroxide in high pH conditions. The spent water after magnesium hydroxide separation is this less amenable to causing fouling effects in the RO membranes.

This product may also be used in many other industries including fire retardants, pollution control, air CO2 capture, magnesium metal production, animal stock feed, light weight concrete and building material. It is also a major flux and alloying agent in the aluminum and steel industries. Price of this material varies widely; the quoted prices vary from $350 to $1200 per ton depending on it physical and chemical characteristics, including reactivity level for specific application. Prices obtained recently from a major supplier in Texas ranged from $45O to over $800 per ton. Geo has extensive experience with the production of magnesium hydroxide in the SAL-PROC™ process. The technology has been extensively tested and already licensed. Based on this previous experience for the purpose of this study a sales value of $250/tonne is used in this analysis. Using this pricing and assuming 100% sales achieved the potential revenue generated annually from the recovery of magnesium hydroxide will be around $0.51 m for Brine A and $1.06 m for Brine C.

4.4 Mixed Salts

Mixed salts will consist of sodium sulfate and sodium chloride with some impurities including silica, potassium and nitrates. As indicated earlier, with further refinement the mixed salts may be separated and possibly used in soap/detergent industries or as a de-icer. However, this analysis for ZLD processing options assumes that the mixed salts will generate no revenue as they will be disposed off at an offsite landfill.

Estimated quantities and values for the byproducts described above are presented in Table 5.

5.0 CONCEPTUAL ESTIMATES

5.1 Material Balances

Table 4 lists the solids (in tons) and flows (in MG) based on 1 MGD flow rate that correspond to the numbered steps illustrated in Figure 5 for each of Brine streams A and C. A complete listing of the total mass and flows for 1MGD flow and corresponding ionic balances are given in Appendix C.

Table 4. Total mass and flows for Brines A and C, based on 1 MGD flow rate.

(Brine A)

Step	Description	Total Mass (T) & Flow (MG)
1	Feed water	1
2	Lime	5.49
3	Carbonated Mg(OH)2	5.53
4	Spent water	1
5	RO1 Permeate	0.6
6	RO1 Reject	0.4
7	Soda ash	11.27
8	PCC product	13.52
9	Spent water	0.4
10	RO2 permeate	0.24
11	RO2 reject	0.16
12	BC water	0.137
13	BC Reject	0.023
14	Crystallizer Water	0.02
15	Mixed salts	23.3

(Brine C)

Step	Description	Total Mass (T) & Flow (MG)
1	Feed water	1
2	Lime	10.67
3	Carbonated Mg(OH)2	11.58
4	Spent water	1
5	Soda ash	31.19
6	PCC product	26.33
7	Spent water	1
8	RO permeate	0.67
9	RO reject	0.33
10	BC water	0.25
11	BC Reject	0.08
12	Crystallizer water	0.053
13	Mixed salts	89.76

5.2 Variance in the Salt Loads Produced and the Projected Revenue Base

The estimated daily and annual quantities and the costs/values for reagents and product streams (based on 365 days-year operation) for a processing plant with 1 MGD flow rate are provided in Table 5. As shown in this table, given the same flow rate for both brine streams, the revenue base is directly influenced by chemical make-up and to a lesser degree by TDS salinity. The salinity level also affects water recovery rate. The variance in total annual revenue from a low of $3.29m in the case of Brine A to a high of $5.26m in the case of Brine C (assuming that revenue from water recovery steps is nearly the same for both) shows a strong dependence of the revenue on the chemical make-up. As it is commonly known, the methodology and chemicals applied in the pretreatment of feed to primary RO have a major influence on the ultimate make-up of the brine streams. From previous experience where the SAL-PROC™ process is applied as part of the pretreatment train for high salinity feed waters to a primary RO, the economics of the ZLD schemes incorporating salt recovery steps show significant improvement. This particularly applies to high salinity ground waters with elevated Si content in which case the carbonated Mg(OH)2 production will advantageously perform as a pretreatment step and for reducing the use of lime for feed softening. The use of Mg(OH)2 will also significantly reduce the costs associated with dewatering of sludge from lime softener units. However, as this study specifically deals with the RO brine management aspects, no further evaluation on the added benefits that may arise from application of SAL-PROC™ technology to the total water recovery system is provided.

Table 5. Estimated quantities and costs/values for the reagents and products . A 365 days-year operation is assumed.

(Brine A)

Description	Reager				Products			
	T/D	T/Yr	Price $/T	$ M/Yr	T/D	T/Yr	Price $/T	$ M/Yr
Lime	5.49	2,005	75	0.15				
Soda Ash	11.27	4,115	135	0.56				
Carbonated Mg(OH)2					5.53	2,020	250	0.51
PCC					13.52	4,935	300	1.48
Fresh Water					0.99MGD	361MGD	$3.61/1000G	1.3
Mixed Salt					23.3	8,505	0	0
			Total:	**$0.71 m**			**Total:**	**$3.29 m**

(Brine C)

Description	Reagents				Products			
	T/D	T/Yr	Price $/T	$ M/Yr	T/D	T/Yr	Price $/T	$ M/Yr
Lime	10.67	3,895	75	0.29				
Soda Ash	31.19	11,385	135	1.54				
Carbonated Mg(OH)2					11.58	4,225	250	1.06
PCC					26.33	9,610	300	2.9
Fresh Water					0.97MGD	354MGD	$3.61/1000G	1.3
Mixed Salt					89.76	32,760	0	0
			Total:	**$1.83**			**Total:**	**$5.26**

5.3 Capital Cost Estimate

To provide a common base for comparative assessment of the economics of ZLD options and their variances, the estimates of capital expenditure (Capex) were prepared for Brines A and C separately. Considering the scope of this study these estimates are Class V estimates. A Class V estimate (as defined by the American Association of Cost Engineers) has an accuracy of -30% to +50%.

The estimates are based on the following assumptions:

- Land is owned by the EMWD
- The brine treatment plant will be financed and operated as a stand-alone facility
- Buildings will be brick veneer
- Electrical power and gas supply to site battery limit
- Ponds and landfills will be lined according to local guidelines
- Costing for piping between the main RO system and the brine treatment plant is not included
- Commercial rates apply for the disposal of BC reject via a commercial disposal facility, whether a brine line or a POWT facility. Where disposal to a brine line can be achieved at a significantly reduced rate or through special arrangements, the cost estimates and economic analysis presented in this report will require refinement to reflect such reduced disposal costs.

The conceptual capital cost estimate for the recommended treatment and disposal options for both Brines A and C are given in Table 6. As indicated, the estimated capital cost of 1 MGD Brine A treatment in the recommended ZLD processing scheme is $16.7 m and for the variance involving partial treatment with commercial disposal is $14.1 m. As shown in Table 6, the comparable estimated capital cost of treating the same volume of Brine C are $28.3 m and $15.2 m, respectively.

From the breakdown of the cost items, it can be seen that the largest cost differentiator between the ZLD and partial treatment options for both brine streams is the crystallizer used in the ZLD option for producing mixed salts for disposal at a dedicated landfill. In the case of partial treatment, this crystallizer is replaced by transferring the BC reject to an off-site commercial disposal site (or a brine line as the case may be). However, as shown in Table 7, this capital saving is at the expense of a dramatic increase in the operating expenses related to the high cost of brine disposal at a commercial facility.

In comparing the Capex of the ZLD option of Brine A and Brine C, a major cost difference relates to the use of two RO units in the case of Brine A, which will be necessary to achieve a water recovery rate. This is in contrast to using one RO unit for Brine C which is several fold more concentrated than the Brine A.

Table 6. Conceptual capital cost estimates for the recommended ZLD and partial treatment with commercial disposal options**.**

(A) Brine A; ZLD Processing Scheme

Item	Ref	Quantity	Material Cost $ m	Delivery Cost (2.5% of Material) $ m	Installation Cost (15% of Material) $ m	Capex $ m
Concrete & Structural Steel			0.3	0.007	0.045	0.35
Buildings	$200/sq.ft.	3,000sq.ft.	0.6	0.015	0.09	0.63
Dedicated Landfill	Capex 0.22 M /acre, includes Haulage	4 Acres		0	0	0.88
SEPCON Plant Equipment	Capex $3.1 m for 1 MGD Flow	Total Treatment System	3.1	0.08	0.46	3.64
RO Desalter #1	Flow rate 1 MGD Capex $1.1 m for 1 MGD	1	1.1	0.03	0.17	1.30
RO Desalter #2	Flow rate 0.24 MGD Capex $3.3 m for 1 MGD	1	0.8	0.02	0.12	0.94
Brine Concentrator	Flow rate 0.16 MGD Capex $6.8 m for 1 MGD	1	1.0	0.03	0.15	1.20
Crystallizer	Flow rate 0.023 MGD Capex $43 m for 1 MGD Flow	1	0.85	0.02	0.13	1.00
Instrumentation	10% of Treatment Equipment					0.68
Mechanical/Piping	5% of Treatment Equipment					0.34
Electrical	7% of Treatment Equipment					0.48
Pumping Plant & Pipelines						0.50
Sub-total A						**11.94**
Contingency	20% of Sub-total A					2.39
Admin	20% of Sub-total A					2.39
TOTAL $ m						**16.72**

(B) Brine A; Partial Treatment with Commercial Disposal

Item	Ref	Quantity	Material Cost $ m	Delivery Cost (2.5% of Material) $ m	Installation Cost (15% of Material) $ m	Capex $ m
Concrete & Structural Steel			0.3	0.007	0.045	0.35
Buildings	$200/sq.ft.	3,000sq.ft.	0.6	0.015	0.09	0.63
SEPCON Plant Equipment	Capex $3.1 m for 1 MGD Flow	Total Treatment System	3.1	0.08	0.46	3.64
RO Desalter #1	Flow rate 1 MGD Capex $1.1 m for 1 MGD	1	1.1	0.03	0.17	1.30
RO Desalter #2	Flow rate 0.24 MGD Capex $3.3 m for 1 MGD	1	0.8	0.02	0.12	0.94
Brine Concentrator	Flow rate 0.16 MGD Capex $6.8 m for 1 MGD	1	1.0	0.03	0.15	1.20
Brine Storage Tanks & Discharge Equipment		15 Days Storage Capacity				0.16
Instrumentation	10% of Treatment Equipment					0.60
Mechanical/Piping	5% of Treatment Equipment					0.30
Electrical	7% of Treatment Equipment					0.42
Pumping Plant & Pipelines						0.50
Sub-total A						**10.04**
Contingency	20% of Sub-total A					2.01
Admin	20% of Sub-total A					2.01
TOTAL $ m						**14.06**

(C) Brine C; ZLD Processing Scheme

Item	Ref	Quantity	Material Cost $ m	Delivery Cost (2.5% of Material) $ m	Installation Cost (15% of Material) $ m	Capex $ m
Concrete & Structural Steel			0.30	0.01	0.05	0.35
Buildings	$200/sq.ft.	2500 sq.ft.	0.500	0.013	0.075	0.588
Dedicated Landfill	Capex 0.22 M /acre including Haulage Cost	17 Acres	3.74	0.00	0.00	3.74
SEPCON Plant Equipment	Capex $2.5 m for 1 MGD Flow	Total Treatment System	3.30	0.08	0.49	3.88
RO Desalter #1	Capex $1.4 m for 1 MGD	1	1.40	0.04	0.21	1.65
Brine Concentrator	Flow rate 0.33 MGD Capex $6.8 m for 1MGD	1	2.24	0.06	0.34	2.64
Crystallizer	Flow rate 0.08 MGD Capex $43m for 1MGD Flow	1	3.440	0.086	0.516	4.04
Instrumentation	10% of Treatment Equipment					1.22
Mechanical/Piping	5% of Treatment Equipment					0.61
Electrical	7% of Treatment Equipment					0.85
Pumping Plant & Pipelines						0.70
Subtotal A						**20.26**
Contingency	20% of Subtotal A					4.05
Admin	20% of Subtotal A					4.05
TOTAL $ m						**28.31**

(D) Brine C; Partial Treatment with Commercial Disposal

Item	Ref	Quantity	Material Cost $ m	Delivery Cost (2.5% of Material) $ m	Installation Cost (15% of Material) $ m	Capex $ m
Concrete & Structural Steel			0.30	0.01	0.05	0.35
Buildings	$200/sq.ft.	2500 sq.ft.	0.500	0.013	0.075	0.588
SEPCON Plant Equipment	Capex $2.5 m for 1 MGD	Total Treatment	3.30	0.08	0.49	3.88
RO Desalter #1	Capex $1.4 m for 1 MGD	1	1.40	0.04	0.21	1.65
Brine Concentrator	Flow rate 0.33 MGD	1	2.24	0.06	0.34	2.64
Brine Storage Tanks & Discharge Equipment		15 Days Storage Capacity				0.16
Instrumentation	10% of Treatment Equipment					0.82
Mechanical/Piping	5% of Treatment Equipment					0.41
Electrical	7% of Treatment Equipment					0.57
Pumping Plant & Pipelines						0.70
Subtotal A						**10.83**
Contingency	20% of Subtotal A					2.17
Admin	20% of Subtotal A					2.17
TOTAL $ m						**15.17**

5.4 Conceptual Operating and Maintenance Costs

Using the conceptual flow diagrams and information from chemical supplies on reagent prices, the conceptual operating and maintenance cost estimates (also referred to as Operation Expenditure “Opex”) were prepared for Brines A and C separately . These estimates are given in Table 7. The estimated Opex for the treatment and disposal of 1 MGD of Brine A in the recommended ZLD processing scheme is $2.9 m and for the variance involving partial treatment with commercial disposal is $3.8 m. The estimated Opex for treating and disposing the same volume of Brine C is $5.1 m for the recommended ZLD processing scheme and $8.2 m for the variance involving partial treatment with commercial disposal.

As indicated earlier and reflected in Table 7, the dramatic rise in the Opex of partial treatment option is because of the costs associated with off-site disposal of reject from brine concentrator. This essentially removes the cost saving achieved through the exclusion of crystallizer and landfill steps from the overall processing scheme. Note that this cost estimation is based on the assumption that a disposal through a brine line will be at cost comparable to that of disposal through a commercial facility.

When comparing the ZLD options for Brines A and C, the Opex difference is largely because of salinity difference between the two brine streams; this is reflected in a significantly higher cost of consumables (raw material) used in the case of treatment of Brine C. To a lesser degree is the higher energy cost for the use of two RO units in the Brine A treatment train. Conversely, the energy cost is significantly lower for both brines when treatment excludes the use of a crystallizer; however this saving is overwhelmingly negated because of the high cost of off-site disposal of BC stream.

Figure 7. Conceptual O&M cost estimate for the recommended ZLD processing scheme and partial treatment with commercial disposal for Brines A and C.

(A) Brine A; ZLD Processing Scheme

Production Cost Summary				
	Product Streams			
	TDS Removal			
	Process Route			
	Plant Capacity	10,000 tpa	(12% Excess Design Capacity)	
	Capacity Utilization	88%		
	Fixed Capital IBL			
	Fixed Capital OBL			
	Total Fixed Capital	$11.94 M		
	Production Cost			Annual cost, $
	Raw Materials	Usage, pa	Unit cost, $/unit	
A	Brine, m^3	0	0	-
B	Polymer & RO Reagents			10,000
C	Soda Ash	4,115	135	555,530
D	Quicklime	2,005	75	150,400
E	RO Membranes			67,000
F	**Total Raw Material cost**			782,930
	Utilities	Usage, PA	Unit cost, $/unit	
G	Electricity, MWh	1,800	70	126,000
H	Gas, GJ	9,000	3.00	27,000
I	Process Water, ML	0	0.00	-
J	**Total Utilities Cost**			153,000
K	**Total Process Labour Wages**	5	70,000	350,000
L	**Payroll Overheads**		22% k	140,000
M	Maintenance		4%Capital	477,600
N	Operating Supplies		15%L	52,500
O	Plant Overheads		25% (K+M)	120,000
P	Insurance		1% Capital	119,300
Q	Book Depreciation		5% Capital	597,000
R	**Sub-total:**			**2,792,330**
	Non Manufacturing Costs			
S	R&D & Selling Prices		5% R	139,620
T	**Total O&M Cost $**			**2,931,950**

(B) Brine A; Partial Treatment with Commercial Disposal

Production Cost Summary				
	Product Streams			
	TDS Removal			
	Process Route			
	Plant Capacity	10,000 tpa	(12% Excess Design Capacity)	
	Capacity Utilization	88%		
	Fixed Capital IBL			
	Fixed Capital OBL			
	Total Fixed Capital	$10.04 M		
	Production Cost			Annual cost, $
	Raw Materials	Usage, pa	Unit cost, $/unit	
A	Brine, m^3	0	0	-
B	Polymer & RO Reagents			10,000
C	Soda Ash	4,115	135	555,530
D	Quicklime	2,005	75	150,400
E	RO Membranes			67,000
F	**Total Raw Material cost**			782,930
	Utilities	Usage, PA	Unit cost, $/unit	
G	Electricity, MWh	1,800	70	81,900
H	Gas, GJ	9,000	3.00	27,000
I	Process Water, ML	0	0.00	-
J	**Total Utilities Cost**			108,900
K	**Total Process Labour Wages**	5	70,000	350,000
L	**Payroll Overheads**		22% k	140,000
	Commercial Brine Disposal Including Haulage	8.40 MG/Yr	$140/1000 G ($6/bbl)	1,176,000
M	Maintenance		4%Capital	401,600
N	Operating Supplies		15%L	52,500
O	Plant Overheads		25% (K+M)	120,000
P	Insurance		1% Capital	114,300
Q	Book Depreciation		5% Capital	502,000
R	**Sub-total:**			**3,748,230**
	Non Manufacturing Costs			
S	R&D & Selling Expenses		1% R	37,480
T	**Total O&M Cost $**			**3,785,710**

(C) Brine C; ZLD Processing Scheme

Production Cost Summary				
	Product Streams		(15% Excess Design Capacity)	
	TDS Removal			
	Process Route			
	Plant Capacity	35,000 tpa		
	Capacity Utilization	85%		
	Fixed Capital IBL			
	Fixed Capital OBL			
	Total Fixed Capital	$20.26 M		
	Product Selling Price			
	Production Cost			Annual cost, $
	Raw Materials	Usage, pa	Unit cost, $/unit	
A	Brine, m^3	-	0	0
B	Polymer & RO Reagents	75	400	30,000
C	Soda Ash	11,385	135	1,537,000
D	Quicklime	3,895	75	292,150
E	RO Membranes			40,000
F	**Total Raw Material cost**			1,899,150
	Utilities		Unit cost, $/unit	-
G	Electricity, MWh	1,500	70	105,000
H	Gas, GJ	20,000	3.00	60,000
I	Process Water, ML		0.00	-
J	**Total Utilities Cost**			165,000
K	**Total Process Labour Wages**	5	70,000	350,000
L	**Payroll Overheads**		22%k	77,000
M	Maintenance		4%Capital	810,400
N	Operating Supplies		15%K	52,500
O	Plant Overheads		25%(K+M)	290,100
P	Insurance		1%Capital	202,600
Q	Book Depreciation		5%Capital	1,013,000
R	**Sub-total:**			**4,859,750**
	Non Manufacturing Costs			
S	R&D & Selling Expenses		5% R	242,990
T	**Total O&M Cost $**			**5,102,740**

(D) Brine C; Partial Treatment with Commercial Disposal

Production Cost Summary				
	Product Streams			
	TDS Removal			
	Process Route			
	Plant Capacity	35,000 tpa	(15% Excess Design Capacity)	
	Capacity Utilization	85%		
	Fixed Capital IBL			
	Fixed Capital OBL			
	Total Fixed Capital	$10.83M		
	Product Selling Price			
	Production Cost			Annual cost, $
	Raw Materials	Usage, pa	Unit cost, $/unit	
A	Brine, m^3	-	0	0
B	Polymer Flocculating Agent	75	400	30,000
C	Soda Ash	11,385	135	1,537,000
D	Quicklime	3,895	75	292,150
E	RO Membranes			40,000
F	**Total Raw Material cost**			1,899,150
	Utilities		Unit cost, $/unit	
G	Electricity, MWh	1,500	70	63,000
H	Gas, GJ	20,000	3.00	60,000
I	Process Water, ML		0.00	-
J	**Total Utilities Cost**			123,000
K	**Total Process Labour Wages**	5	70,000	350,000
L	**Payroll Overheads**		22%k	77,000
M	Commercial Brine Disposal Incling Haulage	29.2 MG/Yr	$140/1000 G ($6/bbl)	4,088,000
N	Maintenance		4% Capital	433,200
O	Operating Supplies		15% K	268,800
P	Plant Overheads		25% (K+N)	195,800
Q	Insurance		1% Capital	108,300
R	Book Depreciation		5% Capital	541,500
S	**Sub-total:**			**8,084,750**
	Non Manufacturing Costs			
T	R&D & Selling Expenses		1% S	80,850
V	**Total O&M Cost $**			**8,165,600**

6.0 RESULTS

6.1 Variation in the Conceptual Costs and Revenue Estimates

Table 8 is a summary listing of the estimated Capex and Opex for treatment of Brines A and B according to recommended process options at a flow rate of 1 MGD. As indicated, in the case of ZLD options the most probable Opex for Brine A is $2.93 m per year and the most probable Capex for the same brine is $16.7 m. For the option involving partial treatment and commercial disposal the Opex is $3.8 m per year and Capex is $14.1m.

The same analysis when applied to higher salinity Brine C provides higher cost ranges essentially reflecting the higher salt load to be removed at the same base flow. Accordingly, in the case of ZLD option the Opex is estimated $5.1 m per year and the Capex is about $28.3 m. For partial treatment and commercial disposal option the Opex is estimated $8.2 m per year and the Capex is about $15.2 m.

In assessing the economics of those processing options where the volume reduction and waste minimization leads to recovery of byproducts with commercial value, the combination of benefits resulting from the sale of recovered byproducts shall form part of the economic assessment. In the case of this project, SAL-PROC™ technology has been applied at a cost to achieve ZLD outcomes and to recover commercial grade byproducts. The sale of these byproducts will assist in offsetting the treatment costs or recover part of the initial capital outlay.

Table 8. Summary listing of the estimated Capex and Opex for treatment of Brines A and B according to recommended process options at a flow rate of 1 MGD.

Option	Annual Revenue, $m	Opex, $ m	Capex, $ m	Annual Cost*, $ m
Brine A - ZLD	3.29	2.93	16.72	4.51
Brine A - Partial Treatment & Commercial Disposal	3.29	3.79	14.06	5.12
Brine B - ZLD	5.26	5.10	28.31	7.77
Brine B - Partial Treatment & Commercial Disposal	5.26	8.17	15.17	9.60

** Sum of Opex and annulized cost at 7% over 20 years*

Given the scope of this study, Class V estimates of the Capex and Opex with accuracy in the range of -30% to +50% were prepared to assist in economic analysis of the proposed brine treatment options. Table 9 gives the range of estimated capex and opex for the treatment of Brines A and B in the recommended options based on Class V estimation (7% interest rate over 20 years operation).

As indicated in Tables 8 and 9, the estimated revenue potential from the sale of salts and fresh water is quite significant, with the most probable revenue base for ZLD option estimated at $3.3 m for Brine A and $5.3 m for Brine C.

Although the product pricing structure and the estimated revenues will need confirmation through further market studies and pilot trials, the data generated as part of this desktop study are considered sufficient in quantity and quality for a first-order economic analysis of the recommended treatment options. This analysis including the conceptual benefit-to-cost ratios and their comparisons are presented in the following section.

Table 9. Range of the estimated Capex and Opex for treatment of brines A and B according to recommended options using Class V Estimation method.

Option or Variance	Revenue($m)		Annual Cost ($m)		Net Cost ($m)	
	Range	Most Probable Case	Range	Most Probable Case	Range	Most Probable Case
Brine A - ZLD Processing	2.3 to 4.95	3.29	3.16 to 6.77	4.51	1.97 to -4.47	-1.22
Brine A - Partial Treatment with Commercial Disposal	2.3 to 4.95	3.29	3.58 to 7.68	14.1	1.37 to -5.38	-1.83
Brine C - ZLD Processing	3.68 to 7.89	5.26	5.44 to 11.66	7.77	2.45 to -7.98	-2.51
Brine C - Partial Treatment with Commercial Disposal	3.68 to 7.89	5.26	6.72 to 14.4	9.6	1.17 to -10.72	-4.34

6.2 Benefit-to-Cost Ratio

Comparing potential income benefits to cost, in a ratio, provides a glimpse into the economics of a wastewater treatment process in terms of potential for offsetting the cost of treatment. This type of analysis excludes the considerations of other benefits that the SAL-PROC™ process may offer to the overall desalination project.

For this desktop study, as shown in Table 10, the calculated benefit-to-cost ratio (B-C ratio) was used to determine if the proposed brine treatment systems offer potential to pay for themselves It is assumed that if the B-C ratio is greater than 1.0 the system has the potential to pay for itself. A ratio >1 means making profit and a ratio <1 means costing money. The revenue from the sale of byproducts as outlined earlier was assigned as the benefit in the numerator of the ratio. Other benefits to the project, briefly discussed earlier, are not included in this analysis. The annual cost is the sum of the annualized capital cost (Capex) and the O&M cost (Opex). A 20 year project life for the capital and a 7% interest rate are assumed. Table 11 details the assumptions and calculations for the most probable B-C ratio for ZLD and partial treatment options for Brines A and C.

Table 10. Method of benefit-to-cost ratio calculation used in this analysis.

(A) Benefit / Revenue
Income from products and water — **$million / year**

(B) Cost / Expenses (Opex)
O&M expenses — **$million / year**

(C) Annualized Capex — **$million / year**
Plant establishment cost annualized over economic life of the plant (20 years at 7%)

Benefit-to-Cost Ratio: $\dfrac{A}{B + C}$

Table 11. Calculations for the most probable B-C ratio for ZLD and partial treatment options for Brines A and C.

(Brine A)

Parameter	ZLD Option	Partial Treatment with Commercial Disposal
Benefit (Annual Revenue), $m	3.29	3.29
Opex, $ m	2.93	3.79
Capex, $ m	16.72	14.06
Annualized Capex (7%, 20 Yrs), $ m	1.58	1.33
Annual Cost, $ m	4.51	5.12
Benefit-to-Cost Ratio	**0.73**	**0.64**
Net Cost, $ m	-1.22	-1.83

(Bine C)

Parameter	ZLD Option	Partial Treatment with Commercial Disposal
Benefit (Annual Revenue), $m	5.26	5.26
Opex, $ m	5.10	8.17
Capex, $ m	28.31	15.17
Annualized Capex (7%, 20 Yrs), $ m	2.67	1.43
Annual Cost, $ m	7.77	9.6
Benefit-to-Cost Ratio	**0.65**	**0.55**
Net Cost, $ m	-2.51	-4.34

6.3 Sensitivity Analysis

The benefit-to-cost ratio is sensitive to the potential income benefit, and the capital and operating costs. A sensitivity analysis was conducted on the cost ranges presented above. The best case would result in the maximum monetary benefit and the least capital and operation and maintenance costs. The worst case would result in just the opposite.

Table 12 presents the range of possible benefit-to-cost ratios from the treatment of Brines A and C in either ZLD or partial treatment options. As indicated, the possible range of benefit-to-cost ratios is considerable; this indicates the sensitivity to costs used and the need to do additional work to improve accuracy. Considering the potential operational, environmental and financial benefits from an integrated desalination-brine treatment process the accuracy can be significantly improved by a better definition of the “whole process” parameters.

In any case, the analysis indicates that the described ZLD and partial treatment options offer cost-effective methods for disposing the EMWD brines from current and future operations.

Table 12. Sensitivity analysis of coats to Benefit-to-Cost Ratio from treatment of Brines A and C in either ZLD or partial treatment option.

(Brine A)

Parameter	ZLD Option			Partial Treatment with Commercial Disposal		
	Best Case	Most Probable Case	Worst Case	Best Case	Most Probable Case	Worst Case
Benefit (Annual Revenue), $m	4.95	3.29	2.3	4.95	3.29	2.3
Annual Cost, $ m	3.16	4.51	6.77	3.58	5.12	7.68
Benefit-to-Cost Ratio	1.57	0.73	0.34	1.38	0.64	0.3
Net Cost, $ m	1.79	-1.22	-4.47	1.37	-1.83	-5.38

(Brine C)

Parameter	ZLD Option			Partial Treatment with Commercial Disposal		
	Best Case	Most Probable Case	Worst Case	Best Case	Most Probable Case	Worst Case
Benefit (Annual Revenue), $m	7.89	5.26	3.68	7.89	5.26	3.68
Annual Cost, $ m	5.44	7.77	11.66	6.72	9.6	14.4
Benefit-to-Cost Ratio	1.45	0.65	0.32	1.17	0.55	0.26
Net Cost, $ m	2.45	-2.51	-7.98	1.17	-4.34	-10.72

7.0 CONCLUDING REMARKS AND RECOMMENDATIONS

The objectives of the study were to evaluate and determine whether the SAL-PROC™ based treatment systems are (a) technically feasible in dramatically reducing the brine flow streams from EMWD's RO processes and (b) economically viable. The answer to both questions is yes. SAL-PROC™ and the recommended ZLD process are technically feasible; SAL-PROC™ has been extensively piloted in Australia and already licensed. The treatment systems are also economically feasible as shown by the Benefit-to-Cost ratios, as presented in the previous chapter.

Furthermore, the proposed ZLD option and its variance involving the partial treatment and disposal provides several added benefits that are unmatched by any other technology or brine management system. These include:

- reduced operation and maintenance costs through recovery and sale of byproducts,
- reduced water loss through improvements in operational efficiency of the conventional volume reduction methods,
- reduced environmental impact and liability, and
- opportunity to achieve a sustainable solution.

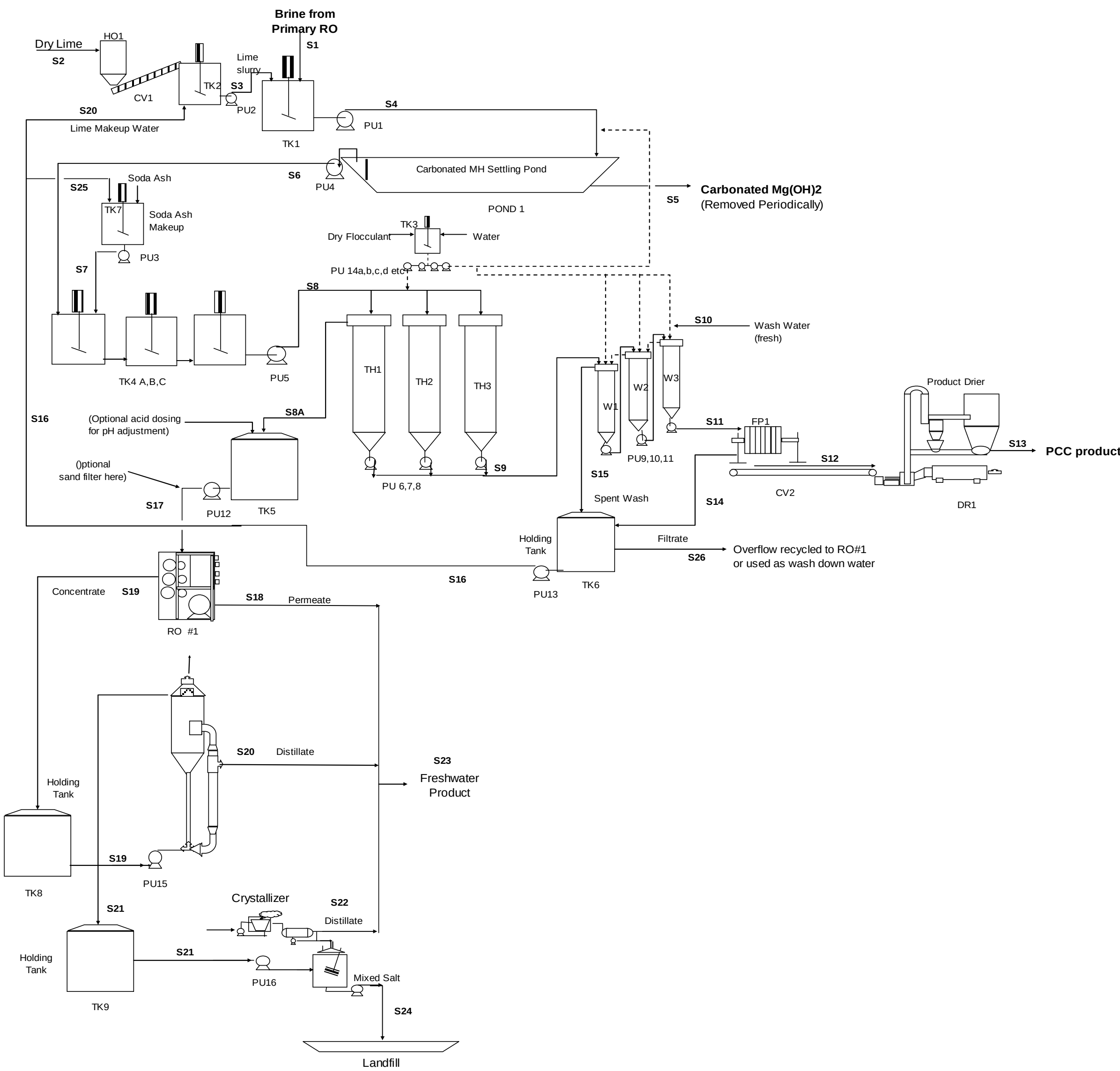
To progress forward some follow-up investigations need to be conducted. The followings are recommended for considerations:

- Pilot study using the EMWD brines and the ZLD processing option 2 to verify the results from the desk top modeling, assess the system performance and optimize treatment efficiencies.
- Compare the economics of SAL-PROC™ based ZLD processing options with the economics of the next best alternative for the disposal of EMWD brines.
- An integrated RO-Brine treatment facility to gain efficiencies.
- Further evaluation of beneficial uses of the mixed salts – including possible downstream processing for value adding.
- Marketing analysis development – confirm the local and regional interests.

APPENDICES

Appendix A

Schematic Flow Diagram Showing Equipment Arrangements for ZLD Processing Scheme Suitable for Brine C, Involving Product Recovery in the SAL-PROC™ Process



Appendix B

Preliminary Streams, Flows, Equipment Specifications and Descriptions. Refer to schematic flow diagram in Appendix A for equipment arrangement. Feed TDS salinity is assumed at 8 g/L, and a chemical make-up similar to that of Brine C.

(A) Preliminary Streams and flow Specifications

Stream	Liquid Flowrate 1000 g/hr	TDS g/L (ppm/1000)	Solids Flowrate US tons/hr	% solids	Description
S1	11	8 (Brine C)	-	-	Process feed water
S2	-	-	0.44	100%	Dry lime [Ca(OH)2]
S3	1.06		4.0	10%	Lime slurry to GMH rxn
S4	11	7.6	4.3	15%	Process slurry w Gyp.Mg(OH)2
S5			0.48	30-40%	GMH removed periodically
S6	11	22	3.47	-	Clarified solution to PCC rxn
S7	1.03	-	1.3	10%	Soda ash solution @ 30% strength
S8	11.1	29.6	4.67	<1%	Process slurry w PCC
S8A	10.9	22.6	3.56	-	Clarified solution (spent water)
S9	2.03	-	1.1	15%	Thickened PCC slurry
S10	7	-			Wash water
S11	2.03	-	1.1	15%	Washed PCC slurry
S12	-	-	2.8	50%	Dewatered PCC filter cake
S13	-	-	1.1	100%	Dry PCC product
S14	6.0	<1	-	-	Filtrate
S15	7	<1	-	-	Spent wash water
S16	1.06	<1	-	-	Lime makeup water
S17	10.9	22.6	3.56	-	RO# 1 feedwater
S18	7.3	<0.05%	-	-	RO#1 permeate
S19	3.6	67.4	3.55	-	RO#1 concentrate
S20	2.75	-	-	-	Distilate from BC
S21	0.85	270	3.55	-	Brine concentrator reject
S22	0.57	809	3.55	-	Distilate from XTL
S23	10.6	-	-	-	Combined freshwater from all streams
S24	-	-	3.7	-	Salts may be further refined. Mixture dominated by NaCl and Na2SO4
S25	3.2	-	-	-	Soda Ash makeup water

APPENDIX C

Total mass and flows for 1MG flows and corresponding ionic balances for Brine A and Brine C

(Brine A)

Step	Description	Total Mass (Tons) & Flow (1 MGD)	Na mg/L	Ca mg/L	Mg mg/L	K mg/L	SO4 mg/L	Cl mg/L	HCO3 (CO3) mg/L	Si mg/L	OH mg/L	TDS mg/L
1	Feed water	1	850	868	260	28.5	1215	2250	850	180	0	6500
2	Lime	5.49T	850	1580	260	28.5	1215	2250	850	180	367	7579
3	Carbonated Mg(OH)2	5.53T	850	1300	0	28.5	1215	2250	423	20	0	6086
4	Spent water	1	850	1300	0	28.5	1215	2250	414	20	0	6077
5	RO1 permeate	0.6	0	0	0	0	0	0	0	50	0	0
6	RO1 reject	0.4	2215	3250	0	71.3	3037.5	5625	1035	50	0	15193
7	Soda ash	11.27T	5069	3250	0	71.3	3037.5	5625	1035	50	0	18138
8	PCC product	13.51T	5069	0	0	71.3	3037.5	5625	1035	50	0	13853
9	Spent water	0.4	5069	0	0	71.3	3037.5	5625	1035	50	0	13853
10	RO2 permeate	0.24	0	0	0	0	0	0	0	0	0	0
11	RO2 reject	0.16	12672.5	0	0	178.3	7594	14062.5	0	125	0	34632
12	BC water	0.137	0	0	0	0	0	0	0	0	0	0
13	BC reject	0.023	88707	0	0	1248	53158	98437	0	875	0	242425
14	Crystallizer water	0.02	0	0	0	0	0	0	0	0	0	0
15	Mixed salts	23.3T	88707	0	0	1248	53158	98437	0	875	0	848487

(Brine C)

Step	Description	Total Mass (T) & Flow (ML), I MGD	Na mg/L	K mg/L	Ca mg/L	Mg mg/L	Cl mg/L	SO4 mg/L	HCO3 mg/L	NO3 mg/L	Si mg/L	OH mg/L	TDS mg/L
1	Feed water	1	5477	71	1550	684	10599	3338	316	60	132	0	22227
2	Lime	2.82	5477	71	2934	684	10599	3338	316	60	132	119	23730
3	Carbonated Mg(OH)2	3.06	5477	71	2726	0	10599	3000	0	60	20	119	22072
4	Spent water	1	5477	71	2726	0	10599	3000	0	60	20	119	22072
5	Soda ash	8.24	8729	71	2726	0	10599	3000	4242	60	20	119	29566
6	PCC product	7.78	8729	71	0	0	10599	3000	0	60	5	119	22583
7	Spent water	1	8729	71	0	0	10599	3000	0	60	5	119	22583
8	RO2 permeate	0.67	0	0	0	0	0	0	0	0	0	0	0
9	RO2 reject	0.33	26187	213	0	0	31797	9000	0	180	15	0	67392
10	BC water	0.25	0	0	0	0	0	0	0	0	0	0	0
11	BC Reject	0.08	104748	852	0	0	127188	36000	0	720	60	0	269568
12	Crystallizer water	0.053	0	0	0	0	0	0	0	0	0	0	0
13	Mixed salts	23.72											808704

Appendix C

HPD – VEOLIA WATER SOLUTIONS REPORT ON BRINE CONCENTRATOR AND CRYSTALLIZER

HPD ZLD Budget Estimates

Item / Capacity	694 gpm	347 gpm	174 gpm	86 gpm
TDS 6300 mg/l :				
Brine Concentrator (BC) - Equipment Only	$9,000,000	$5,500,000	$4,000,000	$3,000,000
Brine Crystallizer (BX) - Equipment Only	$1,500,000	$1,000,000	$700,000	$500,000
Est. Installation Costs	$9,000,000	$7,000,000	$5,000,000	$4,500,000
Total Turnkey Costs	$19,500,000	$13,500,000	$9,700,000	$8,000,000
Power Consumption	3600 kW	1800 kW	900 kW	450 kW
Solids Waste Rate @ 10% moisture	29 tons/day	15 tons/day	7.3 tons/day	3.6 tons/day
Operating Labor Req'ts	13,000 man-hr/yr	10,000 man-hr/yr	8,000 man-hr/yr	6,000 man-hr/yr
Spares & Maintenance Allowance @ 3% of Equipment Costs/yr	$315,000.00	$195,000.00	$141,000.00	$105,000.00
Chemical & Cleaning Costs/yr	$125,000	$85,000	$67,000	$55,000
TDS 28000 mg/l :				
Brine Concentrator - Equipment Only	$9,000,000	$5,500,000	$4,000,000	$3,000,000
Brine Crystallizer (BX) - Equipment Only	$3,000,000	$2,000,000	$1,500,000	$1,000,000
Est. Installation Costs	$9,000,000	$7,000,000	$5,500,000	$4,500,000
Total Turnkey Costs	$21,000,000	$14,500,000	$11,000,000	$8,500,000
Power Consumption	3800 kW	1900 kW	975 kW	500 kW
Solids Waste Rate @ 10% moisture	117 tons/day	58 tons/day	29 tons/day	15 tons/day
Operating Labor Req'ts	14,000 man-hr/yr	11,000 man-hr/yr	9,000 man-hr/yr	7,000 man-hr/yr
Spares & Maintenance Allowance @ 3% of Equipment Costs/yr	$360,000.00	$225,000.00	$165,000.00	$120,000.00
Chemical & Cleaning Costs/yr	$140,000	$95,000	$75,000	$60,000

Appendix D

RAW DATA FILES FOR SEEDED RO PILOT TEST

SPARRO - Batch Test Log Sheet
(1st Run on Secondary RO brine)
Date: October 2, 2006

Operating Time (min)	Time Actual	Pressures (psi)			Flow (gpm)					Seed Conc (g/L)		Conductivity (mS/cm)			Temp (C)	Feed Tank	
		Feed	Before P.Red	After P.Red	Permeate	Feed Bypass	Brine Bypass	Brine (incl b/p)	Pump Output (Calc)	Feed	Brine	Feed	Brine	Permeate	Feed	Level (Approx)	
0	12:30											30.8			26.3		
0	12:50	460	240		0.58	4.29		2.73	7.59			30.5	32.5	14.1	27.8	120	
20	13:10	480	280		0.62	4.62		2.07	7.30			31.8	34.8	19.5	29.2	107	
20	13:30	500	295		0.58	5.00		2.00	7.58			33	36.7	20.5	31	95	
20	13:50	510	340		0.69	3.53		3.16	7.38			34.5	37.7	20.1	32.4	80	
20	14:10	510	420		0.63	4.29		2.50	7.42			37.1	40.7	20.8	33.8		
20	14:30	570	420		0.67	3.53		2.73	6.93			39.4	42.4	22.8	34.6	50	
20	14:50	600	480		0.65	3.75		2.50	6.90			43.4	46.5	24.8	32.3	32	
																30 at 14:55	

= sample taken

MPT 34 Membranes - 200 Mol Mass cut-off
Module Length: 12 ft
Membrane area: 28 ft2
28

Membrane X-area: 0.000122734 m2

Cumulative Time	Permeate Production	Tank Volume	Recovery	CF	Apparent Seed Conc	Apparent Flux	Rejection (Cond)	Delta P	Exit Flow	Exit Velocity	Estimated Feed TDS	Estimated Prod TDS				Permeate	Feed Bypass	Bypass	Brine + Bypass
(min)	(gal)	(gal)	(%)		(g/L)	(gdf)	(%)	(psi)	(m3/sec)	(m/sec)	(mg/L)	(mg/L)				Time for 1 gal (sec)			
0		120			18														
0	0.0	120	0.00	1.00	18.0	29.7	53.8	220	0.000172	1.40	20,435	9,165				104	14		22
20	12.0	108	9.96	1.11	20.0	31.8	38.7	200	0.000131	1.06	21,306	12,675				97	13		29
40	23.9	96	19.92	1.25	22.5	29.7	37.9	205	0.000126	1.03	22,110	13,325				104	12		30
60	36.6	83	30.48	1.44	25.9	35.5	41.7	170	0.000199	1.62	23,115	13,065				87	17		19
80	49.8	70	41.49	1.71	30.8	32.5	43.9	90	0.000158	1.28	24,857	13,520				95	14		24
100	62.8	57	52.37	2.10	37.8	34.7	42.1	150	0.000172	1.40	26,398	14,820				89	17		22
120	76.1	44	63.42	2.73	49.2	33.5	42.9	120	0.000158	1.28	29,078	16,120				92	16		24

SPARRO - Batch Test Log Sheet - Operating on Brine from Secondary RO Pilot Plant
(2nd test on Secondary RO brine)
Date:October 16, 2006

MPT 34 Membranes - 200 Mol Mass cut-off
Module Length: 12 ft
Membrane area: 28 ft2
28

Membrane X-area: 0.00012273 m2

Mass of gypsum added to tank (lbs) **10**

Operating Time (min)	Time Actual	Pressures (psi) Feed	Before P.Red	Delta P	Flow (gpm) Permeate	Feed Bypass	Brine Bypass	Brine (incl b/p)	Pump Output (Calc)	Seed Conc (g/L) Feed	Seed Conc (g/L) Brine	Conductivity (mS/cm) Feed	Brine	Permeate	Temp (C) Feed	Feed Tank Level (Approx)
0	12:30	340				4.29										
0	12:45	340	180	160	0.55	4.62		2.50	7.66			33.1	36.8	21.8	24	120
20	1:05	510	260	250	0.58	4.29		2.22	7.08			34	36.5	21.5	23.2	105
20	1:25	580	300	280	0.67	4.62		2.73	8.02			35.4	38.7	20.4	24.4	95
20	1:45	470	310	160	0.51	2.73		4.29	7.52			37.4	39	20.7	25.6	85
20	2:05	480	330	150	0.49	2.40		4.29	7.18			38.1	40.1	21.2	25.3	72
20	2:25	520	380	140	0.53	3.16		3.75	7.44			40.7	42.6	20.4	26.7	65
20	2:45	460	390	70	0.29	4.00		3.16	7.45			42.9	44.8	21	26.9	53
20	3:05	480	390	90	0.25	4.29		3.16	7.69			45.1	46.5	21.3	28.9	47
20	3:25	640	390	250	0.22	3.33		3.75	7.30			46.7	47.8	20.3	28.7	43
20	3:45	640	240	400	0.15	3.00		4.00	7.15			48.7	49.9	20.5	29.7	38
15	4:00	660	260	400	0.13	3.5294118										~35

= sample taken

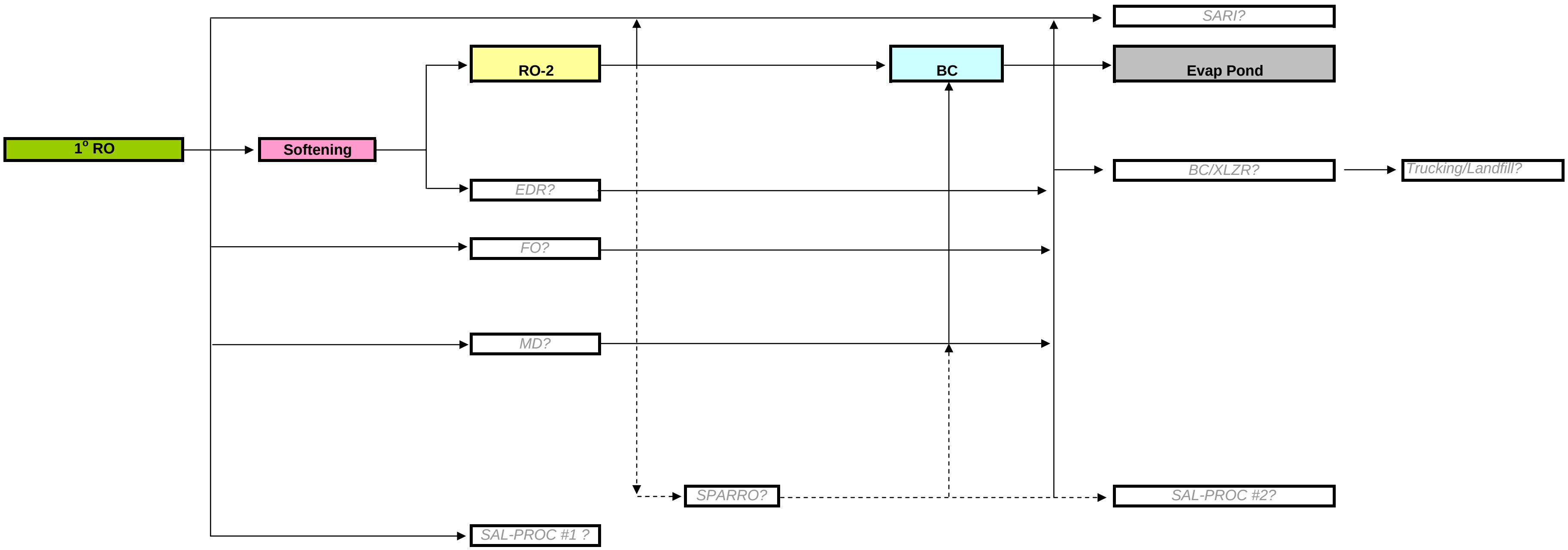
Cumulative Time (min)	Permeate Production (gal)	Tank Volume (gal)	Recovery (%)	CF	Apparent Seed Conc (g/L)	Apparent Flux (gdf)	Rejection (Cond) (%)	Delta P (psi)	Exit Flow (m3/sec)	Exit Velocity (m/sec)	Estimated Feed TDS (mg/L)	Estimated Prod TDS (mg/L)
0		120			10.0							
0	0.0	120	0.00	1.00	10.0	28.1	34.1	160	0.000158	1.28	22,177	14,170
20	11.2	109	9.35	1.10	11.0	29.7	36.8	250	0.000140	1.14	22,780	13,975
40	23.7	96	19.78	1.25	12.5	34.7	42.4	280	0.000172	1.40	23,718	13,260
60	35.6	84	29.63	1.42	14.2	26.2	44.7	160	0.000270	2.20	25,058	13,455
80	45.6	74	37.97	1.61	16.1	25.3	44.4	150	0.000270	2.20	25,527	13,780
100	55.8	64	46.49	1.87	18.7	27.3	49.9	140	0.000237	1.93	27,269	13,260
120	64.0	56	53.37	2.14	21.5	15.1	51.0	70	0.000199	1.62	28,743	13,650
140	69.5	51	57.90	2.38	23.8	12.9	52.8	90	0.000199	1.62	30,217	13,845
160	74.1	46	61.78	2.62	26.2	11.1	56.5	250	0.000237	1.93	31,289	13,195
180	77.7	42	64.79	2.84	28.4	7.5	57.9	400	0.000252	2.06	32,629	13,325
195	79.8	40	66.52	2.99	29.9	6.7		400	0.000000	0.00	-	-

Cumulative Time (min)	Recovery Qp/Qf	Log Mean Conc Factor	Osmotic Pressure (Atmos)	Osmotic Pressure (psi)	Net Driving Pressure NDPt	Temp Correction Factor TCPt	Normalized flow Qn	Time for 1 gal (sec) Permeate	Feed Bypass	Bypass	Brine + Bypass
					86.06	TCFr 1.00					
0									14		
0	0.18	1.10	10.8	159		1.03		110	13		24
20	0.21	1.12	11.3	166	204.05	1.06	0.230	104	14		27
40	0.20	1.11	11.7	172	253.05	1.02	0.225	89	13		22
60	0.11	1.06	11.7	173	202.69	0.98	0.220	118	22		14
80	0.10	1.06	11.9	176	214.76	0.99	0.199	122	25		14
100	0.12	1.07	12.9	190	245.58	0.95	0.196	113	19		16
120	0.09	1.05	13.3	196	214.54	0.94	0.125	204	15		19
140	0.07	1.04	13.9	204	215.80	0.89	0.112	240	14		19
160	0.05	1.03	14.3	210	290.67	0.89	0.071	279	18		16
180	0.04	1.02	14.7	216	208.84	0.87	0.069	410	20		15
195	(0.04)	0.98	-	-					17		17

Appendix E

DETAILED COST ESTIMATES

Selection criteria:



User Selection:

Design Capacity	MGD 1.00		MGD 1.00		MGD 0.3		MGD 0.3		MGD 0.045	
1° RO	→	Softening	→	RO-2	→	none	→	BC only	→	Evap Pond

Total Capital Cost, $	$4,350,375	+	$2,295,000	+	$0	+	$17,339,739	+	$1,578,377	=	$25,563,491
Capital Costs, $/annum	$379,286	+	$200,089	+	$0	+	$1,512,757	+	$137,610	=	$2,229,742
O&M Costs, $/annum	$763,921	+	$181,390	+	$0	+	$1,791,495	+	$902,022	=	$3,638,828
Total Costs, $/annum	$1,143,207	+	$381,478	+	$0	+	$3,304,253	+	$1,039,632	=	$5,868,570
Generated Revenue, $/annum	$0	+	$430,501	+	$0	+	$156,825	+	$0	=	$587,326

Notes: Cells that are RED contain user-put options

DESCRIPTION	TREATMENT OPTION										
	Softening	**Secondary RO**	**EDR**	**SPARRO**	**SAL-PROC[4]**		**FO**	**Evap Pond**	**SARI**	**B.C only**	**B.C and XLZR[5]**
Assumed recovery (%)		70%	75%	60%	**Brine A (#1)**	**Brine C (#2)**	70%			**Brine B**	**Brine B**
					Total ZLD	Total ZLD					
Design Informatior											
Capacity (mgd)	1.00	1.00	0.00	0.00	0.00	0	0.00	0.05	0.00	0.30	0.05
Utilization Rate (%)											
Footprint Area (ft^2) [2]											
Length (ft) x Width (ft)											
Fenced Area - ft x ft (ft^2)											
Capital Cost											
Equipment Cost (incl. Installation)	$2,050,000	$1,000,000	$462,197	$28,274	$12,668,802	$7,168,672	$3,849,011			$8,858,104	$6,480,972
Interconnecting Pipework, 12.50%	$256,250	$125,000	$50,000	$3,534	$1,583,600	$896,084	$481,126			$1,107,263	$810,122
Electrical and Instrumentation, 18%	$369,000	$180,000		$5,089	$2,280,384	$1,290,361	$692,822			$1,594,459	$1,166,575
Building ($150/sq ft)	$225,000	$225,000	$0	$0	$450,000	$67,500	$0			Outdoors - No Bldg	Outdoors - No Bldg
Subtotal Construction Cost	$2,900,250	$1,530,000	$512,197	$36,898	$16,982,786	$9,422,617	$5,022,959			$11,559,826	$8,457,669
Engineering, 15%	$435,038	$229,500	$76,830	$5,535	$2,547,418	$1,413,393	$753,444			$1,733,974	$1,268,650
Legal and Admin, 10%	$290,025	$153,000	$51,220	$3,690	$1,698,279	$942,262	$502,296			$1,155,983	$845,767
Contingency, 25%	$725,063	$382,500	$128,049	$9,224	$4,245,696	$2,355,654	$1,255,740			$2,889,957	$2,114,417
Total Capital Cost	**$4,350,375**	**$2,295,000**	**$839,081**	**-$168**	**$25,474,179**	**$14,133,926**	**$3,211**	**$1,578,377**	**$8,100,000**	**$17,339,739**	**$12,686,503**
Annual Cost of Capita	**$379,286**	**$200,089**	**$73,155**	**$4,825**	**$2,220,955**	**$1,232,260**	**$280**	**$137,610**	**$706,195**	**$1,512,757**	**$1,106,067**
O&M Cost											
Labor	$278,096	incl with Softening	incl with Softening	-	$350,000	$350,000	$105,693			$441,106	$321,314
Electricity[1]	$35,040	$97,440	$0	$0	$279,000	$36,000	$410,537			$1,167,928	$416,747
Chemicals	$763,921	$25,550	$0	$0	$2,099,360	$801,396	$258,362			$64,713	$44,045
Media / Equipment Replacement	-	$58,400	$0	-$168	$67,000	$67,000	$413,609			$117,749	$86,078
Waste Handling/Disposal	$259,106	-	-	-					**$4**	-	$94,780
					$3,331,147	$1,926,282					
Including disposal ?[2] (YES/NO)	no	no									
Annual O&M Costs	**$763,921**	**$181,390**	**$0**	**-$168**	**$6,551,732**	**$3,245,557**	**$1,170**	**$902,022**	**$4**	**$1,791,495**	**$962,964**
Present Worth (2007 USD)											
Revenue Generated		**$430,501**	**$0**	**$0**	**#REF!**	**#REF!**	**$0**			**$156,825**	**$24,908**
Capital Cost[3]	**$379,286**	**$200,089**	**$73,155**	**$4,825**	**$2,220,955**	**$1,232,260**	**$280**	**$137,610**	**$706,195**	**$1,512,757**	**$1,106,067**
O&M Cost	**$763,921**	**$181,390**	**$0**	**-$168**	**$6,551,732**	**$3,245,557**	**$1,170**	**$902,022**	**$4**	**$1,791,495**	**$962,964**
Total Costs	**$1,143,207**	**$381,478**	**$73,155**	**$4,658**	**$8,772,687**	**$4,477,817**	**$1,450**	**$1,039,632**	**$706,199**	**$3,304,253**	**$2,069,031**

Notes:
1. Electrical power cost of $0.12/kWh assumed
2. User chooses whether or not to include disposal costs (sludge for Softening, brine for RO - SARI, and solids for BC/XLZR)
3. Annual costs assume 6% interest rate, and 20 yr loan period
4. SAL-PROC: Partial Treatment includes price for commercial brine disposal, ZLD does not include landfill costs
5. Assume solids disposal of $50/ton
6. Labor cost is priced at $47.75hr.
7. Soda ash is priced at $0.165/LB
8. Lime is priced at $300/ton
9. Polymer & RO Reagents are priced at $0.07/1000 gal
10. Natural gas is priced at $7/GJ
11. RO Membranes priced at $0.16/1000 gal

Process Combination: Base Case 1°RO + SARI

DESCRIPTION	TREATMENT OPTION										
	Softening	Secondary RO	EDR	SPARRO	SAL-PROC[4]		FO	Evap Pond	SARI	B.C only	B.C and XLZR[5]
Assumed recovery (%)		70%	75%	60%	Brine A (#1)	Brine C (#2)	70%			Brine B	Brine B
Design Information											
Capacity (mgd)									1.00		
Utilization Rate (%)											
Footprint Area (ft^2) [2]											
Length (ft) x Width (ft)											
Fenced Area - ft x ft (ft^2)											
Capital Cost											
Equipment Cost (incl. Installation)											
Interconnecting Pipework, 12.50%											
Electrical and Instrumentation, 18%											
Building ($150/sq ft)											
Subtotal Construction Cost											
Engineering, 15%											
Legal and Admin, 10%											
Contingency, 25%											
Total Capital Cost									**$8,100,000**		
Annual Cost of Capital									**$1,071,195**		
O&M Cost											
Labor											
Electricity[1]											
Chemicals											
Media / Equipment Replacement											
Waste Handling/Disposal											
Including disposal ?[2] (YES/NO)											
Annual O&M Costs											
Present Worth (2007 USD)											
Revenue Generated											
Capital Cost[3]											
O&M Cost											
Total Costs									**$1,071,195**		

Process Combination: 0 1^{o}RO + Softening + 2^{o}RO + SARI

DESCRIPTION	TREATMENT OPTION										
	Softening	Secondary RO	EDR	SPARRO	SAL-PROC[4]		FO	Evap Pond	SARI	B.C only	B.C and XLZR[5]
Assumed recovery (%)		70%	75%	60%	Brine A (#1)	Brine C (#2)	70%			Brine B	Brine B
Design Information											
Capacity (mgd)	1.00	1.00							0.30		
Utilization Rate (%)											
Footprint Area (ft^2) [2]											
Length (ft) x Width (ft)											
Fenced Area - ft x ft (ft^2)											
Capital Cost											
Equipment Cost (incl. Installation)	$2,050,000	$1,000,000									
Interconnecting Pipework, 12.50%	$256,250	$125,000									
Electrical and Instrumentation, 18%	$369,000	$180,000									
Building ($150/sq ft)	$225,000	$225,000									
Subtotal Construction Cost	$2,900,250	$1,530,000									
Engineering, 15%	$435,038	$229,500									
Legal and Admin, 10%	$290,025	$153,000									
Contingency, 25%	$725,063	$382,500									
Total Capital Cost	**$4,350,375**	**$2,295,000**							**$8,100,000**		
Annual Cost of Capital	**$379,286**	**$200,089**							**$706,195**		
O&M Cost											
Labor	$278,096	incl with Softening									
Electricity[1]	$35,040	$97,440									
Chemicals	$763,921	$25,550									
Media / Equipment Replacement	-	$58,400									
Waste Handling/Disposal	$259,106	-							**$32,854**		
Including disposal ?[2] (YES/NO)	no	no									
Annual O&M Costs	**$763,921**	**$181,390**							**$32,854**		
Present Worth (2007 USD)											
Revenue Generated		**$430,501**									
Capital Cost[3]	**$379,286**	**$200,089**							**$706,195**		
O&M Cost	**$763,921**	**$181,390**							**$32,854**		
Total Costs	**$1,143,207**	**$381,478**							**$739,049**		

Process Combination:	1	1°RO + Softening + 2°RO + BC + SARI									
DESCRIPTION	**TREATMENT OPTION**										
	Softening	**Secondary RO**	**EDR**	**SPARRO**	**SAL-PROC[4]**		**FO**	**Evap Pond**	**SARI**	**B.C only**	**B.C and XLZR[5]**
Assumed recovery (%)		70%	75%	60%	**Brine A (#1)**	**Brine C (#2)**	70%			**Brine B**	**Brine B**
Design Information											
Capacity (mgd)	1.00	1.00							0.05	0.30	
Utilization Rate (%)											
Footprint Area (ft^2) [2]											
Length (ft) x Width (ft)											
Fenced Area - ft x ft (ft^2)											
Capital Cost											
Equipment Cost (incl. Installation)	$2,050,000	$1,000,000								$8,858,104	
Interconnecting Pipework, 12.50%	$256,250	$125,000								$1,107,263	
Electrical and Instrumentation, 18%	$369,000	$180,000								$1,594,459	
Building ($150/sq ft)	$225,000	$225,000								Outdoors - No Bldg	
Subtotal Construction Cost	$2,900,250	$1,530,000								$11,559,826	
Engineering, 15%	$435,038	$229,500								$1,733,974	
Legal and Admin, 10%	$290,025	$153,000								$1,155,983	
Contingency, 25%	$725,063	$382,500								$2,889,957	
Total Capital Cost	**$4,350,375**	**$2,295,000**							**$8,100,000**	**$17,339,739**	
Annual Cost of Capital	**$379,286**	**$200,089**							**$706,195**	**$1,512,757**	
O&M Cost											
Labor	$278,096	incl with Softening								$441,106	
Electricity[1]	$35,040	$97,440								$1,167,928	
Chemicals	$763,921	$25,550								$64,713	
Media / Equipment Replacement	-	$58,400								$117,749	
Waste Handling/Disposal	$259,106	-							**$4,932**	-	
Including disposal ?[2] (YES/NO)	no	no									
Annual O&M Costs	**$763,921**	**$181,390**							**$4,932**	**$1,791,495**	
Present Worth (2007 USD)											
Revenue Generated		**$430,501**								**$156,825**	
Capital Cost[3]	**$379,286**	**$200,089**							**$706,195**	**$1,512,757**	
O&M Cost	**$763,921**	**$181,390**							**$4,932**	**$1,791,495**	
Total Costs	**$1,143,207**	**$381,478**							**$711,127**	**$3,304,253**	

Process Combination: **2** **1°RO + Softening + 2°RO + BC + XLZR-Disposal**

DESCRIPTION	TREATMENT OPTION										
	Softening	**Secondary RO**	**EDR**	**SPARRO**	**SAL-PROC[4]**		**FO**	**Evap Pond**	**SARI**	**B.C only**	**B.C and XLZR[5]**
Assumed recovery (%)		70%	75%	60%	**Brine A (#1)**	**Brine C (#2)**	70%			**Brine B**	**Brine B**
Design Information											
Capacity (mgd)	1.00	1.00									0.30
Utilization Rate (%)											
Footprint Area (ft^2) [2]											
Length (ft) x Width (ft)											
Fenced Area - ft x ft (ft^2)											
Capital Cost											
Equipment Cost (incl. Installation)	$2,050,000	$1,000,000									$11,497,476
Interconnecting Pipework, 12.50%	$256,250	$125,000									$1,437,185
Electrical and Instrumentation, 18%	$369,000	$180,000									$2,069,546
Building ($150/sq ft)	$225,000	$225,000									Outdoors - No Bld
Subtotal Construction Cost	$2,900,250	$1,530,000									$15,004,207
Engineering, 15%	$435,038	$229,500									$2,250,631
Legal and Admin, 10%	$290,025	$153,000									$1,500,421
Contingency, 25%	$725,063	$382,500									$3,751,052
Total Capital Cost	**$4,350,375**	**$2,295,000**									**$22,506,310**
Annual Cost of Capital	**$379,286**	**$200,089**									**$1,962,203**
O&M Cost											
Labor	$278,096	incl with Softening									$441,106
Electricity[1]	$35,040	$97,440									$1,167,928
Chemicals	$763,921	$25,550									$77,192
Media / Equipment Replacement	-	$58,400									$172,055
Waste Handling/Disposal	$259,106	-									$609,521
Including disposal ?[2] (YES/NO)	no	no									
Annual O&M Costs	**$763,921**	**$181,390**									**$2,467,802**
Present Worth (2007 USD)											
Revenue Generated		**$430,501**									**$166,050**
Capital Cost[3]	**$379,286**	**$200,089**									**$1,962,203**
O&M Cost	**$763,921**	**$181,390**									**$2,467,802**
Total Costs	**$1,143,207**	**$381,478**									**$4,430,004**

Process Combination: 3 — **1°RO + Softening + 2°RO + BC + Evap Pond**

DESCRIPTION	TREATMENT OPTION										
	Softening	**Secondary RO**	**EDR**	**SPARRO**	**SAL-PROC[4]**		**FO**	**Evap Pond**	**SARI**	**B.C only**	**B.C and XLZR[5]**
Assumed recovery (%)		70%	75%	60%	**Brine A (#1)**	**Brine C (#2)**	70%			**Brine B**	**Brine B**
Design Information											
Capacity (mgd)	1.00	1.00						0.05		0.30	
Utilization Rate (%)											
Footprint Area (ft^2) [2]											
Length (ft) x Width (ft)											
Fenced Area - ft x ft (ft^2)											
Capital Cost											
Equipment Cost (incl. Installation)	$2,050,000	$1,000,000								$8,858,104	
Interconnecting Pipework, 12.50%	$256,250	$125,000								$1,107,263	
Electrical and Instrumentation, 18%	$369,000	$180,000								$1,594,459	
Building ($150/sq ft)	$225,000	$225,000								Outdoors - No Bldg	
Subtotal Construction Cost	$2,900,250	$1,530,000								$11,559,826	
Engineering, 15%	$435,038	$229,500								$1,733,974	
Legal and Admin, 10%	$290,025	$153,000								$1,155,983	
Contingency, 25%	$725,063	$382,500								$2,889,957	
Total Capital Cost	**$4,350,375**	**$2,295,000**						**$1,578,377**		**$17,339,739**	
Annual Cost of Capital	**$379,286**	**$200,089**						**$137,610**		**$1,512,757**	
O&M Cost											
Labor	$278,096	incl with Softening								$441,106	
Electricity[1]	$35,040	$97,440								$1,167,928	
Chemicals	$763,921	$25,550								$64,713	
Media / Equipment Replacement	-	$58,400								$117,749	
Waste Handling/Disposal	$259,106	-								-	
Including disposal ?[2] (YES/NO)	no	no									
Annual O&M Costs	**$763,921**	**$181,390**						**$902,022**		**$1,791,495**	
Present Worth (2007 USD)											
Revenue Generated		**$430,501**								**$156,825**	
Capital Cost[3]	**$379,286**	**$200,089**						**$137,610**		**$1,512,757**	
O&M Cost	**$763,921**	**$181,390**						**$902,022**		**$1,791,495**	
Total Costs	**$1,143,207**	**$381,478**						**$1,039,632**		**$3,304,253**	

Process Combination:	4	1°RO + Softening + EDR + BC + SARI									
DESCRIPTION	**TREATMENT OPTION**										
	Softening	**Secondary RO**	**EDR**	**SPARRO**	**SAL-PROC[4]**		**FO**	**Evap Pond**	**SARI**	**B.C only**	**B.C and XLZR[5]**
Assumed recovery (%)		70%	75%	60%	**Brine A (#1)**	**Brine C (#2)**	70%			**Brine B**	**Brine B**
Design Information											
Capacity (mgd)	1.00		1.00						0.04	0.25	
Utilization Rate (%)											
Footprint Area (ft^2) [2]											
Length (ft) x Width (ft)											
Fenced Area - ft x ft (ft^2)											
Capital Cost											
Equipment Cost (incl. Installation)	$2,050,000		$2,195,972							$8,218,922	
Interconnecting Pipework, 12.50%	$256,250		$50,000							$1,027,365	
Electrical and Instrumentation, 18%	$369,000									$1,479,406	
Building ($150/sq ft)	$225,000		$225,000							Outdoors - No Bldg	
Subtotal Construction Cost	$2,900,250		$2,470,972							$10,725,694	
Engineering, 15%	$435,038		$370,646							$1,608,854	
Legal and Admin, 10%	$290,025		$247,097							$1,072,569	
Contingency, 25%	$725,063		$617,743							$2,681,423	
Total Capital Cost	**$4,350,375**		**$4,019,117**						**$8,100,000**	**$16,088,541**	
Annual Cost of Capital	**$379,286**		**$350,405**						**$706,195**	**$1,403,672**	
O&M Cost											
Labor	$278,096		incl with Softening							$419,155	
Electricity[1]	$35,040		$410,625							$1,011,208	
Chemicals	$763,921		$6,844							$58,393	
Media / Equipment Replacement	-		$62,963							$106,905	
Waste Handling/Disposal	$259,106		-						**$4,110**	-	
Including disposal ?[2] (YES/NO)	no										
Annual O&M Costs	**$763,921**		**$480,431**						**$4,110**	**$1,595,660**	
Present Worth (2007 USD)											
Revenue Generated			**$461,251**							**$130,688**	
Capital Cost[3]	**$379,286**		**$350,405**						**$706,195**	**$1,403,672**	
O&M Cost	**$763,921**		**$480,431**						**$4,110**	**$1,595,660**	
Total Costs	**$1,143,207**		**$830,836**						**$710,305**	**$2,999,332**	

Process Combination:	5	1°RO + Softening + EDR + BC + XLZR-Disposal									
DESCRIPTION	**TREATMENT OPTION**										
	Softening	**Secondary RO**	**EDR**	**SPARRO**	**SAL-PROC[4]**		**FO**	**Evap Pond**	**SARI**	**B.C only**	**B.C and XLZR[5]**
Assumed recovery (%)		70%	75%	60%	**Brine A (#1)**	**Brine C (#2)**	70%			**Brine B**	**Brine B**
Design Information											
Capacity (mgd)	1.00		1.00								0.25
Utilization Rate (%)											
Footprint Area (ft^2) [2]											
Length (ft) x Width (ft)											
Fenced Area - ft x ft (ft^2)											
Capital Cost											
Equipment Cost (incl. Installation)	$2,050,000		$2,195,972								$10,580,163
Interconnecting Pipework, 12.50%	$256,250		$50,000								$1,322,520
Electrical and Instrumentation, 18%	$369,000										$1,904,429
Building ($150/sq ft)	$225,000		$225,000								Outdoors - No Bld
Subtotal Construction Cost	$2,900,250		$2,470,972								$13,807,113
Engineering, 15%	$435,038		$370,646								$2,071,067
Legal and Admin, 10%	$290,025		$247,097								$1,380,711
Contingency, 25%	$725,063		$617,743								$3,451,778
Total Capital Cost	**$4,350,375**		**$4,019,117**								**$20,710,670**
Annual Cost of Capital	**$379,286**		**$350,405**								**$1,805,651**
O&M Cost											
Labor	$278,096		incl with Softening								$419,155
Electricity[1]	$35,040		$410,625								$1,011,208
Chemicals	$763,921		$6,844								$71,136
Media / Equipment Replacement	-		$62,963								$155,954
Waste Handling/Disposal	$259,106		-								$508,214
Including disposal ?[2] (YES/NO)	no										
Annual O&M Costs	**$763,921**		**$480,431**								**$2,165,666**
Present Worth (2007 USD)											
Revenue Generated			**$461,251**								**$138,375**
Capital Cost[3]	**$379,286**		**$350,405**								**$1,805,651**
O&M Cost	**$763,921**		**$480,431**								**$2,165,666**
Total Costs	**$1,143,207**		**$830,836**								**$3,971,317**

Process Combination: 6 — 1°RO + Softening + EDR + BC + Evap Pond

DESCRIPTION	Softening	Secondary RO	EDR	SPARRO	SAL-PROC[4] Brine A (#1)	SAL-PROC[4] Brine C (#2)	FO	Evap Pond	SARI	B.C only Brine B	B.C and XLZR[5] Brine B
	TREATMENT OPTION										
Assumed recovery (%)		70%	75%	60%			70%				
Design Information											
Capacity (mgd)	1.00		1.00					0.04		0.25	
Utilization Rate (%)											
Footprint Area (ft^2) [2]											
Length (ft) x Width (ft)											
Fenced Area - ft x ft (ft^2)											
Capital Cost											
Equipment Cost (incl. Installation)	$2,050,000		$2,195,972							$8,218,922	
Interconnecting Pipework, 12.50%	$256,250		$50,000							$1,027,365	
Electrical and Instrumentation, 18%	$369,000									$1,479,406	
Building ($150/sq ft)	$225,000		$225,000							Outdoors - No Bldg	
Subtotal Construction Cost	$2,900,250		$2,470,972							$10,725,694	
Engineering, 15%	$435,038		$370,646							$1,608,854	
Legal and Admin, 10%	$290,025		$247,097							$1,072,569	
Contingency, 25%	$725,063		$617,743							$2,681,423	
Total Capital Cost	**$4,350,375**		**$4,019,117**					**$1,315,314**		**$16,088,541**	
Annual Cost of Capital	**$379,286**		**$350,405**					**$114,675**		**$1,403,672**	
O&M Cost											
Labor	$278,096		incl with Softening							$419,155	
Electricity[1]	$35,040		$410,625							$1,011,208	
Chemicals	$763,921		$6,844							$58,393	
Media / Equipment Replacement	-		$62,963							$106,905	
Waste Handling/Disposal	$259,106		-							-	
Including disposal ?[2] (YES/NO)	no										
Annual O&M Costs	**$763,921**		**$480,431**					**$751,685**		**$1,595,660**	
Present Worth (2007 USD)											
Revenue Generated			**$461,251**							**$130,688**	
Capital Cost[3]	**$379,286**		**$350,405**					**$114,675**		**$1,403,672**	
O&M Cost	**$763,921**		**$480,431**					**$751,685**		**$1,595,660**	
Total Costs	**$1,143,207**		**$830,836**					**$866,360**		**$2,999,332**	

Process Combination:	7	1°RO + FO + BC + SARI									
DESCRIPTION	**TREATMENT OPTION**										
	Softening	**Secondary RO**	**EDR**	**SPARRO**	**SAL-PROC[4]**		**FO**	**Evap Pond**	**SARI**	**B.C only**	**B.C and XLZR[5]**
Assumed recovery (%)		70%	75%	60%	**Brine A (#1)**	**Brine C (#2)**	70%			**Brine B**	**Brine B**
Design Information											
Capacity (mgd)							1.00		0.05	0.30	
Utilization Rate (%)											
Footprint Area (ft^2) [2]											
Length (ft) x Width (ft)											
Fenced Area - ft x ft (ft^2)											
Capital Cost											
Equipment Cost (incl. Installation)							$3,849,011			$8,858,104	
Interconnecting Pipework, 12.50%							$481,126			$1,107,263	
Electrical and Instrumentation, 18%							$692,822			$1,594,459	
Building ($150/sq ft)							$225,000			Outdoors - No Bldg	
Subtotal Construction Cost							$5,247,959			$11,559,826	
Engineering, 15%							$787,194			$1,733,974	
Legal and Admin, 10%							$524,796			$1,155,983	
Contingency, 25%							$1,311,990			$2,889,957	
Total Capital Cost							**$8,320,333**		**$8,100,000**	**$17,339,739**	
Annual Cost of Capital							**$725,405**		**$706,195**	**$1,512,757**	
O&M Cost											
Labor							$105,693			$441,106	
Electricity[1]							$410,537			$1,167,928	
Chemicals							$258,362			$64,713	
Media / Equipment Replacement							$413,609			$117,749	
Waste Handling/Disposal									**$4,932**	-	
Including disposal ?[2] (YES/NO)											
Annual O&M Costs							**$1,187,450**		**$4,932**	**$1,791,495**	
Present Worth (2007 USD)											
Revenue Generated							**$301,350**			**$156,825**	
Capital Cost[3]							**$725,405**		**$706,195**	**$1,512,757**	
O&M Cost							**$1,187,450**		**$4,932**	**$1,791,495**	
Total Costs							**$1,912,855**		**$711,127**	**$3,304,253**	

Process Combination: 8 1^{o}RO + FO + BC + XLZR-Disposal

DESCRIPTION	TREATMENT OPTION										
	Softening	Secondary RO	EDR	SPARRO	SAL-PROC[4]		FO	Evap Pond	SARI	B.C only	B.C and XLZR[5]
Assumed recovery (%)		70%	75%	60%	Brine A (#1)	Brine C (#2)	70%			Brine B	Brine B
Design Information											
Capacity (mgd)							1.00				0.30
Utilization Rate (%)											
Footprint Area (ft^2) [2]											
Length (ft) x Width (ft)											
Fenced Area - ft x ft (ft^2)											
Capital Cost											
Equipment Cost (incl. Installation)							$3,849,011				$11,497,476
Interconnecting Pipework, 12.50%							$481,126				$1,437,185
Electrical and Instrumentation, 18%							$692,822				$2,069,546
Building ($150/sq ft)							$225,000				Outdoors - No Bld
Subtotal Construction Cost							$5,247,959				$15,004,207
Engineering, 15%							$787,194				$2,250,631
Legal and Admin, 10%							$524,796				$1,500,421
Contingency, 25%							$1,311,990				$3,751,052
Total Capital Cost							**$8,320,333**				**$22,506,310**
Annual Cost of Capital							**$725,405**				**$1,962,203**
O&M Cost											
Labor							$105,693				$441,106
Electricity[1]							$410,537				$1,167,928
Chemicals							$258,362				$77,192
Media / Equipment Replacement							$413,609				$172,055
Waste Handling/Disposal											$609,521
Including disposal ?[2] (YES/NO)											
Annual O&M Costs							**$1,187,450**				**$2,467,802**
Present Worth (2007 USD)											
Revenue Generated							**$301,350**				**$166,050**
Capital Cost[3]							**$725,405**				**$1,962,203**
O&M Cost							**$1,187,450**				**$2,467,802**
Total Costs							**$1,912,855**				**$4,430,004**

Process Combination:	9	1^{o}RO + FO + BC + Evap Pond									
DESCRIPTION	TREATMENT OPTION										
	Softening	Secondary RO	EDR	SPARRO	SAL-PROC[4]		FO	Evap Pond	SARI	B.C only	B.C and XLZR[5]
Assumed recovery (%)		70%	75%	60%	Brine A (#1)	Brine C (#2)	70%			Brine B	Brine B
Design Information											
Capacity (mgd)							1.00	0.05		0.30	
Utilization Rate (%)											
Footprint Area (ft^2) [2]											
Length (ft) x Width (ft)											
Fenced Area - ft x ft (ft^2)											
Capital Cost											
Equipment Cost (incl. Installation)							$3,849,011			$8,858,104	
Interconnecting Pipework, 12.50%							$481,126			$1,107,263	
Electrical and Instrumentation, 18%							$692,822			$1,594,459	
Building ($150/sq ft)							$225,000			Outdoors - No Bldg	
Subtotal Construction Cost							$5,247,959			$11,559,826	
Engineering, 15%							$787,194			$1,733,974	
Legal and Admin, 10%							$524,796			$1,155,983	
Contingency, 25%							$1,311,990			$2,889,957	
Total Capital Cost							**$8,320,333**	**$1,578,377**		**$17,339,739**	
Annual Cost of Capital							**$725,405**	**$137,610**		**$1,512,757**	
O&M Cost											
Labor							$105,693			$441,106	
Electricity[1]							$410,537			$1,167,928	
Chemicals							$258,362			$64,713	
Media / Equipment Replacement							$413,609			$117,749	
Waste Handling/Disposal										-	
Including disposal ?[2] (YES/NO)											
Annual O&M Costs							**$1,187,450**	**$902,022**		**$1,791,495**	
Present Worth (2007 USD)											
Revenue Generated							**$301,350**			**$156,825**	
Capital Cost[3]							**$725,405**	**$137,610**		**$1,512,757**	
O&M Cost							**$1,187,450**	**$902,022**		**$1,791,495**	
Total Costs							**$1,912,855**	**$1,039,632**		**$3,304,253**	

Process Combination: **10** 1^{o}RO + SAL-PROC#1

DESCRIPTION	TREATMENT OPTION										
	Softening	**Secondary RO**	**EDR**	**SPARRO**	**SAL-PROC[4]**		**FO**	**Evap Pond**	**SARI**	**B.C only**	**B.C and XLZR[5]**
Assumed recovery (%)		70%	75%	60%	**Brine A (#1)**	**Brine C (#2)**	70%			**Brine B**	**Brine B**
					Total ZLD						
Design Information											
Capacity (mgd)					1.00						
Utilization Rate (%)											
Footprint Area (ft^2) [2]											
Length (ft) x Width (ft)											
Fenced Area - ft x ft (ft^2)											
Capital Cost											
Equipment Cost (incl. Installation)					$12,668,802						
Interconnecting Pipework, 12.50%					$1,583,600						
Electrical and Instrumentation, 18%					$2,280,384						
Building ($150/sq ft)					$450,000						
Subtotal Construction Cost					$16,982,786						
Engineering, 15%					$2,547,418						
Legal and Admin, 10%					$1,698,279						
Contingency, 25%					$4,245,696						
Total Capital Cost					**$25,474,179**						
Annual Cost of Capital					**$2,220,955**						
O&M Cost											
Labor					$350,000						
Electricity[1]					$279,000						
Chemicals					$2,099,360						
Media / Equipment Replacement					$67,000						
Waste Handling/Disposal					$3,331,147						
Including disposal ?[2] (YES/NO)											
Annual O&M Costs					**$6,551,732**						
Present Worth (2007 USD)											
Revenue Generated											
Capital Cost[3]					**$2,220,955**						
O&M Cost					**$6,551,732**						
Total Costs					**$8,772,687**						

Process Combination:	**11**	**1°RO + Softening + 2°RO + Seeded RO + SAL-PROC#2**									
DESCRIPTION	**TREATMENT OPTION**										
	Softening	**Secondary RO**	**EDR**	**SPARRO**	**SAL-PROC[4]**		**FO**	**Evap Pond**	**SARI**	**B.C only**	**B.C and XLZR[5]**
Assumed recovery (%)		70%	75%	60%	**Brine A (#1)**	**Brine C (#2)**	70%			**Brine B**	**Brine B**
						Total ZLD					
Design Information											
Capacity (mgd)	1.00	1.00		0.30		0.15					
Utilization Rate (%)											
Footprint Area (ft^2) [2]											
Length (ft) x Width (ft)											
Fenced Area - ft x ft (ft^2)											
Capital Cost											
Equipment Cost (incl. Installation)	$2,050,000	$1,000,000		$3,346,813		$7,168,672					
Interconnecting Pipework, 12.50%	$256,250	$125,000		$418,352		$896,084					
Electrical and Instrumentation, 18%	$369,000	$180,000		$602,426		$1,290,361					
Building ($150/sq ft)	$225,000	$225,000		$67,500		$67,500					
Subtotal Construction Cost	$2,900,250	$1,530,000		$4,435,091		$9,422,617					
Engineering, 15%	$435,038	$229,500		$665,264		$1,413,393					
Legal and Admin, 10%	$290,025	$153,000		$443,509		$942,262					
Contingency, 25%	$725,063	$382,500		$1,108,773		$2,355,654					
Total Capital Cost	**$4,350,375**	**$2,295,000**		**$7,047,144**		**$14,133,926**					
Annual Cost of Capital	**$379,286**	**$200,089**		**$619,242**		**$1,232,260**					
O&M Cost											
Labor	$278,096	incl with Softening		-		$350,000					
Electricity[1]	$35,040	$97,440		$59,484		$36,000					
Chemicals	$763,921	$25,550		$7,665		$801,396					
Media / Equipment Replacement	-	$58,400		$85,635		$67,000					
Waste Handling/Disposal	$259,106	-		-							
						$1,926,282					
Including disposal ?[2] (YES/NO)	no	no									
Annual O&M Costs	**$763,921**	**$181,390**		**$152,783**		**$3,245,557**					
Present Worth (2007 USD)											
Revenue Generated		**$430,501**		**$110,700**							
Capital Cost[3]	**$379,286**	**$200,089**		**$619,242**		**$1,232,260**					
O&M Cost	**$763,921**	**$181,390**		**$152,783**		**$3,245,557**					
Total Costs	**$1,143,207**	**$381,478**		**$772,025**		**$4,477,817**					

Process Combination:	12	1^{o}RO + Softening + 2^{o}RO + Seeded RO + BC + SARI									
DESCRIPTION	TREATMENT OPTION										
	Softening	Secondary RO	EDR	SPARRO	SAL-PROC[4]		FO	Evap Pond	SARI	B.C only	B.C and XLZR[5]
Assumed recovery (%)		70%	75%	60%	Brine A (#1)	Brine C (#2)	70%			Brine B	Brine B
					Total ZLD	Total ZLD					
Design Information											
Capacity (mgd)	1.00	1.00		0.30					0.02	0.15	
Utilization Rate (%)											
Footprint Area (ft^2) [2]											
Length (ft) x Width (ft)											
Fenced Area - ft x ft (ft^2)											
Capital Cost											
Equipment Cost (incl. Installation)	$2,050,000	$1,000,000		$3,346,813						$6,890,069	
Interconnecting Pipework, 12.50%	$256,250	$125,000		$418,352						$861,259	
Electrical and Instrumentation, 18%	$369,000	$180,000		$602,426						$1,240,212	
Building ($150/sq ft)	$225,000	$225,000		$67,500						Outdoors - No Bldg	
Subtotal Construction Cost	$2,900,250	$1,530,000		$4,435,091						$8,991,540	
Engineering, 15%	$435,038	$229,500		$665,264						$1,348,731	
Legal and Admin, 10%	$290,025	$153,000		$443,509						$899,154	
Contingency, 25%	$725,063	$382,500		$1,108,773						$2,247,885	
Total Capital Cost	**$4,350,375**	**$2,295,000**		**$7,047,144**					**$8,100,000**	**$13,487,310**	
Annual Cost of Capital	**$379,286**	**$200,089**		**$619,242**					**$706,195**	**$1,176,885**	
O&M Cost											
Labor	$278,096	incl with Softening		-						$373,002	
Electricity[1]	$35,040	$97,440		$59,484						$711,567	
Chemicals	$763,921	$25,550		$7,665						$44,746	
Media / Equipment Replacement	-	$58,400		$85,635						$84,704	
Waste Handling/Disposal	$259,106	-		-					**$2,468**	-	
Including disposal ?[2] (YES/NO)	no	no									
Annual O&M Costs	**$763,921**	**$181,390**		**$152,783**					**$2,468**	**$1,214,019**	
Present Worth (2007 USD)											
Revenue Generated		**$430,501**		**$110,700**						**$78,413**	
Capital Cost[3]	**$379,286**	**$200,089**		**$619,242**					**$706,195**	**$1,176,885**	
O&M Cost	**$763,921**	**$181,390**		**$152,783**					**$2,468**	**$1,214,019**	
Total Costs	**$1,143,207**	**$381,478**		**$772,025**					**$708,663**	**$2,390,904**	

Process Combination:	13	$1^oRO + Softening + 2^oRO + Seeded RO + BC + XLZR-Disposal$									
DESCRIPTION	**TREATMENT OPTION**										
	Softening	**Secondary RO**	**EDR**	**SPARRO**	**SAL-PROC[4]**		**FO**	**Evap Pond**	**SARI**	**B.C only**	**B.C and XLZR[5]**
Assumed recovery (%)		70%	75%	60%	**Brine A (#1)**	**Brine C (#2)**	70%			**Brine B**	**Brine B**
Design Information											
Capacity (mgd)	1.00	1.00		0.30							0.15
Utilization Rate (%)											
Footprint Area (ft^2) [2]											
Length (ft) x Width (ft)											
Fenced Area - ft x ft (ft^2)											
Capital Cost											
Equipment Cost (incl. Installation)	$2,050,000	$1,000,000		$3,346,813							$8,648,491
Interconnecting Pipework, 12.50%	$256,250	$125,000		$418,352							$1,081,061
Electrical and Instrumentation, 18%	$369,000	$180,000		$602,426							$1,556,728
Building ($150/sq ft)	$225,000	$225,000		$67,500							Outdoors - No Bld
Subtotal Construction Cost	$2,900,250	$1,530,000		$4,435,091							$11,286,280
Engineering, 15%	$435,038	$229,500		$665,264							$1,692,942
Legal and Admin, 10%	$290,025	$153,000		$443,509							$1,128,628
Contingency, 25%	$725,063	$382,500		$1,108,773							$2,821,570
Total Capital Cost	**$4,350,375**	**$2,295,000**		**$7,047,144**							**$16,929,421**
Annual Cost of Capital	**$379,286**	**$200,089**		**$619,242**							**$1,475,984**
O&M Cost											
Labor	$278,096	incl with Softening		-							$373,002
Electricity[1]	$35,040	$97,440		$59,484							$711,567
Chemicals	$763,921	$25,550		$7,665							$58,376
Media / Equipment Replacement	-	$58,400		$85,635							$122,644
Waste Handling/Disposal	$259,106	-		-							$306,152
Including disposal ?[2] (YES/NO)	no	no									
Annual O&M Costs	**$763,921**	**$181,390**		**$152,783**							**$1,571,741**
Present Worth (2007 USD)											
Revenue Generated		**$430,501**		**$110,700**							**$83,025**
Capital Cost[3]	**$379,286**	**$200,089**		**$619,242**							**$1,475,984**
O&M Cost	**$763,921**	**$181,390**		**$152,783**							**$1,571,741**
Total Costs	**$1,143,207**	**$381,478**		**$772,025**							**$3,047,725**

Process Combination:	14	1^oRO + Softening + 2^oRO + Seeded RO + BC + Evap Pond									
DESCRIPTION	**TREATMENT OPTION**										
	Softening	**Secondary RO**	**EDR**	**SPARRO**	**SAL-PROC**[4]		**FO**	**Evap Pond**	**SARI**	**B.C only**	**B.C and XLZR**[5]
Assumed recovery (%)		70%	75%	60%	**Brine A (#1)**	**Brine C (#2)**	70%			**Brine B**	**Brine B**
Design Information											
Capacity (mgd)	1.00	1.00		0.30				0.02		0.15	
Utilization Rate (%)											
Footprint Area (ft^2) [2]											
Length (ft) x Width (ft)											
Fenced Area - ft x ft (ft^2)											
Capital Cost											
Equipment Cost (incl. Installation)	$2,050,000	$1,000,000		$3,346,813						$6,890,069	
Interconnecting Pipework, 12.50%	$256,250	$125,000		$418,352						$861,259	
Electrical and Instrumentation, 18%	$369,000	$180,000		$602,426						$1,240,212	
Building ($150/sq ft)	$225,000	$225,000		$67,500						Outdoors - No Bldg	
Subtotal Construction Cost	$2,900,250	$1,530,000		$4,435,091						$8,991,540	
Engineering, 15%	$435,038	$229,500		$665,264						$1,348,731	
Legal and Admin, 10%	$290,025	$153,000		$443,509						$899,154	
Contingency, 25%	$725,063	$382,500		$1,108,773						$2,247,885	
Total Capital Cost	**$4,350,375**	**$2,295,000**		**$7,047,144**				**$789,188**		**$13,487,310**	
Annual Cost of Capital	**$379,286**	**$200,089**		**$619,242**				**$68,805**		**$1,176,885**	
O&M Cost											
Labor	$278,096	incl with Softening		-						$373,002	
Electricity[1]	$35,040	$97,440		$59,484						$711,567	
Chemicals	$763,921	$25,550		$7,665						$44,746	
Media / Equipment Replacement	-	$58,400		$85,635						$84,704	
Waste Handling/Disposal	$259,106	-		-						-	
Including disposal ?[2] *(YES/NO)*	no	no									
Annual O&M Costs	**$763,921**	**$181,390**		**$152,783**				**$451,011**		**$1,214,019**	
Present Worth (2007 USD)											
Revenue Generated		**$430,501**		**$110,700**						**$78,413**	
Capital Cost[3]	**$379,286**	**$200,089**		**$619,242**				**$68,805**		**$1,176,885**	
O&M Cost	**$763,921**	**$181,390**		**$152,783**				**$451,011**		**$1,214,019**	
Total Costs	**$1,143,207**	**$381,478**		**$772,025**				**$519,816**		**$2,390,904**	

Notes:

1. Electrical power cost of $0.12/kWh assumed

2. User chooses whether or not to include disposal costs (sludge for Softening, brine for RO - SARI, and solids for BC/XLZR)

3. Annual costs assume 6% interest rate, and 20 yr loan period
4. SAL-PROC: ZLD includes solids disposal in O&M
5. Assume solids disposal of $50/ton
6. Labor cost is priced at $47.75hr.
7. Soda ash is priced at $0.165/LB
8. Lime is priced at $300/ton
9. Polymer & RO Reagents are priced at $0.07/1000 gal
10. Natural gas is priced at $7/GJ
11. RO Membranes priced at $0.16/1000 gal

Recalculation of SAL-PROC#1 according to Carollo's project cost estimates
(Brine A, 1 MGD, ZLD only)

<u>Capital Costs</u>

Item	Ref	Quantity	Material Cost $ m	Delivery Cost (2.5% of Material) $	Installation Cost (15% of Material) $	Capex, $
SEPCON Plant Equipment	$3.1M/ mgd(A)	Total Treatment System	$3,100,000		$465,000	$3,565,000
RO Desalter #1	$1.1M/mgd(A)	1	$1,000,000		included	$1,000,000
RO Desalter #2	$3.3M/mgd, at 0.24 mgd		$240,000		included	$240,000
BC + XLZR	0.16 mgd into BC, 0.023 mgd into XLZR	1	$7,863,802		included	$7,863,802
				Subtotal for Equipment:		$12,668,802
				Interconnecting Pipework		$1,583,600
				Electrical & Instrumentation		$2,280,384
Buildings	$150/sq.ft	3000 sq ft	$450,000			$450,000
				Subtotal for Construction:		$16,982,786
				Engineering and Instrumentation		$2,547,418
				Legal and administration		$1,698,279
				Contingency		$4,245,696
				Estimated Project Cost		**$25,474,179**
	Annualized Capital Cost @	6%	20	yrs	=	**$2,220,955**

<u>O&M Costs</u>		Unit	Annual Cost, $
Raw Materials:			
	Polymer and RO Reagents	None needed for Brine A	$0
	Soda Ash, ton	4115 pa, $364/unit	$1,497,860
	Quicklime, ton	2005 pa, $300/unit	$601,500
	RO Membranes		$67,000
		Total Raw Material Cost	*$2,166,360*
Utilities:	Electricity, MWh	1800 @ $120/unit	$216,000
	Gas, GJ	9000 at $7/unit	$63,000
	Process water, ML		$0
		Total Utilities Cost	*$279,000*
Total Process Labor			
Wages:		5 at $70000 per	$350,000
Payroll Overheads:		22% Total Process Labor	$77,000
Maintenance		4% Capital	$1,018,967
Operating supplies		15% Total Process labor	$52,500
Plant Overheads		25% (process labor + maintenance)	$342,242
Insurance		1% capital	$254,742
Book Depreciation		5% capital	$1,273,709
Solids Disposal		$50/ton at 23.3 tons/mgd	$425,225
		Sub-total	*$6,239,745*
Non-manufacturing Costs			
	R&D Selling Prices	5% (subtotal from above)	*$311,987*
Total O&M Costs for Brine A --			***$6,551,732***

(Brine C, 0.15 MGD, ZLD only)
<u>Capital Costs</u>

Item	Ref	Quantity	Material Cost $ m	Delivery Cost (2.5% of Material) $	Installation Cost (15% of Material) $	Capex, $
SEPCON Plant Equipment	$2.5M/ mgd(C)	Total Treatment System	$375,000		$56,250	$431,250
RO Desalter #1	$1.4M/mgd(C)	1	$150,000		included	$150,000
BC + XLZR	Assum 67% Recovery from RO, therefore 0.05 MGD going into the BC/XLZR system	1	$6,587,422		included	$6,587,422
				Subtotal for Equipment:		$7,168,672
				Interconnecting Pipework		$896,084
				Electrical & Instrumentation		$1,290,361
Buildings	$150/sq.ft	15% of 3000 sq ft	$67,500			$67,500
				Subtotal for Construction:		$9,422,617
		4.052		Engineering and Instrumentation		$1,413,393
				Legal and administration		$942,262
				Contingency		$2,355,654
				Estimated Project Cost		**$14,133,926**
	Annualized Capital Cost @	6%	20	yrs	=	**$1,232,260**

<u>O&M Costs</u>		Unit	Annual Cost, $
Raw Materials:			
	Polymer and RO Reagents	75 pa/MGD @ $400/UNIT	$4,500
	Soda Ash, ton	11385 pa/MGD, $364/unit	$621,621
	Quicklime, ton	3895 pa/MGD, $300/unit	$175,275
	RO Membranes		$67,000
		Total Raw Material Cost	*$868,396*
Utilities:	Electricity, MWh	1500/mgd @ $120/unit	$27,000
	Gas, GJ	20000/mgd at $7/unit	$9,000
	Process water, ML		$0
		Total Utilities Cost	*$36,000*
Total Process Labor			
Wages:		5 at $70000 per	$350,000
Payroll Overheads:		22% Total Process Labor	$77,000
Maintenance		4% Capital	$565,357
Operating supplies		15% Total Process labor	$52,500
Plant Overheads		25% (process labor + maintenance)	$228,839
Insurance		1% capital	$141,339
Book Depreciation		5% capital	$706,696
Solids Disposal		$50/ton at 23.7tons/mgd	$64,879
		Sub-total	*$3,091,007*
Non-manufacturing Costs			
	R&D Selling Prices	5% (subtotal from above)	*$154,550*
Total O&M Costs for Brine C --			***$3,245,557***

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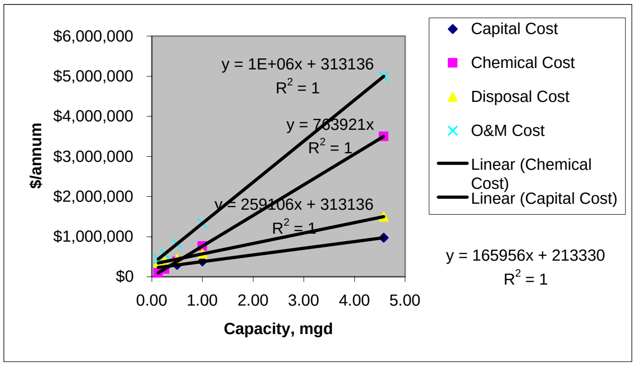
SOFTENING COST ESTIMATES

Variables		**Units**					
- Flow from 1^o RO to softening system		mgd	**4.58**	**1.00**	**0.5**	**0.25**	**0.125**
		gpm	**3180**	**694**	**347**	**174**	**87**
Equipment Costs:							
- Lime softerning equipment		$	$3,663,565	$800,000	$400,000	$200,000	$100,000
- lime dosing system		$	$750,000	$750,000	$750,000	$750,000	$750,000
- acid&soda ash dosing system		$	$500,000	$500,000	$500,000	$500,000	$500,000
- *Subtotal1 for equipment cost*		**$**	**$4,913,565**	**$2,050,000**	**$1,650,000**	**$1,450,000**	**$1,350,000**
- Interconnecting pipework	12.50%	$	$614,196	$256,250	$206,250	$181,250	$168,750
- Electrical and instrumentation	18%	$	$884,442	$369,000	$297,000	$261,000	$243,000
- Building @ $150/sq ft and 1500 sq ft/mgd		$	$1,030,378	$225,000	$112,500	$56,250	$28,125
- *Subtotal2 for construction costs*		**$**	**$7,442,581**	**$2,900,250**	**$2,265,750**	**$1,948,500**	**$1,789,875**
- Engineering	15%	$	$1,116,387	$435,038	$339,863	$292,275	$268,481
- Legal and Admin	10%	$	$744,258	$290,025	$226,575	$194,850	$178,988
- Contingency	25%	$	$1,860,645	$725,063	$566,438	$487,125	$447,469
- Estimated Project Cost		**$**	**$11,163,871**	**$4,350,375**	**$3,398,625**	**$2,922,750**	**$2,684,813**
Annual Cost of Capital:							
- Average Annual Interest Rate		%	6%	6%	6%	6%	6%
- Loan Period		years	20	20	20	20	20
- Annual Payment		**$/annum**	**$973,317**	**$379,286**	**$296,308**	**$254,819**	**$234,074**
- Annual Payment		$/AF					
Chemicals:							
-							
Lime:							
- Dosage		mg/L	555	555	555	555	555
- Lime required (100% chemical)		lb/day	21196.93	4628.70	2314.35	1157.18	578.59
- Lime purity($Ca(OH)_2$)		%	98.5%	98.5%	98.5%	98.5%	98.5%
- Lime required as chemical		lb/day	21519.73	4699.19	2349.59	1174.80	587.40
- Cost of lime		$/ton	$300	$300	$300	$300	$300
- - Annual cost of lime		**$/yr**	**$1,178,205**	**$257,281**	**$128,640**	**$64,320**	**$32,160**
Soda Ash:							
- Dosage		mg/L	150	150	150	150	150
- Soda ash required (100% chemical)		lb/day	5728.90	1251.00	625.50	312.75	156.38
- Soda ash purity (Na_2CO_3), solution		%	15%	15%	15%	15%	15%
- Soda ash required as chemical		lb/day	38193	8340	4170	2085	1043
- Soda ash solution required		gal/day	3954	863	432	216	108
- Cost of soda ash		$/gal	$1.594	$1.594	$1.594	$1.594	$1.594
- - Annual cost of soda ash		**$/yr**	**$2,300,154**	**$502,277**	**$251,138**	**$125,569**	**$62,785**
Acid:							
- Dosage		mg/L	20	20	20	20	20
- acid required (100% chemical)		lb/day	764	167	83	42	21
- acid purity (H_2SO_4)		%	93%	93%	93%	93%	93%
- Acid required as chemical		lb/day	821.35	179.35	89.68	44.84	22.42
- Acid solution required		gal/day	55	12	6	3	1
- Cost of acid		$/gal	1	1	1	1	1
- Annual cost of acid		**$/yr**	**$19,986**	**$4,364**	**$2,182**	**$1,091**	**$546**
- Total cost for chemicals		**$/annum**	**$3,498,345**	**$763,921**	**$381,961**	**$190,980**	**$95,490**
Chemical Sludge Disposal:							
- Raw brine flowrate		gpm	3,180	694	347	174	87
- Average Ca in raw brine		mg/L	1,080	1,080	1,080	1,080	1,080
- Average Ca in Softened brine		mg/L	480	480	480	480	480
- Average Ca removed by softening		mg/L	600	600	600	600	600
- Moles Ca removed		mmol/L	15	15	15	15	15
- Mass CaCO3 formed		mg/L	1500	1500	1500	1500	1500
- Mass CaCO3 formed		lb/day	57,289	12,510	6,255	3,128	1,564
- Add the mass of lime added		lb/day	21,520	4,699	2,350	1,175	587
- Estimated total mass of solids formed		lb/day	86,690	18,930	9,465	4,733	2,366
- Dewatered sludge % solids		%	40%	40%	40%	40%	40%
- Mass of wet sludge for disposal		ton/day	108	24	12	6	3
- Disposal cost for chemical sludge		$/ton	30	30	30	30	30
- Annual Cost for Sludge Disposal		$	**$1,186,564**	**$259,106**	**$129,553**	**$64,776**	**$32,388**
- NPV for Sludge Disposal		20 yrs					
- Power requirements for belt press		kw	100	100	100	100	100
- Belt Press Hours of operation per day		hr	8	8	8	8	8
- Annual cost for power		$	**$35,040**	**$35,040**	**$35,040**	**$35,040**	**$35,040**
- No. Shifts Required per Week			14	14	14	14	14
- No. Hours per Shift		hr/shift	8	8	8	8	8
- Labor Cost per Hour		$/hr (incl. Benefits	$ 47.75	$ 47.75	$ 47.75	$ 47.75	$ 47.75
- Total Hours per Year - Labor		hr/yr	5824	5824	5824	5824	5824
- Total Laber Costs per Year		$/annum	**$ 278,096**	**$ 278,096**	**$ 278,096**	**$ 278,096**	**$ 278,096**
- Total cost for sludge disposal		**$/annum**	**$1,499,700**	**$572,242**	**$442,689**	**$377,912**	**$345,524**
- Total O&M Cost (chemicals + disposal)		**$/annum**	**$4,998,045**	**$1,336,163**	**$824,650**	**$568,893**	**$441,014**
- Total O&M Cost (chemicals only)		**$/annum**	**$3,498,345**	**$763,921**	**$381,961**	**$190,980**	**$95,490**



Annualized Capital Cost Curve			
y=	165955.8042	x+	213329.7129
Chemical Cost Curve			
y=	763921.3341	x+	0
Disposal Cost Curve			
y=	259105.8341	x+	313136
Total O&M Cost Curve (with disposal)			
y=	1023027.168	x+	313136
Total O&M Cost Curve (without disposal)			
y=	**763921.3341**	x+	0
Total Capital Cost Curve			
y=	1903500	x+	2446875

[redacted]

SECONDARY RO COST ESTIMATES

RO Power Cost:						
- RO Feed flowrate	mgd	**4.58**	**1**	**0.5**	**0.25**	**0.125**
- RO Feed flow	gpm	**3180**	**694**	**347**	**174**	**87**
- Average operating pressure	psi	215	215	215	215	215
- Pumping power	hp	399	87	44	22	11
- Overall efficiency	%	70%	70%	70%	70%	70%
- Power consumed	kWh/yr	3718511	811998	405999	203000	101500
- Power cost	$/Kwh	$0.12	$0.12	$0.12	$0.12	$0.12
- Power cost	$/yr	**$446,221**	**$97,440**	**$48,720**	**$24,360**	**$12,180**
- Power cost on per 1000 gal basis	$/1000 gal	$0.27	$0.27	$0.27	$0.27	$0.27
Membrane Replacement:						
- Assume typical life	$/1000 gal	$0.16	$0.16	$0.16	$0.16	$0.16
- Annual cost for membranes	$/annum	**$267,440**	**$58,400**	**$29,200**	**$14,600**	**$7,300**
Chemicals (Antiscalants, cleaning solution, etc):						
- Assume typical operation	$/1000 gal	0.07	0.07	0.07	0.07	0.07
- Annual cost for RO Chemicals	$	**117,005**	**25,550**	**12,775**	**6,388**	**3,194**
- Total O&M Cost for secondary RO	$/annum	**$830,667**	**$181,390**	**$90,695**	**$45,347**	**$22,674**
Capital Cost:						
- Unit cost of RO Equipment	$/gal	1.0	1	1	1	1
- Additional cost of installation	$	0	0	0	0	0
- Capital cost of RO Equipment	$	$4,579,457	$1,000,000	$500,000	$250,000	$125,000
Subtotal for equipment costs	$	$4,579,457	$1,000,000	$500,000	$250,000	$125,000
- Interconnecting pipework	12.50%	$572,432	$125,000	$62,500	$31,250	$15,625
- Electrical and instrumentation	18%	$824,302	$180,000	$90,000	$45,000	$22,500
- Building @ $150/sq ft and 1500 sq ft/mgd		$1,030,378	$225,000	$112,500	$56,250	$28,125
Subtotal for construction costs	$	$7,006,569	$1,530,000	$765,000	$382,500	$191,250
- Engineering	15%	$1,050,985	$229,500	$114,750	$57,375	$28,688
- Legal and Administration	10%	$700,657	$153,000	$76,500	$38,250	$19,125
- Contingency	25%	$1,751,642	$382,500	$191,250	$95,625	$47,813
Estimated Project Cost	$	**$10,509,853**	**$2,295,000**	**$1,147,500**	**$573,750**	**$286,875**
Annual Cost of Capital:						
- Average Annual Interest Rate	%	6%	6%	6%	6%	6%
- Loan Period	years	20	20	20	20	20
- Annual Payment	$	**$916,297**	**$200,089**	**$100,044**	**$50,022**	**$25,011**
- Annual Payment	$/AF					
Revenue from New Water Production:						
- Assumed 2^o RO recovery	%	70%	70%	70%	70%	70%
- Potable water produced	mgd	3.21	0.70	0.35	0.18	0.09
- Potable water produced	AF/yr	3,591	784	392	196	98
- Selling price of new water	$/AF	549	549	549	549	549
- Revenue generated	$/annum	**$1,971,459**	**$430,501**	**$215,250**	**$107,625**	**$53,813**
Disposal Cost of Remaining Brine:						
- remaining brine flowrate	mgd	1.37	0.30	0.15	0.08	0.04
OCSD Treatment Capacity Cost	$/mgd	$4,300,000	$4,300,000	$4,300,000	$4,300,000	$4,300,001
- EMWD Total treatment cap needed	mgd	1.37	0.30	0.15	0.08	0.04
- Treatment cost	$	$5,907,499	$1,290,000	$645,000	$322,500	$161,250
SARI Line Capacity Cost	$/mgd	$3,800,000	$3,800,000	$3,800,000	$3,800,000	$3,800,001
- Capacity needed	mgd	1.37	0.30	0.15	0.08	0.04
- SARI Line cost	$	$5,220,581	$1,140,000	$570,000	$285,000	$142,500
Total Capital Expenditure	$	$11,128,080	$2,430,000	$1,215,000	$607,500	$303,750
- Interest Rate	%	6%	6%	6%	6%	6%
- Loan Period	years	20	20	20	20	20
- Annual payment	$	**$970,197**	**$211,858**	**$105,929**	**$52,965**	**$26,482**
Disposal Cost						
- OCSD Cost	$/mil gal	$1,000	$1,000	$1,000	$1,000	$1,001
- Brine Volume for disposal	mgd	1.37	0.30	0.15	0.08	0.04
- Disposal Cost	$/day	$1,374	$300	$150	$75	$38
- Disposal Cost per year	$/year	$501,451	$109,500	$54,750	$27,375	$13,701
Total annual disposal cost						
- (assuming excess capacity sold)	$/yr	**$1,471,647**	**$321,358**	**$160,679**	**$80,340**	**$40,184**
(assume capacity not sold, excl capital)	$/yr	**$1,207,645**	**$815,695**	**$760,945**	**$733,570**	**$719,896**
Total O&M Cost with disposal	$/yr	**$2,302,314**	**$502,748**	**$251,374**	**$125,687**	**$62,857**

SEEDED RO COST ESTIMATE

Item	Part Number	Price per	Quantity	Total	
SelRO Tubular MPT-34 module		$3,000.00	122	$366,000.00	
Replacement membranes		$875.00	1	$875.00	
U-bend connector	3012000	$90.00	0	$0.00	(used to be $55)
Inlet/Outlet adapter	3012002	$85.00	1320	$0.00	

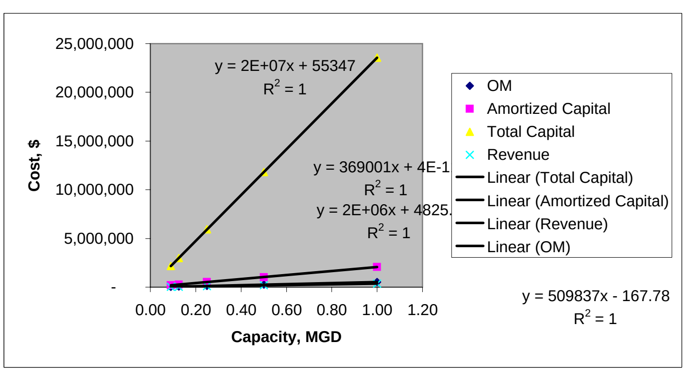
Further Concentration of Secondary Brine with Seeded RO

Cost Constants:

-	Electrical Power ($/kWh)	0.12
-	Sludge disposal Cost ($/t)	50
-	Selling Price of New Water ($/AF)	549
-	SARI Line Disposal Cost ($/mil gal)	1,100
-	Retail Value of Potable Water ($/AF)	727.25

$7.7 M for plant with 1500 modules @ $3000 each. So, membranes = $4.5 M; therefore ancillaries = $3.2 M
With 800 modules = $2.4 M for membrane => $5.6 M for plant

Seeded RO (SPARRO):			**Capacity --> (mgd)**	**0.09**	**1.00**	**0.50**	**0.25**	**0.13**
	Power Cost:							
(max op. pressure 510 psi)	-	Feed Pressure (psi)		500	500	500	500	500
	-	Feed Flowrate (mgd)		0.09	1.00	0.50	0.25	0.13
	-	Feed Flowrate (gpm)		63	694	347	174	87
	-	Pump horsepower (Hp)		18.23	202.55	101.27	50.64	25.32
	-	Efficiency (%)		80%	80%	80%	80%	80%
	-	Pump Power (kW)		17	189	94	47	24
	-	Power per year (kWh/y)		148,709	1,652,322	826,161	413,081	206,540
	-	Power cost per year ($)		**17,845**	**198,279**	**99,139**	**49,570**	**24,785**
	-	Power costs as $/1000 gal		0.54	0.54	0.54	0.54	0.54
	Membrane Replacement:							
	-	Assumed membrane life (years)		3	3	3	3	3
	-	No. modules in plant		88	980	490	244	122
	-	Replacement cost of membranes per module ($)		$875.00	$875.00	$875.00	$875.00	$875.00
	-	Annual Cost for Membranes ($)		**25,667**	**285,833**	**142,917**	**71,167**	**35,583**
	Chemicals (Antiscalant, cleaning solution etc):							
	-	Assume Typical operation ($/1000 gal)		0.07	0.07	0.07	0.07	0.07
	-	Annual Cost for RO Chemicals ($)		**2,300**	**25,550**	**12,775**	**6,388**	**3,194**
-	Total O&M Cost for Secondary RO ($/annum)			**45,811**	**509,662**	**254,831**	**127,124**	**63,562**
	Capital Cost:							
	Capital Cost of membranes			$264,000	$2,940,000	$1,470,000	$732,000	$366,000
	-	RO Feed Flowrate (mgd)		0.09	1.00	0.50	0.25	0.13
	-	Unit cost of RO Equipment ($/gal)		5.6	5.6	5.6	5.6	5.6
	-	Additional Cost for Installation (million $)		0.2268	2.52	1.26	0.63	0.315
	-	Capital Cost of RO Equipment ($)		730,800	8,120,000	4,060,000	2,030,000	1,015,000
	-	Chemical cleaning System		30,000	30,000	30,000	30,000	30,000
	-	Sub-total 1 Treatment Equipment ($)		1,024,800	11,090,000	5,560,000	2,792,000	1,411,000
	-	Interconnecting Pipework	12.50%	128,100	1,386,250	695,000	349,000	176,375
	-	Electrical and Instrumentatic	18%	184,464	1,996,200	1,000,800	502,560	253,980
	-	Building (@ $150/sq ft and 8167 sq ft/mgd))		110,255	1,225,050	612,525	306,263	153,131
	-	Sub-total 2 Construction Cost Estimate ($)		1,447,619	15,697,500	7,868,325	3,949,823	1,994,486
	-	Engineering	15%	217,143	2,354,625	1,180,249	592,473	299,173
	-	Legal and Admin	10%	144,762	1,569,750	786,833	394,982	199,449
	-	Contingency	25%	361,905	3,924,375	1,967,081	987,456	498,622
	-	Estimated Project Cost ($)		**2,171,428**	**23,546,250**	**11,802,488**	**5,924,734**	**2,991,729**
	Annual Cost of Capital:							
	-	Average Annual Interest Rate (%)		6%	6%	6%	6%	6%
	-	Loan Period (years)		20	20	20	20	20
	-	Annual Payment ($)		**$189,315**	**$2,052,869**	**$1,028,995**	**$516,545**	**$260,833**
	-	Annual Payment ($/AF)						
Revenue from New Water Production:								
	-	Assumed SPARRO Recovery (%)		60%	60%	60%	60%	60%
	-	Potable Water Produced (mgd)		0.05	0.60	0.30	0.15	0.08
	-	Potable Water Produced (AF/y)		60	672	336	168	84
	-	Selling price of New Water ($/AF)		549	549	549	549	549
	-	Revenue Generated ($/annum)		**$ 33,210**	**$ 369,001**	**$ 184,500**	**$ 92,250**	**$ 46,125**



Annualized Capital Cost Curve			
y=	2048055.94	x+	4825.389648
O&M Cost Curve			
y=	509836.572	x+	-167.781511
Revenue Cost Curve			
y=	369000.53	x+	2.91038E-11
Total Capital Cost Curve			
y=	23491040.3	x+	-167.781511

[redacted] EDR COST ESTIMATE

Goal: Produce 450 ppm TDS water (including silica), at 75% water recovery, using worse case 850µS feedwater to EDR

Estimated Budget Capital Costs						
Resulting feed, mgd	**6.67**	**5.00**	**1.33**	**0.67**	**0.33**	**0.17**
	Product Water Capacity (mgd)					
	5[a]	**3.75**[a]	**1**	**0.5**	**0.25**	**0.125**
EDR	$9,000,000	$7,200,000	$1,900,000	$1,600,000	$830,000	$580,000
Installation, % of capital	30.0%	30.0%	28.0%	28.0%	13.0%	13.0%
Installation, $	$2,700,000	$2,200,000	$540,000	$450,000	$110,000	$75,000
equipment subtotal	*$11,850,000*	*$9,400,000*	*$2,440,000*	*$2,050,000*	*$940,000*	*$655,000*
Building Cost (@$100/sq-ft)	$1,650,000	$1,300,000	$320,000	$290,000	$100,000	$80,000
Revised Bldg Cost for Carollo	$2,475,000	$1,950,000	$480,000	$435,000	$150,000	$120,000
Ancillary Items	$150,000	$150,000	$50,000	$50,000	$25,000	$20,000
Sub-total EDR Costs	$14,475,000	$11,500,000	$2,970,000	$2,535,000	$1,115,000	$795,000
To calculate for Carollo Project Cost,						
Engineering (15%)	$2,171,250	$1,725,000	$445,500	$380,250	$167,250	$119,250
Legal and Admin (10%)	$1,447,500	$1,150,000	$297,000	$253,500	$111,500	$79,500
Contingency (25%)	$3,618,750	$2,875,000	$742,500	$633,750	$278,750	$198,750
Estimated project cost	**$21,712,500**	**$17,250,000**	**$4,455,000**	**$3,802,500**	**$1,672,500**	**$1,192,500**

Annual Cost of Capital:						
- Average Annual Interest Rate (%)	6%	6%	6%	6%	6%	6%
- Loan Period (years)	20	20	20	20	20	20
- Annual Payment ($)	**$1,892,995**	**$1,503,934**	**$388,407**	**$331,519**	**$145,816**	**$103,968**
- Annual Payment ($/AF)						

Estimated O&M costs (for a single EDR system - EDR 2020 at 1.25 mgd product flow and 75% recovery)

	O&M Costs in Dollars per 1,000 gal Reclaimed water						
1 Consumable Costs, $ per year		**5**	**3.75**	**1**	**0.5**	**0.25**	**0.125**
- Electrical Energy ($0.10/kwh)	$1.250	$2,281,250	$1,710,938	$456,250	$228,125	$114,063	$57,031
Carollo Electrical ($012/kwh)		$2,737,500	$2,053,125	$547,500	$273,750	$136,875	$68,438
- Chemical Consumption	$0.025	$45,625	$34,219	$9,125	$4,563	$2,281	$1,141
- Filter Cartridge Replacement	$0.040	$73,000	$54,750	$14,600	$7,300	$3,650	$1,825
Sub-total	$1.315	$2,856,125	$2,142,094	$571,225	$285,613	$142,806	$71,403
2 Long-Term Reserves							
- EDR Membrane Replacement	$0.100	$182,500	$136,875	$36,500	$18,250	$9,125	$4,563
- Other EDR Parts Replacement	$0.090	$164,250	$123,188	$32,850	$16,425	$8,213	$4,106
Sub-total	$0.190	$346,750	$260,063	$69,350	$34,675	$17,338	$8,669
Carollo Estimate - Parts & Replacement		*$419,750*	*$314,813*	*$83,950*	*$41,975*	*$20,988*	*$10,494*
3 **Sub-total EDR O&M Costs**	**$1.505**	**$3,202,875**	**$2,402,156**	**$640,575**	**$320,288**	**$160,144**	**$80,072**

Revenue from New Water Production:							
1 Potable Water Produced (mgd)	@75% recovery	5	3.75	1	0.5	0.25	0.125
2 Potable Water Produced (AF/y)		5,601	4,201	1,120	560	280	140
3 Selling price of New Water ($/AF)		549	549	549	549	549	549
4 Revenue Generated ($/annum)		**$3,075,004**	**$2,306,253**	**$615,001**	**$307,500**	**$153,750**	**$76,875**

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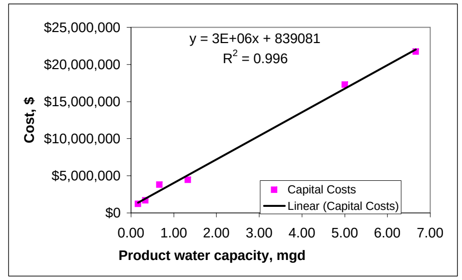
Equipment Subtotal:
y= 1733775 x+ 462196.7411

Ancillary Items:
y= 21143.03 x+ 24245.62462

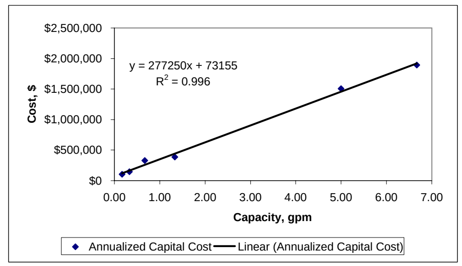
Electricity:
y= 410625 x+ 0

Chemicals:
y= 6843.75 x+ -5.45697E-12

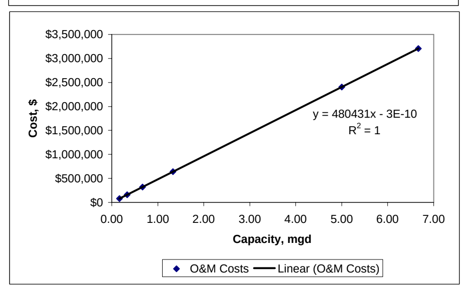
Replacement Parts
y= 62962.5 x+ 2.91038E-11



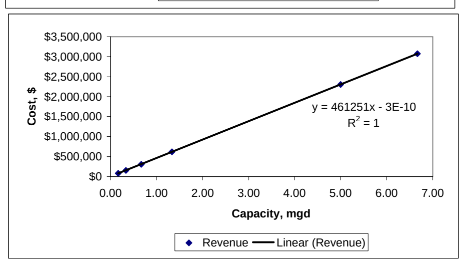
Total Capital Cost Curve
y= 3180036 x+ 839081.1708



Annualized Capital Cost Curve
y= 277250 x+ 73154.92014



O&M Cost Curve
y= 480431.3 x+ 0



Revenue from water sale
y= 461250.7 x+ -2.32831E-10

FORWARD OSMOSIS COST ESTIMATE

FO Subsystem

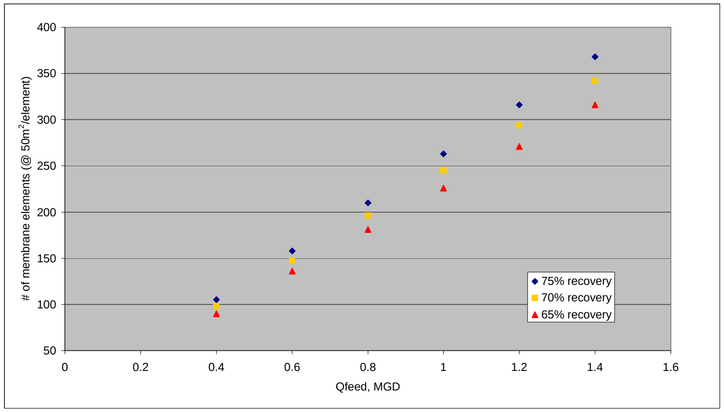
Q_{feed}			# of pumps	Power (W)	$Q_{permeare}$ (MGD)			kWh/m^3			# of FO elements (@ 50 m^2)		
MGD	gpm	lpm	@ 70 gpm/pump	(1,200 W/pump)	R = 75%	R = 70%	R = 65%	R = 75%	R = 70%	R = 65%	R = 75%	R = 70%	R = 65%
1.4	972	3,680	14	16,800	1.05	0.98	0.91	0.101	0.109	0.117	368	342	316
1.2	833	3,154	12	14,400	0.9	0.84	0.78	0.101	0.109	0.117	316	294	271
1	694	2,628	10	12,000	0.75	0.7	0.65	0.101	0.109	0.117	263	245	226
0.8	556	2,103	8	9,600	0.6	0.56	0.52	0.101	0.109	0.117	210	196	181
0.6	417	1,577	6	7,200	0.45	0.42	0.39	0.101	0.109	0.117	158	147	136
0.4	278	1,051	4	4,800	0.3	0.28	0.26	0.101	0.109	0.117	105	98	90

$Q_{permeare}$ (m^3/hr)		
R = 75%	R = 70%	R = 65%
165.6	154.6	143.5
141.9	132.5	123.0
118.3	110.4	102.5
94.6	88.3	82.0
71.0	66.2	61.5
47.3	44.2	41.0

RO Subsystem

FO Q_{feed}	FO Q_{perm}	# of RO elements	pressure	power consumption	specific energy	$C_{permeate}$
MGD	R = 75%	SW30HRLE-400i	psig	kW	kWh/m^3	(ppm)
1.00	0.75	24x6	680	282	2.38	114
0.60	0.45	14x6	686	171	2.41	110
0.40	0.30	9x6	695	115	2.44	106

FO Q_{feed}	FO Q_{perm}	# of RO elements	pressure	power consumption	specific energy	$C_{permeate}$	Qperm
MGD	R = 70%	SW30HRLE-400i	psig	kW	kWh/m^3	ppm	m^3/hr
1.00	0.70	22x6	680	264	2.40	112	157.725476
0.60	0.42	13x6	687	160	2.41	110	94.6352856
0.40	0.28	9x6	677	105	2.38	114	63.0901904



Cost estimate assumptions:

Pump	$	$400			
RO Mem. Element	$	$550			
FO Mem. Element	$/ft2	$5	$/m^2		$53.82
FO Mem. Area	m^2/elem	50			
Unit RO Mem. Replacement	$/m^3	$0.01			
Power cost	$/kWh	$0.12			
Net Energy Cost	$/m^3	$0.21			
Labor Cost (for RO)	$/m^3	$0.05			
Chemical Cost (for RO)	$/m^3	$0.11	$/1000 gal		$0.07
Hours of operation	hr/yr	8,760			

Assuming 70% recovery for FO and RO,
Carollo Proj. Cost Estimate:

MGD	Unit	FO Pump Cost	FO Mem.Elem. Cost	RO Subsystem Mem.Elem. Cost	Equip. Instal.	Subtotal	Interconnecting Pipework, 12.50%	Electrical and Instrumentation, 18%	Building ($150/sq ft and 3500 sq ft/mgd)	Subtotal-Construction	Engineering, 15%	Legal and Admin, 10%	Contingency, 25%	Total Capital	Annualized Capital, $/yr
1.4	$	$5,600	$920,314	$4,200,000	$259,256	$5,385,170	$673,146	$969,331	$735,000	$7,762,647	$1,164,397	$776,265	$1,940,662	$11,643,971	$1,015,174
1.2	$	$4,800	$791,147	$3,600,000	$222,865	$4,618,813	$577,352	$831,386	$630,000	$6,657,551	$998,633	$665,755	$1,664,388	$9,986,326	$870,653
1	$	$4,000	$659,289	$3,000,000	$185,721	$3,849,011	$481,126	$692,822	$525,000	$5,547,959	$832,194	$554,796	$1,386,990	$8,321,938	$725,544
0.8	$	$3,200	$527,432	$2,400,000	$148,577	$3,079,208	$384,901	$554,258	$420,000	$4,438,367	$665,755	$443,837	$1,109,592	$6,657,551	$580,436
0.6	$	$2,400	$395,574	$1,800,000	$111,433	$2,309,406	$288,676	$415,693	$315,000	$3,328,775	$499,316	$332,878	$832,194	$4,993,163	$435,327
0.4	$	$1,600	$263,716	$1,200,000	$74,288	$1,539,604	$192,451	$277,129	$210,000	$2,219,184	$332,878	$221,918	$554,796	$3,328,775	$290,218

Carollo O&M Cost Estimate:

MGD	Unit	Net Energy Cost FO	Net Energy Cost RO Subsystem	Membrane Replacement Costs FO	Membrane Replacement Costs RO Subsystem	Chemical Costs FO	Chemical Costs RO	Labor Costs FO	Labor Costs RO Subsystem	Total O&M
1.4	$/yr	$17,660		$81,760		$148,928		$60,925		$309,274
1.2	$/yr	$15,137		$70,080		$127,653		$52,222		$265,092
1	$/yr	$12,614	$397,922	$58,400	$355,209	$106,377	$151,984	$43,518	$62,175	$1,188,201
0.8	$/yr	$10,092		$46,720		$85,102		$34,814		$176,728
0.6	$/yr	$7,569	$239,748	$35,040	$209,896	$63,826	$91,191	$26,111	$37,305	$710,686
0.4	$/yr	$5,046	$157,843	$23,360	$145,313	$42,551	$60,794	$17,407	$24,870	$477,183

Values to generate O&M cost curve

MGD	O&M
1	$1,188,201
0.6	$710,686
0.4	$477,183

Note 1: The diagram provides the system setup for a conceptual pilot FO/RO system. The secondary RO system is used to continually concentrate the draw solution at 50g/L while recovering water from the brine stream (from 1°RO).
Note 2: The FO/RO model in this case is based on a plate and frame unit using flat sheet membranes. Hydration Technologies in Oregon currently produces a 4" diameter, 12" long spiral wound FO module but UNR reports poor rejection and poor flow characteristics after testing. Tzahi believes that future FO membranes should have either hollow fiber or tubular configurations.
Note 3: 2°RO recovery for the DS+water is slightly less energy intensive than seawater desalination - achievable recovery 50% to 70% and the final product TDS achieved from secondary RO is 88 mg/L TDS
Updated: Feb 28 2007

Annualized Capital Cost Curve
y= 725124.602 x+ 279.9252516

Total Capital Cost Curve
y= 8317122.059 x+ 3210.720583

O&M Cost Curve
y= 1186280.745 x+ 1169.63086

Revenue from New Water Production:

MGD	Q_{FO} mgd	Potable water produced mgd	Potable water produced AF/yr	Selling price of new water $/AF	Revenue generated $/annum
1.4	0.98	0.686	768	549.00	$421,891
1.2	0.84	0.588	659	549.00	$361,621
1	0.7	0.49	549	549.00	$301,350
0.8	0.56	0.392	439	549.00	$241,080
0.6	0.42	0.294	329	549.00	$180,810
0.4	0.28	0.196	220	549.00	$120,540

Revenue Generated
y= 301350.4331 x+ -5.82077E-11

Equipment Cos FO system (1 MGD at 70% recovery)
y= $3,849,011

Labor
y= $105,693

[redacted] MEMBRANE DISTILLATION

1. Feed concentration has minimal effect on water flux in VEDCMD
2. No membrane companies make membranes specifically designed for MD - currently MD membranes are merely MF membranes that have specific pore size and are hydrophobic
3. Assumes energy to heat the water is FREE

For the purpose of cost estimating,

Membrane		Microdyn (MD 150 CP 2N) - Tube and Shell configuration		
Membrane Area	m^2	10		
Retail Cost	$	5000-7000		
# of capillary membranes		1800	Polypropylene capillary membranes	
Feed Temp	${}^{o}C$	40	60	
Permeate Temp	${}^{o}C$	20	20	
Permeate Flow	L/m^2-hr	2-4	6-8	
approx.	L/day (per membrane element)	750	1600	
	gpm	0.1375896	0.293524	
Feed Flow rate	L/min (per membrane element)	200-250		
	m^3/min	0.2-0.25		
Single-element recovery	%	0.2	0.4	
To obtain 70% recovery from 1 MGD of Brine A,	# of elements required	31000	15000	
Feed Flow rate	m^3/min (overall)	6200	3000	(at approximately 10 psig)
	gpm	1637867	792516	
	mgd	2359	1141	
If 25 membranes were connected in series, lowered feed flow rate would be	m^3/min (overall)	310		approx 10 psig (feed side would have reheating stages in between membranes, and the permeate side would require interstage cooling to maintain $20^{o}C$)
	gpm	81893		
	mgd	118		

<u>Comment from Amy:</u>
From the above numbers, we found FO to be the ONLY viable process at this point in time.

Evaporation Ponds

Evaporation pond regression model from USBR (Mickley & Associates, 2001)

Model valid from 10 to 100 acres
20% contigency included

Factors	Unit	Range	Choose configuration
Liner Thickness	mil	20-120	60
Land cost	$/acre	$8,000	$5,000
Land clearing cost	$/acre	Sparsely Woode	$2,000
- clearing brush		$1,000	
- sparsely wooded area		$2,000	
- medium wooded area		$4,000	
- heavily wooded area		$7,000	
Dike height	ft	4-12	8
Evaporation area	acres	10-100	10
Total Unit Area Capital Cost	$/acre		**$42,258**
Total Area (plus 20% contigency factor)	acre		16.71
Total Capital Cost	$		**$705,940**

Another Evaporation Pond Model (Svensson, 2005)

Amount of brine produced per day	m3/day	170.3435203	31	gpm
Evaporation rate (annual mean)	m/day	0.0136	195	in/yr
Precipitation rate (annual mean)	m/day	0.0025	36	in/yr
Assumed Recovery rate	%	80%		
Evaporation pond area needed	m2	43712	**11**	acres

Model from Graham - modified from Erin's Model *(H:\Common Use\TOOLBOX\Membranes\evaporat*

Notes:
- (a) Mean Annual Precip - Mean Annual Evaporation
- (b) Evap Ratio accounts for reduced evaporation rates with increases in salinity
- (c) Year contingencies reflect design flow accepted
- (d) Square pond assumed
- (e) Saturation assumed for NaCl
- (f) Salt Density assumed to be 135 lb/ft3 (NaCl)
- (g) self propelled scraper 3000' haul
- (h) Source: www.fieldliningsystems.com, (3-4 layers, 1 clay, 2 plastic liners 60 mil, with seapage water collection)
  $2/ ft installed
  clay liner + 2 layers of 60 mil HDPE + Geonet

City	in/yr (a)	
Albuquerque	-44	
Fresno	-55	
Las Vegas	-70	
Los Angeles	-34	
Phoenix	-92	
Reno	-51	
Sacramento	-37	
Salt Lake City	-40	
San Diego	-23	
Ridgecrest	-85 to 90	highest in the USA

Parameter		Value	Units	Erin	Mickley	
Discharge Flow (mgd)	0.045	31	gpm			
TDS		263,600	mg/l			
Net Annual Evaporation Rate		70	in/yr			
Evaporation Ratio (b)		1				
Assumed Water Depth		18	in			
Designed Water Depth		21.6	in			
Min. Freeboard		2	ft			
First Year Contingency (.c)		150%				
Second Year Contingency (c)		130%		56 inches evap rate		
Remaining Years (.c)		120%				
Flow per Year		2,195,704	ft3/yr		74,000	capital costs per acre
Required Area (plus 20% contingency)		**12.96**	**acres**	**10.31249934**	1.2	contingency
Required Liner (PVC or Hypalon) (d)		13.09	acres			
Pond Water Volume		1.02	mega-ft3			
Deposition		98852	lbs/d			
Yearly Deposition		18040	tons/yr			
Time to Saturation (264,000 mg/L) (e)		0.5	years			
Half-Water Depth Consumed (f)		1.90	years			
Volume to be cleared		79,463	C.Y.			
Excavation cost ($)		**$397,316**				
land costs per acre		$4,000	acre			
land costs		$51,846				
multiple barrier p acre(h)		2	$/sq ft			
multiple barrier installed		$1,129,214				
Estimated capital costs		**$1,578,377**		**$928,125**	**$1,150,990**	
capital costs per acre		$121,773.33		$71,605.76	$88,800.00	
payment period	20					
interest rate	6					
annual payment		**$137,610**				

50	**3375**	**1637**	
75	5062	2455	
100	6750	3274	405000

O&M Costs	Unit Cost	Annual costs
sludge disposal ($/ton)	$50	$902,022
seepage monitoring	?	
clean up of contaminated soil	?	
Total		**$902,022**

		Accumulation	
Influent flowrate (AF/yr)	50.48	(AF)	(inches)
Year 1 - amount evaporated (AF/yr)	75.61	-64.45	-59.67
Years to fill pond to assumed water depth	-0.3		

3.9	Trucks per day
9000	Truck volume
35100	Trucked per day (gal)
39.32	AF/yr
176400	
4.05	Acres

Water Quality Data for Brine A, B and C
Brine A: Primary RO Brine from Menifee Desalter
Brine B: Secondary RO Brine (after softening_
Brine C: SPARRO Brine (Treating Brine B)

Table : Summary of Water Quality Data:

Evaluation and Selection of Available Processes for a ZLD System

Eastern Municipal Water District, Menifee Desalter, Sun City, CA

Parameter	Units	Detection Limit [a]	Brine A 1° RO Brine	Brine B 2°RO Brine	Brine C SPARRO Brine
pH	pH units	-	7	7.2	7.2
Bicarbonate	mg/L	3	792	223	316
Total Alkalinity	mg/L as $CaCO_3$	3	652	188	259
Chloride	mg/L	1	2439	9891	10598.8
Fluoride	mg/L	0.1	0	0	0.48
Nitrate – N	mg/L	0.1	20	65	60
Sulfate	mg/L	1	462	2202	3338.4
Boron	mg/L	0.01	0	0	0
Calcium	mg/L	1	994	2200	1550
Hardness [b]	mg/L as $CaCO_3$		3470	6222	11000
Magnesium	mg/L	1	234	614	684.33
Potassium	mg/L	0.1	26	101	70.7
Silica	mg/L as SiO_2	1	165	184	132.33
Dissolved Silica	mg/L as SiO_2	1	165	166	230
Reactive Silica	mg/L as SiO_2	1	182	78	
Sodium	mg/L	10	873	4142	5476.91
Iron	µg/L	5	26	13	2700
Manganese	µg/L	2	8	3	840
Aluminum	µg/L	5	23	23	2450
Arsenic	µg/L	1	8	17	9.5
Barium	µg/L	1	657	496	450
Selenium	µg/L	2	19	70	93
Strontium	µg/L	1	5061	11325	26000
Ammonia as N	mg/L	1	1	1	1
Total Dissolved Soli	mg/L	25	5701	18605	22264.49
Total Organic Carbo	mg/L	0.7	1.4	5	-
Total Phosphate - P	mg/L	0.1	0.4	2	5
Total Suspended So	mg/L	3	18	34	19500
Electrical Conductar	umhos/cm	1	8900	30309	45267

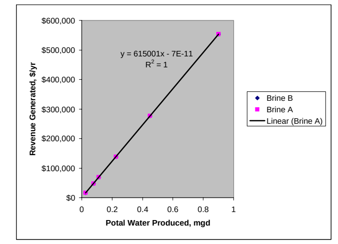
Notes:
Calculations assume values for non-detect results are at the detection limit.
Calculations are based on EMWD lab analysis for the period 06/23/2006 to 12/04/2006
Reported values are average values from multiple data sets, wherever possible
a: Based on reporting detection limit as provided by EMWD laboratory

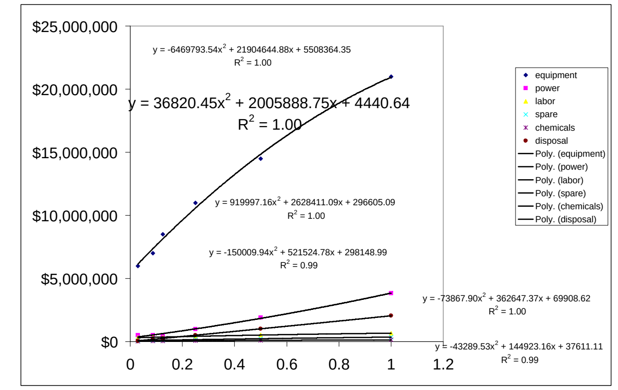
		180	Balanced 120
Bicarbonate (HCO3)	mg/L	750	316
Carbonate (CO3)	mg/L	3	3
Hydroxide (OH)	mg/L	3	3
Total Alkalinity as CaCO3	mg/L	610	259
Chloride	mg/L	17800	10598.8
Fluoride	mg/L	0.6	0.48
Nitrate as N	mg/L	99	60
Sulfate	mg/L	5770	3338.4
Boron	mg/L	0	
Calcium	mg/L	3000	1550
Hardness	mg/L	14000	
Magnesium	mg/L	1600	684.33
Potassium	mg/L	180	70.7
Silica	mg/L	210	132.33
Sodium	mg/L	6600	5476.91
Electrical Conductance	umhos/cm	51200	
Iron	ug/L	21	2700
Manganese	ug/L	3	840
Aluminum	ug/L	50	2450
Arsenic	ug/L	20	9.5
Barium	ug/L	740	450
Selenium	ug/L	130	93
Strontium	ug/L	26000	26000
Ammonia as N	mg/L	1	1
pH	pH units	7.9	7.2
Silica: Dissolved	mg/L	-	230
Total Dissolved Solids	mg/L	37800	22264
Total Organic Carbon	mg/L	16	-
Total Phosphate as P	mg/L	9	5
Total Suspended Solids		-	19500

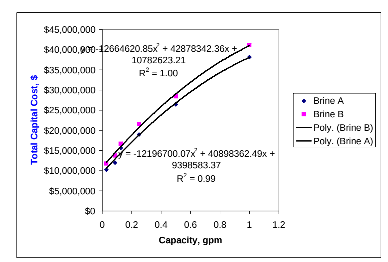
BRINE CONCENTRATOR AND CRYSTALLIZER COST ESTIMATE

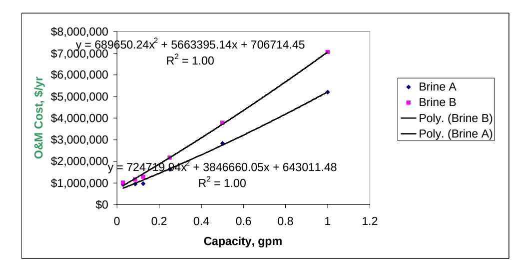
NOTE: GRAY AREAS NOT USED FOR SPREADSHEET

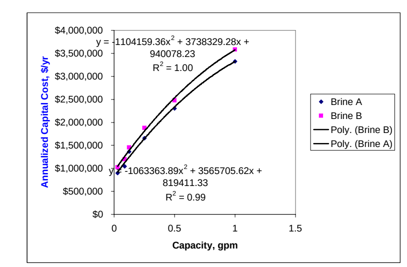
On stream	hr/yr	8,400				
(MGD)	1	0.50	0.250	0.125	0.086	0.029
Item/Capacity (gpm)	**694**	**371**	**174**	**86**	**60**	**20**
*For a brine stream of TDS **6300** mg/L,*						
Brine Concentrator (BC) - Equipment only	$9,000,000	$5,500,000	$4,000,000	$3,000,000	$2,437,500	$1,875,000
Brine Crystallizer (BX) - Equipment only	$1,500,000	$1,000,000	$700,000	$500,000	$406,250	$312,500
Est. Installation costs	$9,000,000	$7,000,000	$5,000,000	$4,500,000	$3,270,313	$3,062,500
Total turnkey costs	$19,500,000	$13,500,000	$9,700,000	$8,000,000	$6,114,063	$5,250,000
Interconnecting Pipework (12.50%)	$2,437,500	$1,687,500	$1,212,500	$1,000,000	$764,258	$656,250
Electrical and Instrumentation (18%)	$3,510,000	$2,430,000	$1,746,000	$1,440,000	$1,100,531	$945,000
Sub-total 2 Construction Cost Estimate (sans building cost)	$25,447,500	$17,617,500	$12,658,500	$10,440,000	$7,978,852	$6,851,250
Engineering (15%)	$3,817,125	$2,642,625	$1,898,775	$1,566,000	$1,196,828	$1,027,688
Legal and Admin (10%)	$2,544,750	$1,761,750	$1,265,850	$1,044,000	$797,885	$685,125
Contingency (25%)	$6,361,875	$4,404,375	$3,164,625	$2,610,000	$1,994,713	$1,712,813
Estimated Project Cost	**$38,171,250**	**$26,426,250**	**$18,987,750**	**$15,660,000**	**$11,968,277**	**$10,276,875**
Annual Cost of capital:						
average annual interest rate (%)	6%	6%	6%	6%	6%	6%
Loan period (yrs)	20	20	20	20	20	20
Annual Payment ($/yr)	**$3,327,944**	**$2,303,961**	**$1,655,439**	**$1,365,310**	**$1,043,449**	**$895,985**
Power Consumption (kw)	3,600	1,800	900	450	335	130
USD/kw-hr	*0.12*	*0.12*	*0.12*	*0.12*	*0.12*	*0.12*
USD/hr	*$432*	*$216*	*$108*	*$54*	*$54*	*$54*
USD/yr	*$3,628,800*	*$1,814,400*	*$907,200*	*$453,600*	*$453,600*	*$453,600*
Solids waste rate @ 10% moisture (tons/day)	29.0	15.0	7.3	3.6	2.6	1.0
assume disposal cost of $50/ton, disposal cost /yr	$507,500	$262,500	$127,750	$63,000	$45,975	$16,695
Operating Labor Req'ts (man-hr/yr)	13,000	10,000	8,000	6,000	6,001	6,002
assume man hour of $47.75/hr, labor cost /yr	$620,750	$477,500	$382,000	$286,500	$286,548	$286,596
Spares & Maintenance, allowance at 3% of:						
Equipment Costs/yr	$315,000	$195,000	$141,000	$105,000	$105,000	$105,000
Chemical & Cleaning costs/yr	$125,000	$85,000	$67,000	$55,000	$55,000	$55,000
Total O&M Cost /yr	**$5,197,050**	**$2,834,400**	**$1,624,950**	**$963,100**	**$946,123**	**$916,890**
Revenue from water recovered						
Potable water produced (mgd)	0.9	0.45	0.225	0.1125	0.077809798	0.025936599
Potable water produced (AF/yr)	1,008	504	252	126	87	29
Selling price of new water ($/AF)	549	549	549	549	549	549
Revenue generated ($/annum)	$553,501	$276,750	$138,375	$69,188	$47,853	$15,951
*For a brine stream of TDS **28000** mg/L,*					$3,250,000	$2,500,000
Brine Concentrator (BC) - Equipment only	$9,000,000	$5,500,000	$4,000,000	$3,000,000	$2,437,500	$1,875,000
Brine Crystallizer (BX) - Equipment only	$3,000,000	$2,000,000	$1,500,000	$1,000,000	$812,500	$625,000
Est. Installation costs	$9,000,000	$7,000,000	$5,500,000	$4,500,000	$3,750,000	$3,500,000
Total turnkey costs	$21,000,000	$14,500,000	$11,000,000	$8,500,000	$7,000,000	$6,000,000
Interconnecting Pipework (12.50%)	$2,625,000	$1,812,500	$1,375,000	$1,062,500	$875,000	$750,000
Electrical and Instrumentation (18%)	$3,780,000	$2,610,000	$1,980,000	$1,530,000	$1,260,000	$1,080,000
Sub-total 2 Construction Cost Estimate (sans building cost)	$27,405,000	$18,922,500	$14,355,000	$11,092,500	$9,135,000	$7,830,000
Engineering (15%)	$4,110,750	$2,838,375	$2,153,250	$1,663,875	$1,370,250	$1,174,500
Legal and Admin (10%)	$2,740,500	$1,892,250	$1,435,500	$1,109,250	$913,500	$783,000
Contingency (25%)	$6,851,250	$4,730,625	$3,588,750	$2,773,125	$2,283,750	$1,957,500
Estimated Project Cost	**$41,107,500**	**$28,383,750**	**$21,532,500**	**$16,638,750**	**$13,702,500**	**$11,745,000**
Annual Cost of capital:						
average annual interest rate (%)	6%	6%	6%	6%	6%	6%
Loan period (yrs)	20	20	20	20	20	20
Annual Payment ($/yr)	**$3,583,939**	**$2,474,625**	**$1,877,301**	**$1,450,642**	**$1,194,646**	**$1,023,983**
Power Consumption (kw)	3,800	1,900	975	500	375	150
USD/kw-hr	*0.12*	*0.12*	*0.12*	*0.12*	*0.12*	*0.12*
USD/hr	*$456*	*$228*	*$117*	*$60*	*$60*	*$60*
USD/yr	*$3,830,400*	*$1,915,200*	*$982,800*	*$504,000*	*$504,000*	*$504,000*
Solids waste rate @ 10% moisture (tons/day)	117	58	29	15	10	3
assume disposal cost of $50/ton, disposal cost /yr	*2,047,500*	*1,015,000*	*507,500*	*262,500*	*176,715*	*58,905*
Operating Labor Req'ts (man-hr/yr)	14,000	11,000	9,000	7,000	7,001	7,002
assume man hour of $47.75/hr, labor cost /yr	*$668,500*	*$525,250*	*$429,750*	*$334,250*	*$334,298*	*$334,346*
Spares & Maintenance, allowance at 3% of:						
Equipment Costs/yr	$360,000	$225,000	$165,000	$120,000	$97,500	$75,000
Chemical & Cleaning costs/yr	$140,000	$95,000	$75,000	$60,000	$48,750	$37,500
Total O&M Cost /yr	**$7,046,400**	**$3,775,450**	**$2,160,050**	**$1,280,750**	**$1,161,262**	**$1,009,750**
Revenue from water recovered						
Potable water produced (mgd)	0.9	0.45	0.225	0.1125	0.077809798	0.025936599
Potable water produced (AF/yr)	1,008	504	252	126	87	29
Selling price of new water ($/AF)	$549	$549	$549	$549	$549	$549
Revenue generated ($/annum)	**$553,501**	**$276,750**	**$138,375**	**$69,188**	**$47,853**	**$15,951**







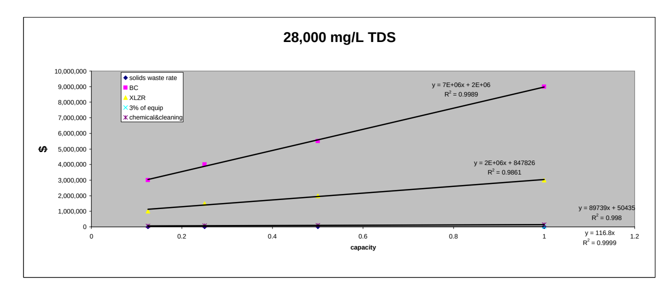




$y=Ax^2+bx+c$

	A	b	c
Total Capital Cost Curve For A	-12196700.07	40898362.49	9398583.37
Total Capital Cost Curve For B	-12664620.85	42878342.36	10782623.21
Annualized Cost Curve for A	-1063363.89	3565705.62	819411.33
Annualized Cost Curve for B	-1104159.36	3738329.28	940078.23
O&M Cost Curve for A	724719.94	384660.05	643011.48
O&M Cost Curve for B	689650.24	5663395.14	706714.45

			Equipment Cost
BC			
y=	6782608.7 x+	2195652.174	
60gpm	$2,782,045		$3,819,321
20gpm	$2,391,116		$3,302,092
XLZR			
y=	2191304.35 x+	847826.087	
60gpm	$1,037,276		
20gpm	$910,976		
3% of equipment			
y=	269217.391 x+	91304.34783	
60gpm	$114,580		
20gpm	$99,063		
3% of chemical and cleaning			
y=	89739.1304 x+	30434.78261	
60gpm	$38,193		
20gpm	$33,021		
solids waste rate			
y=	116.8 x+	0	
60gpm	10		
20gpm	3		



Notes from Dave Ciszewski:

1. Turnkey installation estimates given are for mechanical and electrical on a customer supplied foundation. No buildings are allowed for. It is assumed that the brine concentrator (BC) and brine crystallizer (XZLR) are outdoors and the PLC computer interface and MCC are installed in the same electrical/contro building as the RO plant.

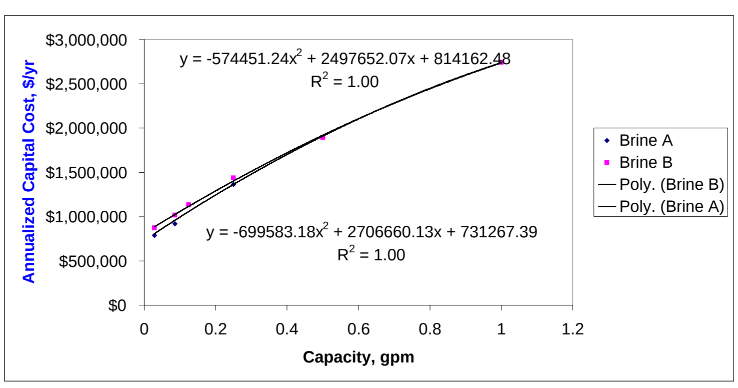
2. Flow out of the BC to the XZLR is approximately 15% at 28000 mg/L TDS feed and 4% at 6300 mg/L feed

3. The smallest BC provided is 20 gpm. Budget pricing for a 20 gpm BC+ XLZR is $2,500,000 for the equipment and $6,000,000 turnkey with a power consumption of 150 kW. Budget pricing for a 60 gpm BC+XLZR is $3,250,000 for the equipment and $7,000,000 turnkey with a power consumption of 375 kW. These are based on 28000 mg/L TDS feed. Subtract 5-10% at 6300 mg/L feed. Add 10% to individual prices if the BC and XLZR are split up.

BRINE CONCENTRATOR COST ESTIMATE

On stream	hr/yr		8,400				
	(MGD)	1	0.50	0.250	0.125	0.086	0.029
Item/Capacity (gpm)		**694**	**371**	**174**	**86**	**60**	**20**
For a brine stream of TDS ***6300*** *mg/L,*							
Brine Concentrator (BC) - Equipment only		$9,900,000	$6,050,000	$4,400,000	$3,300,000	$2,949,375	$2,268,750
Est. Installation costs		$9,000,000	$7,000,000	$5,000,000	$4,500,000	$2,440,608	$2,359,500
Total turnkey costs		$16,065,000	$11,092,500	$7,990,000	$6,630,000	$5,389,983	$4,628,250
Interconnecting Pipework (12.50%)		$2,008,125	$1,386,563	$998,750	$828,750	$673,748	$578,531
Electrical and Instrumentation (18%)		$2,891,700	$1,996,650	$1,438,200	$1,193,400	$970,197	$833,085
Sub-total 2 Construction Cost Estimate (sans building cost)		$20,964,825	$14,475,713	$10,426,950	$8,652,150	$7,033,928	$6,039,866
Engineering (15%)		$3,144,724	$2,171,357	$1,564,043	$1,297,823	$1,055,089	$905,980
Legal and Admin (10%)		$2,096,483	$1,447,571	$1,042,695	$865,215	$703,393	$603,987
Contingency (25%)		$5,241,206	$3,618,928	$2,606,738	$2,163,038	$1,758,482	$1,509,967
Estimated Project Cost		**$31,447,238**	**$21,713,569**	**$15,640,425**	**$12,978,225**	**$10,550,891**	**$9,059,799**
Annual Cost of capital:							
average annual interest rate (%)		6%	6%	6%	6%	6%	6%
Loan period (yrs)		20	20	20	20	20	20
Annual Payment ($/yr)		**$2,741,713**	**$1,893,088**	**$1,363,604**	**$1,131,501**	**$919,875**	**$789,875**
Power Consumption (kw)		3,600	1,800	900	450	335	130
USD/kw-hr		*0.12*	*0.12*	*0.12*	*0.12*	*0.12*	*0.12*
USD/hr		*$432*	*$216*	*$108*	*$54*	*$54*	*$54*
USD/yr		*$3,628,800*	*$1,814,400*	*$907,200*	*$453,600*	*$453,600*	*$453,600*
Flow out from BC, 4% of feed at 6300 mg/L (mgd)		0.040	0.020	0.010	0.005	0.003	0.001
Operating Labor Req'ts (man-hr/yr)		13,000	10,000	8,000	6,000	6,001	6,002
assume man hour of $47.75/hr, labor cost /yr		$620,750	$477,500	$382,000	$286,500	$286,548	$286,596
Spares & Maintenance, allowance at 3% of:							
Equipment Costs/yr		$315,000	$195,000	$141,000	$105,000	$105,000	$105,000
Chemical & Cleaning costs/yr		$125,000	$85,000	$67,000	$55,000	$55,000	$55,000
Total O&M Cost /yr		**$4,689,550**	**$2,571,900**	**$1,497,200**	**$900,100**	**$900,148**	**$900,196**
For a brine stream of TDS ***28000*** *mg/L,*						$3,250,000	$2,500,000
Brine Concentrator (BC) - Equipment only		$9,900,000	$6,050,000	$4,400,000	$3,300,000	$2,681,250	$2,062,500
Est. Installation costs		$9,000,000	$7,000,000	$5,500,000	$4,500,000	$2,700,000	$2,600,000
Total turnkey costs		$16,065,000	$11,092,500	$8,415,000	$6,630,000	$5,950,000	$5,100,000
Interconnecting Pipework (12.50%)		$2,008,125	$1,386,563	$1,051,875	$828,750	$743,750	$637,500
Electrical and Instrumentation (18%)		$2,891,700	$1,996,650	$1,514,700	$1,193,400	$1,071,000	$918,000
Sub-total 2 Construction Cost Estimate (sans building cost)		$20,964,825	$14,475,713	$10,981,575	$8,652,150	$7,764,750	$6,655,500
Engineering (15%)		$3,144,724	$2,171,357	$1,647,236	$1,297,823	$1,164,713	$998,325
Legal and Admin (10%)		$2,096,483	$1,447,571	$1,098,158	$865,215	$776,475	$665,550
Contingency (25%)		$5,241,206	$3,618,928	$2,745,394	$2,163,038	$1,941,188	$1,663,875
Estimated Project Cost		**$31,447,238**	**$21,713,569**	**$16,472,363**	**$12,978,225**	**$11,647,125**	**$9,983,250**
Annual Cost of capital:							
average annual interest rate (%)		6%	6%	6%	6%	6%	6%
Loan period (yrs)		20	20	20	20	20	20
Annual Payment ($/yr)		**$2,741,713**	**$1,893,088**	**$1,436,136**	**$1,131,501**	**$1,015,449**	**$870,385**
Power Consumption (kw)		3,230	1,615	829	425	319	128
USD/kw-hr		*0.12*	*0.12*	*0.12*	*0.12*	*0.12*	*0.12*
USD/hr		*$456*	*$228*	*$117*	*$60*	*$60*	*$60*
USD/yr		*$3,830,400*	*$1,915,200*	*$982,800*	*$504,000*	*$504,000*	*$504,000*
Flow out from BC, 15% of feed for 28 000 mg/L (mgd)		0.150	0.075	0.038	0.019	0.013	0.004
Potable water produced (mgd)		0.850	0.425	0.213	0.106	0.073	0.024
Potable water produced (AF/yr)		952.187	476.094	238.047	119.023	82.322	27.441
Selling price of new water ($/AF)		$549	$549	$549	$549	$549	$549
Revenue generated ($/annum)		**$522,751**	**$261,375**	**$130,688**	**$65,344**	**$45,195**	**$15,065**
Operating Labor Req'ts (man-hr/yr)		14,000	11,000	9,000	7,000	7,001	7,002
assume man hour of $47.75/hr, labor cost /yr		*$668,500*	*$525,250*	*$429,750*	*$334,250*	*$334,298*	*$334,346*
Spares & Maintenance, allowance at 3% of:							
Equipment Costs/yr		$252,450	$154,275	$112,200	$84,150	$68,372	$52,594
Chemical & Cleaning costs/yr		$119,000	$80,750	$63,750	$51,000	$29,058	$22,352
Total O&M Cost /yr		**$4,870,350**	**$2,675,475**	**$1,588,500**	**$973,400**	**$935,728**	**$913,292**

Notes from Dave Ciszewski:
1. Turnkey installation estimates given are for mechanical and electrical on a customer supplied foundation. No buildings are allowed for. It is assumed that the brine concentrator (BC) and brine crystallizer (XZLR) are outdoors and the PLC computer inter
2. Flow out of the BC to the XZLR is approximately 15% at 28000 mg/L TDS feed and 4% at 6300 mg/L feed
3. The smallest BC provided is 20 gpm. Budget pricing for a 20 gpm BC+ XLZR is $2,500,000 for the equipment and $6,000,000 turnkey with a power consumption of 150 kW. Budget pricing for a 60 gpm BC+XLZR is $3,250,000 for the equipment and $7,000,000 turnk
4. The BC portion makes up approximately 85% of the turnkey, power consumption, O&M, spares, and chemical/cleaning costs of the BC+BX systems



$y=Ax^2+bx+c$

	A	b	c
Total Capital Cost Curve For A	-8024163.94	31045178.51	8387579.33
Total Capital Cost Curve For B	-6588910.43	28647872.46	9338379.54
Annualized Cost Curve for A	-699583.18	2706660.13	731267.39
Annualized Cost Curve for B	-574451.24	2497652.07	815162.48
O&M Cost Curve for A	749053.46	3313456.10	643891.82
O&M Cost Curve for B	668607.61	3548965.05	666630.82

Equipment Cost

MGD	OM Brine A	Brine B	Annual Capital Brine A	Brine B
1.00	$4,689,550	$4,870,350	$7,431,263	$7,612,063
0.50	$2,571,900	$2,675,475	$4,464,988	$4,568,563
0.25	$1,497,200	$1,588,500	$2,860,804	$3,024,636
0.13	$900,100	$973,400	$2,031,601	$2,104,901