Getting the Most From Your Brackish Groundwater: Alternatives for Treating RO Brine to Increase Recovery and Reduce the Volume for Disposal

Graham J.G. Juby, Partner

Wen-Yi Shih, Engineer

G. Adam Zacheis, Associate

Wendie James, Engineer

Carollo Engineers, P.C., Fountain Valley, California

Michael D. Nusser, Water Resources Planning Analyst II

Behrooz Mortazavi, Assistant General Manager, Resource Development

Parameshwaran Ravishanker, Deputy General Manager

Eastern Municipal Water District, Perris, California

ABSTRACT

EMWD recently completed two brine treatment studies, one for the California Department of Water Resources (DWR) and the other for the United States Department of Interior, Bureau of Reclamation (USBR). These studies investigated the cost of different alternative process treatment trains to recover additional usable water from primary reverse osmosis (RO) brine, and either reduce the volume of brine significantly, or go to complete zero-liquid discharge (ZLD). The studies included pilot scale, bench scale and desktop evaluations of both established and emerging technologies for brine treatment. Chemical softening of primary RO brine followed by either a second RO process or electrodialysis reversal (EDR) is an example of so-called established technologies; emerging technologies investigated included membrane distillation (MD), forward osmosis (FO) and seeded reverse osmosis. A 50-gpm chemical softening process was operated for 6-months and provided feed water to RO and EDR pilot units. MD and FO were tested at bench scale, and seeded RO was operated in a batch pilot-plant. Capital and O&M cost estimates were established for several treatment configurations that also included brine concentration, crystallization and pond evaporation. These were then compared. Overall, total annual costs for 14 different treatment train combinations were generally within about 10-percent of each other. Treatment for removal and recovery of salable salt by-products was, however, more costly; probably because of the capital costs associated with selective precipitation and washing and processing of the salts. Treatment trains that included a third concentration step (using the seeded RO process) ahead of thermal mechanical evaporation appeared to have slightly lower overall costs. This is attributed to the reduced size of the thermal evaporation process. Additional work on MD, FO and the seeded RO processes is needed.

INTRODUCTION

Each year more groundwater reverse osmosis (RO) plants come online. This is in response to the growing need for more potable water and the demands placed on the existing fresh water resources. Agencies close to the ocean are able to make use of existing ocean outfalls or brine lines to dispose off RO brine from their desalters, but the cost for such disposal methods is increasing. Agencies further inland, however, face significantly greater challenges when it comes to disposal of RO brine. A recent study (Juby et al., 2006) showed that the additional cost for desalination of brackish groundwater at an inland location could be around $600/AF. This cost is almost entirely due to the costs associated with managing and disposal of the RO brine. Eastern Municipal Water District (EMWD) operates two groundwater desalters and currently discharges its brine to the ocean via a wastewater treatment facility. Brine disposal via this system, although convenient, is not cheap and costs are expected to rise in the future. EMWD has been investigating ways to reduce the volume of brine for discharge and increase the recovery of potable water. This would keep more water in the District's basin, produce more potable water and potentially lower the long-term costs of brine disposal.

EMWD covers an area of 555-square-miles, serving a population of over 650,000 people. Over time EMWD has come to rely on groundwater wells to provide between 20 and 30 percent of its potable water supply. The District owns and operates the Menifee desalters, which treat a challenging groundwater with a total dissolved solids (TDS) concentration of around 2,500 mg/L and high concentrations of silica (60 mg/L), calcium (331 mg/L), sulfate (430 mg/L), iron (0.3 mg/L) and barium (0.13 mg/L). Typically, acid or scale inhibitors are added to RO feed water to prevent membrane scaling, allowing a somewhat higher recovery than would be otherwise possible. Even with such chemical addition, the Menifee Desalters currently achieve only about 70-percent recovery due primarily to the high concentration of silica. This translates to a high volume of concentrate, 30 percent of the feed, which is currently not usable and has to be discharged at relatively high costs.

EMWD's primary RO desalters produces brine of about 6,000 mg/L TDS which presently discharges to the Santa Ana Regional Interceptor (SARI) pipeline that transports it about 50 miles to the Pacific Ocean. Brine disposal via the SARI line is expensive and is likely to become more expensive in the future. Also, the capacity in the SARI line is limited and purchasing additional capacity is costly. As a result, EMWD has decided to investigate brine treatment to recover additional water and reduce the volume of brine for disposal, as well as investigate the alternative to convert the entire system to zero liquid discharge (ZLD).

The major benefits to EMWD of treating brine include:

- Increasing potable water production
- Retaining more water in the ground water basin
- Reducing the volume and cost of brine disposal to the ocean

In order to assist with the cost of such evaluations, EMWD applied for and received funding assistance from two sources; the California Department of Water Resources (DWR) and the United States Bureau of Reclamation (USBR). This paper focuses on the project funded by the USBR, and performance results from the DWR study were used as part of the study.

The objectives of the project were to research a wide range of existing and emerging technologies that could be used as part of a Zero-Liquid Discharge (ZLD) brine management system, and establish preliminary cost estimates for various ZLD process combinations.

PROCESS SELECTION

The following processes were evaluated as part of the brine treatment studies:

- Conventional lime/soda softening followed by secondary desalting treatment using RO or EDR (part of the DWR funded project).
- Seeded RO (or SPARRO process) for treatment of second stage RO brine.
- Membrane Distillation (MD) and/or Forward Osmosis (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 in 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 to landfill. Additional water will be recovered as a result of this treatment.
- Evaporation ponds as a final disposal point in a desalination treatment train once the brine stream has been concentrated to a manageable volume.

OVERALL TECHNICAL APPROACH

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

Experimental Methods

Three brine streams were evaluated. Brine A, produced from EMWD’s primary RO process (TDS of approximately 6,000 mg/L), was softened and fed to a secondary RO pilot and/or EDR pilot - see Figure 1 which is a simplified schematic illustrating the sources of the three brine streams. The brine stream produced from the secondary RO pilot was designated as “Brine B” and had a TDS concentration of approximately 18,000 mg/L. The brine from the EDR process was similar in composition to “Brine B” and 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.

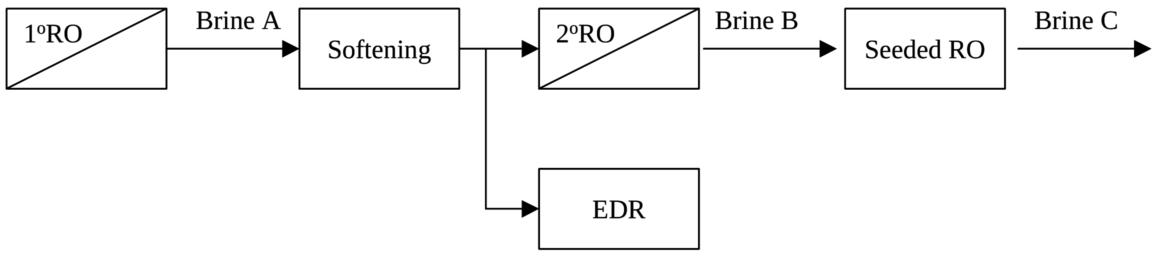


Figure 1. Schematic showing brine designation from pilot tests for subsequent bench-scale and desktop analyses

Samples of "Brine A" and "Brine C" were used as feed water for FO and MD bench-scale testing, and were sent to the University of Nevada, Reno (UNR) for this work to be carried out . Water analysis data for both "Brine A" and "Brine B" were sent to Geo-Processors USA as feed water quality input to their SAL-PROC™ technology model to determine process stream configurations to optimize salt by-product production and water recovery.

Concentrate discharges from the secondary and tertiary treatment systems (RO, EDR, SPARRO, FO and MD) can be processed downstream for further water recovery to achieve zero-liquid discharge. Downstream processes included in this study were a combination of a thermal evaporation process (brine concentrator) and a crystallizer, or a brine concentrator and an evaporation pond. Brine composition and volume information were sent to brine concentrator and crystallizer manufacturers for accurate price quotes specific to the brine stream. For comparison, direct discharge to the SARI line was also considered.

A cost analysis model was developed to analyze and compare the costs for different treatment combinations. Within the model, each brine treatment process is treated as a block with associated capital and operating cost information. Different process blocks were combined to form alternate treatment trains in the model and the resulting costs were compared. A total of 14 process trains were compared in the study; plus two "base-case" alternatives.

Pilot Study

Chemical Softening

Brine A was pumped at 20-50 gpm (76-190 L/min) to a USFilter ZIMPRO unit for preliminary softening and clarification via rapid mixing and flocculation, followed by a plate settling basin mounted above a sludge hopper. Suspended solids were subsequently removed by conventional dual-media filtration. The brine stream was dosed with 600-650 mg/L of 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 floc agglomeration of calcium carbonate and magnesium hydroxide.

Reverse Osmosis (RO)

The pH of the softened effluent was adjusted to 7.0 by dosing 25-percent sulfuric acid before secondary RO and EDR treatment. The RO skid-mounted pilot from Harn RO Systems consisted of two stages of RO membranes arranged in a 2:1 array of three and four-element pressure vessels in series. This particular array was selected to mimic the configuration of the primary desalter. Dow Filmtec™ XLE-4040 spiral wound 40-inch (1.02 m) by 4-inch (0.10 m) membrane elements were housed in the pressure vessels. Vitec 3000 from Avista Technologies

(San Marcos, CA) was the chosen antiscalant for this study, dosed at 4.0 mg/L. Softened brine was fed to the RO pilot unit at a flow rate of around 20 gpm (76 L/min).

Electrodialysis-Reversal (EDR)

The EDR pilot unit (Aquamite V, GE) utilizes a single EDR membrane stack with two electrical stages and four hydraulic stages. The cathode/anode operation alternates every 15 to 30 minutes by reversing the polarity, or direction of current flow. This aided in preserving the integrity of the membranes by preventing scale build-up. During charge reversal, approximately 30 to 45 seconds, out-of-spec product water is diverted as a waste stream. Softened Brine A was fed at 12 (45) to 15 (57) gpm (L/min) to the unit and hydrochloric acid was continuously fed to the circulating brine stream and the anode flush water to prevent calcium carbonate scaling. The feed was passed through each electrical stage twice to provide greater residence time for ion transfer. Brine formed within the concentrate cell pairs was recirculated back to the concentrate system in a loop until the salts become supersaturated, at which point the salts are removed as a reject stream.

SPARRO (Seeded RO)

Tubular NF membranes from Koch Membrane Systems (San Diego, CA) were utilized in this pilot setup due to the unavailability of tubular RO membranes. The membranes were the SelRO® MPT-34 pH-stable composite membranes. Gypsum slurry (10 to 18 g/L) was circulated within the membranes with the brine stream to promote preferential calcium sulfate crystal precipitation. The NF membrane module was 150-inches (3.8 m) in length with an active membrane area of 28 ft^2 (2.6 m^2). The module contains eighteen tubes connected in series, and has a nominal permeate production rate of about 1,000 gpd (2.6 L/min).

A batch-feed tank was filled with 120 gallons (454 L) of Brine B and commercial gypsum powder was then added to the brine to produce the desired concentration of gypsum slurry (10 to 18 g/L). The prepared gypsum slurry was then pumped through the Koch tubular membranes using a high-pressure positive displacement pump at between 200 and 600 psi (13.6 to 40.8 atm). Permeate produced by the membrane module was removed from the system and the concentrate stream was recycled to the feed tank. Hence the solution in the feed tank became more concentrated with time which allowed the system to simulate operation at different water recovery levels. Permeate produced from the membrane vessel was collected periodically for laboratory analysis.

Bench-Scale Study

Vacuum Enhanced Direct Contact Membrane Distillation (VEDCMD)

Membrane distillation experiments were carried out by the University of Nevada, Reno using flat sheet polypropylene membranes (GE Osmonics, MN) and polytetrafluoroethylene membranes (GE Osmonics, MN) using a bench-scale membrane test unit. Warm feed solutions (Brine A and Brine B) at either 40°C or 60°C were circulated on the feed side of the membrane at 0.4 gpm (1.5 L/min). De-ionized water at 20°C was circulated counter currently on the support side of the membrane, also at 0.4 gpm (1.5 L/min). A vacuum was applied to the de-ionized side of the membrane. The driving force for transport is the vapor pressure difference across the membrane. Separation occurs when vapor passes through the membrane pores by a convective or diffusive mechanism. As water evaporates through the membrane, the permeate volume continually increases and this change in volume and conductivity was recorded to calculate water flux, recovery and rejection.

Forward Osmosis (FO)

Forward Osmosis (FO) is a process that uses natural osmosis to separate water from salts. Like RO, FO uses a semi-permeable membrane to effect separation of water from dissolved solutes, the difference is that pressure is not used as the driving force. Flat sheet cellulose triacetate membranes were used for this study. The driving force for FO is an osmotic pressure gradient, such that a "draw" solution of high concentration (relative to that of the feed solution) is used to induce a net flow of water through the membrane into the draw solution, thus effectively separating the feed water from its solutes. A bench-scale FO test unit was set up at the University of Nevada, Reno. This unit was coupled with a pilot-scale RO system used to continually concentrate the draw solution. Brine A was used as the feed solution and was circulated at 0.4 gpm (1.5 L/min) on the feed side of the FO membrane. A sodium chloride draw solution (at about 50,000 mg/L concentration) was circulated counter currently at the same flow rate on the support side. As water from the feed stream diffused into the draw solution, the change in volume was monitored for flux and recovery calculations. This information was obtained during the testing and then used to estimate the size and cost of the treatment system.

Simulation

Residual Recovery (SAL-PROC™)

Geo-Processors USA, Inc., was engaged to perform a desktop pre-feasibility study of the SAL-PROC™ process for recovery of useful byproducts such as magnesium salts and calcium carbonate. A nominal 1-mgd (2628 L/min) flow rate was used for analysis for Brine A and Brine C treatment. Desktop modeling software developed by Geo-Processors was used to identify technically feasible ZLD process streams. Each treatment stream was comprised of subsystems including one or more selective salt recovery steps that are linked with RO desalination, thermo-mechanical brine concentration and crystallization steps. Three salt by-products were identified; precipitated calcium carbonate (PCC), carbonated magnesium hydroxide and mixed salt (mainly sodium sulfate and sodium chloride). These salt products were considered as one of the parameters for conceptual design and sizing of the SAL-PROC™ plants, and estimates were made of the potential selling price for the salts.

Brine Concentrator (BC) and Crystallizer (XLZR)

HPD-Veolia Water Solutions were given water quality data for Brines B and C, and responded with budget estimates for the treatment of both brines using a combination of a brine concentrator with and without a crystallizer for treatment capacities ranging from 0.125 to 1 mgd (328 to 2628 L/min).

Evaporation Ponds

An desktop model was developed based on a report from the United States Bureau of Reclamation (Mickley & Associates, 2001) to estimate evaporation pond costs using the local net evaporation rate of 70 inches/year (1.78 m/year).

PROCESS PERFORMANCE

Pilot Study Results

Table 1 Average Water Quality Data for Three Brine Types

Parameter	Units	Detection Limit(1)	Brine A(2) 1° RO Brine	Brine B(3) 2° RO Brine	Brine C(4) SPARRO Brine
pH	pH units	-	7	7.2	7.2
Total Alkalinity	mg/L as $CaCO_3$	3	652	188	259
Chloride	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(5)	mg/L as $CaCO_3$		3,470	6,222	11,000
Magnesium	mg/L	1	234	614	684
Dissolved Silica	mg/L as 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	µmhos/cm	1	8,900	30,309	45,267

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.
(1) Based on reporting detection limit as provided by EMWD laboratory.
(2) Data averaged over 6-month period
(3) Brine B was produced from treatment of Brine A after softening - data averaged over 6-month period.
(4) Average of three samples.
(5) Reported hardness of 23,000 mg/L as $CaCO_3$ on 11/01/2006 omitted as aberration.

Results from the pilot, bench and modeling work were combined to generate operating performance criteria for each treatment process. These criteria were then used in the process treatment trains to determine the quality and flow rate of the brine after each step in the process. This allowed the size of each treatment step to be estimated and therefore estimates of the capital and operating cost of each process step could be made. Table 2 summarizes the operating performance criteria for each treatment process. More details can be obtained in the USBR Report on the project (U.S. Department of the Interior, 2008).

Table 2	**Operating Performance Criteria for each Treatment Process**
PILOT STUDY	
Chemical Softening	• Lime dose of 555 mg/L • Soda ash dose of 150 mg/L • Acid dose of 20 mg/L • 40-percent dewatered sludge with a sludge disposal cost of $30/ton • Product water TDS: 4,500 mg/L
Reverse Osmosis (RO)	• 70-percent recovery • Operating pressure 215 psi • Chemical cost is $0.07/1,000 gallons feed volume • Product water TDS: 282 mg/L
Electrodialysis-Reversal (EDR)	• 75-percent recovery • Chemical cost is $0.025/1,000 gallons product volume • Product water TDS: 415 mg/L
Seeded RO	• 60-percent recovery • Chemical cost is $0.07/1,000 gallons feed volume • Feed pressure 500 psi • Product water TDS: 10,423 mg/L
BENCH SCALE TESTS	
Membrane Distillation (MD)	• 70-percent recovery • Irreversible membrane fouling occurred during the testing. Accordingly, estimating full-scale performance and costs seemed unrealistic. Therefore MD was eliminated from further consideration.
Forward Osmosis (FO)	• 70-percent recovery • Energy consumption of 9 kWh/1,000 gal • Modeled product water 88 mg/L TDS
DESKTOP SIMULATION	
SAL-PROC™	• Salinity variations between Brine A and C required two different treatment trains • Brine streams undergo treatment in a series of $Mg(OH)_2$ and $CaCO_3$ recovery, with additional potable water recovery through RO, brine concentration and thermal crystallization
Brine Concentrator	• Flow of out BC = 15 percent of feed Brine C (28,000 mg/L)

Table 2	Operating Performance Criteria for each Treatment Process
(BC) and Crystallizer (XLZR)	• BC portion of the BC-XLZR system makes up ~85 percent of the turnkey costs, power consumption, operation and maintenance costs, spare equipment and chemical cleaning costs. • Solid waste:10-percent moisture • 3,800 kW/mgd
Evaporation Ponds	• 263,600 mg/L TDS assumed for incoming brine salinity based on projections from BC-XLZR treatment of Brine B • Assumed water depth for pond was 18 inches, with a minimum freeboard of 2 ft. • Excavation costs of $5/cubic yard, land costs of $4,000/acre, and installed liner cost of $2/sq ft were assumed • 20-percent required land contingency built into the model

COST ANALYSIS ASSUMPTIONS

Table 3 presents a summary of the assumptions used in the cost model for determining the annual operating cost estimates for each treatment process.

Table 3 Cost Analysis Assumptions

Parameter	Unit
Annual Interest Rate	6 percent
Loan Period	20 yrs
Disposal Cost to Landfill	$50/ton
Labor Cost	$47.75/hr
Electricity	$0.12/kWh
Building	$150/sq ft

PROCESS COMBINATION ALTERNATIVES

Fourteen process combination alternatives and two base cases were developed, as described below. The overall concept was to develop unit capital and operating costs for each treatment technology, as well as develop performance criteria for each. Then, assuming a starting point of 1-mgd of primary RO brine (Brine A) (which is approximately the brine production expected from a 3.3-mgd desalter operating at 70-percent recovery) the treatment technology performance criteria can be applied sequentially along the treatment train until the brine is completely disposed of.

For example, for the process combination of Brine A followed by chemical softening, then reverse osmosis, and then by a brine concentrator and then finally a crystallizer, the configuration would be:

- 1-mgd (2628 L/min) of Brine A to the softening unit.

- The softening unit would produce 1-mgd (2628 L/min) of softened brine, which would be fed to a 1-mgd (2628 L/min) secondary RO process.

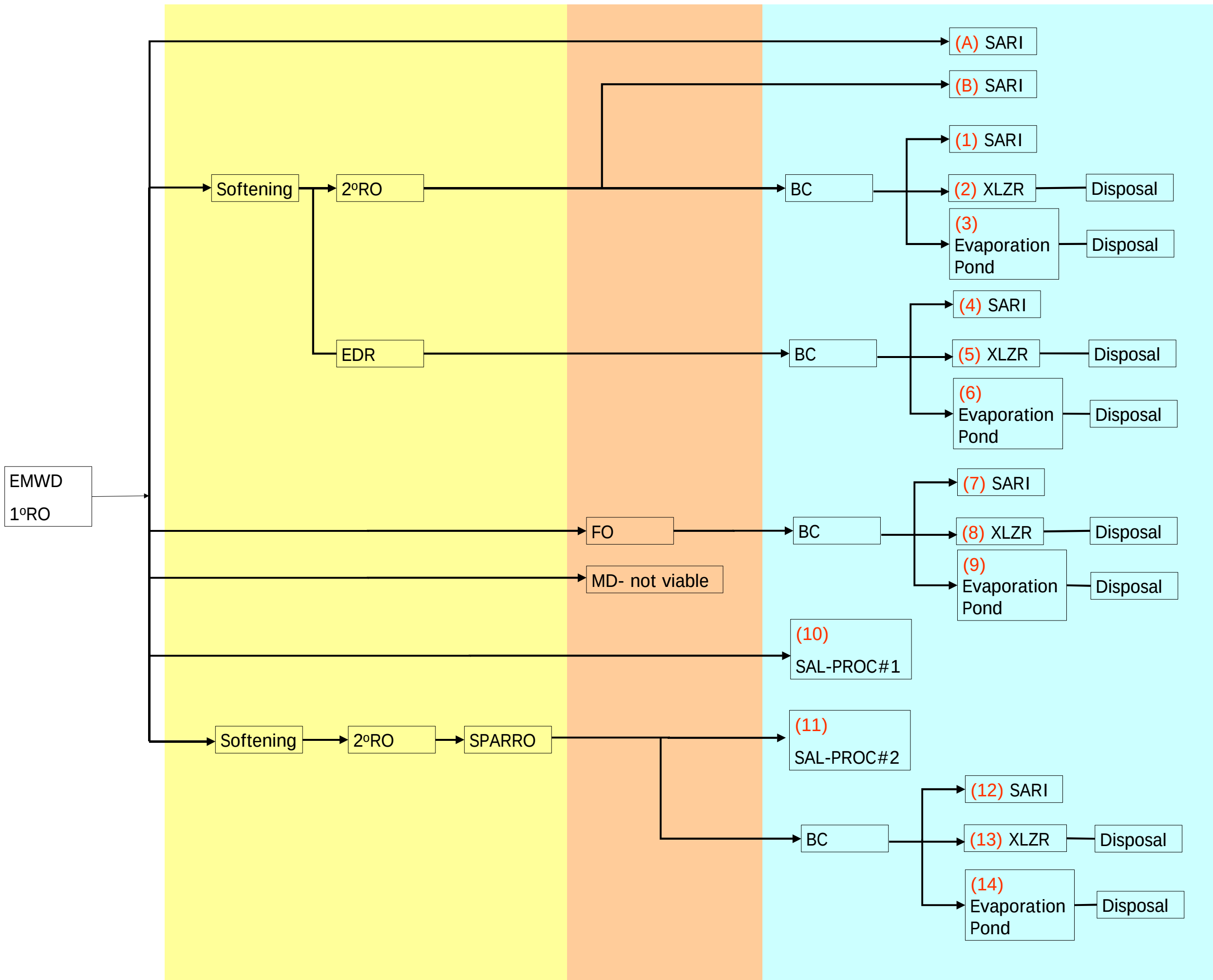
- 0.3-mgd (788 L/min) of Brine B would be produced by the secondary RO process, which would be fed to a 0.3-mgd (788 L/min) brine concentrator.

- The brine concentrator would produce 0.05-mgd (131 L/min) of highly concentrated brine (about 250,000 mg/L) which would be fed to the crystallizer.

- The crystallizer would produce 3 tons/d of solids, which would be trucked to a landfill for final disposal.

- More than 97 percent of the initial 1-mgd (2628 L/min) of brine would be recovered as potable water.

The performance results from the desktop modeling, bench-scale and pilot testing were used to develop 14 treatment trains that would support the minimization (or elimination) of brine. Two “base case” scenarios were also evaluated, for comparison. These were firstly the status quo brine disposal route; which is disposal of the primary RO brine directly to the SARI line. The second base case considered was disposal of brine resulting from chemical softening followed by secondary desalting (using RO) (i.e. Brine B) to the SARI line. Conceptual line diagrams showing the proposed treatment trains are provided in Figure 2.



LEGEND:

(RED) = Treatment Option number

= Pilot Test

= Bench Scale Test

= Desktop Simulation

Figure 2. Treatment alternatives developed or Brines A, B, and C including two base cases

Since the SARI disposal alternative is available to EMWD, additional alternatives also included the use of the SARI line as a final disposal approach. It is recognized that this may be impractical given the very high concentration of the final brine streams, but nevertheless the SARI disposal alternative was included to provide a comparative cost.

RESULTS AND DISCUSSION

As seen from Figure 2, 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 softening for calcium, silica and magnesium removal in a pilot-scale softening unit before secondary treatment by either RO or EDR. Results showed that the brine from the secondary RO and EDR processes was similar in quality, and was designated as Brine B. This secondary brine stream was introduced as feed to the SPARRO process, and the resulting concentrate stream was termed Brine C.

As mentioned, samples of Brines A and B were sent to UNR for FO and MD bench-scale experiments to test the viability of these two processes for brine minimization. MD of both Brine A and Brine B required high energy for heating the feed stream and resulted in catastrophic membrane scaling and low water flux. For these reasons it was decided that it would not be possible to estimate large scale performance and so MD was removed from further evaluation.

FO, 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, due to the lower operating flux. 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. UNR used the performance results to size the process to treat 1-mgd of brine influent and computed the energy requirements and equipment sizing for such a system. The resulting information was incorporated into the overall 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.

Salt recovery equipment cost estimates provided by Geo-Processors for treatment of Brines A and C, were used as-is in the cost models, except that interest rates were adjusted to be consistent and RO, BC/XLZR and landfill costs were modified to match the costs estimates of other process trains to allow for common comparisons.

Modeled recoveries and anticipated amounts of solids generated were used for evaporation pond sizing, in addition to solid disposal/landfill costs estimated from prior projects.

Table 4 summarizes the overall costs for all 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. All process alternatives are able to recover between about $400,000 and $600,000 per year for potable water generated. The reason that the revenue from the sale of salt byproducts is excluded from the analysis is that the revenue is likely to go to the operator of the SAL-PROC™ system, rather than EMWD, if such a system were to be implemented.

Two base cases are included for comparison. The first base case assumes that the primary RO brine (Brine A) stream does not undergo further treatment and is discharged into the SARI line at 1-mgd capacity, the current situation. The second base case considers softening and secondary RO treatment of the primary brine and discharges the remaining 0.3 mgd (788 L/min) of secondary RO brine (Brine B) to the SARI line (refer to Base Case B in Table 4). For this study, it was assumed that the excess SARI capacity would not be sold when calculating the costs for brine disposal to the SARI line. In other words, no income was assumed from the sale of excess SARI line capacity for any treatment alternative.

Among the 14 treatment alternatives evaluated, 10 are true zero-liquid discharge systems, and the remaining 4 include discharge of more concentrated brine to the SARI ocean outfall. The treatment combinations that include final disposal to the SARI line (A, B, 1, 4, 7, 12) are shown in green in Figure 3, which presents a summary of the annualized capital, annual O&M and total annual costs.

Table 4 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^{o}RO + SARI	=	**$14,745,375**	**$1,285,569**	**$978,165**	**$2,263,734**
1	Softening + 2^{o}RO + BC + SARI	=	**$32,085,114**	**$2,798,326**	**$2,741,738**	**$5,540,064**
2	Softening + 2^{o}RO + BC + XLZR-Disposal	=	**$29,151,685**	**$2,541,577**	**$3,413,113**	**$5,954,690**
3	Softening + 2^{o}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^{o}RO + Seeded RO + SAL-PROC#2-Disposal	=	**$27,826,445**	**$2,430,876**	**$4,343,651**	**$6,774,528**
12	Softening + 2^{o}RO + Seeded RO + BC + SARI	=	**$35,279,829**	**$3,081,696**	**$2,314,582**	**$5,396,278**
13	Softening + 2^{o}RO + Seeded RO + BC + XLZR-Disposal	=	**$30,621,940**	**$2,674,600**	**$2,669,835**	**$5,344,435**
14	Softening + 2^{o}RO + Seeded RO + BC + Evap Pond-Disposal	=	**$27,969,018**	**$2,444,306**	**$2,763,125**	**$5,207,431**

Notes:
(1) Total Cost per annum = Annualized Capital Cost + O&M Cost.
(2) Costs do not include primary desalting step and revenue generated from the sale of potable water or salt by-products.

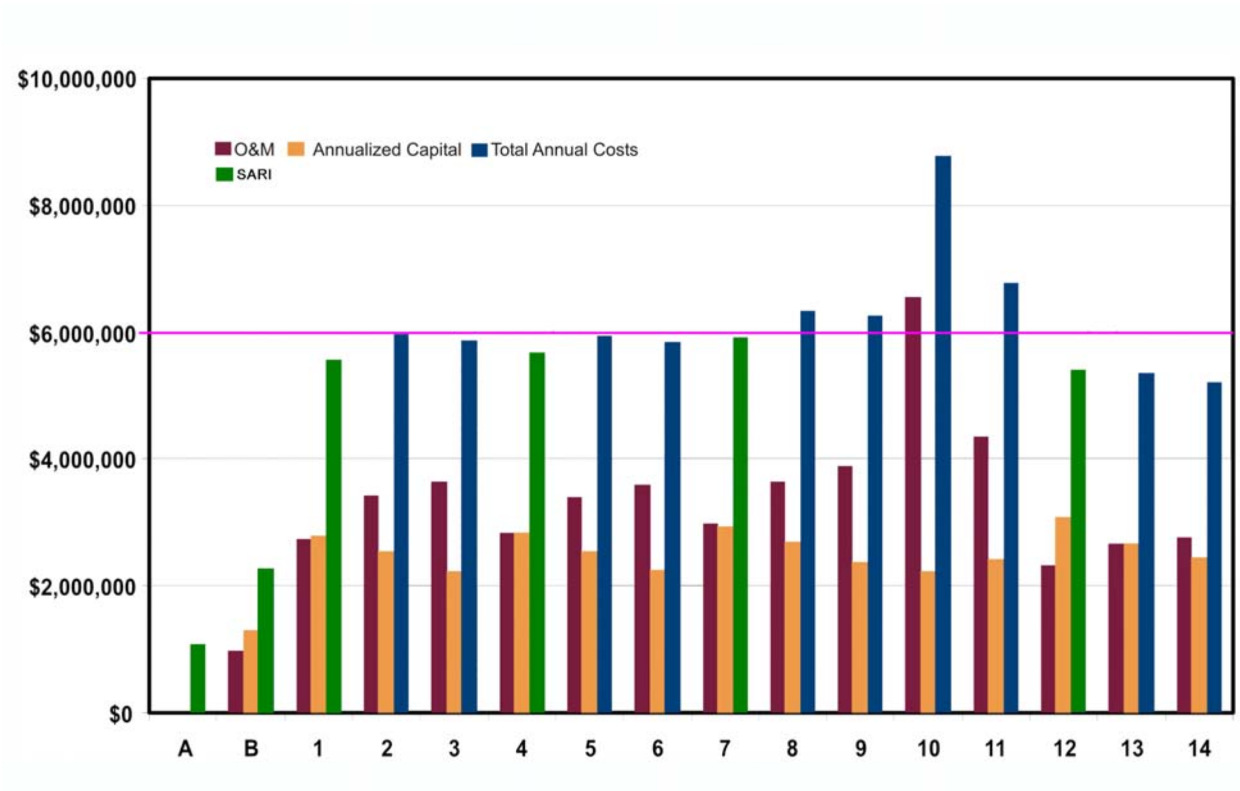


Figure 3. Total annual costs for various treatment combinations evaluated

From Figure 3, it is apparent that the treatment alternatives (apart from the base-case alternatives) have a fairly similar total annual cost, and seem to be within about 10-percent of each other; with the exception of the two SAL-PROC™ processes (Alternatives 10 and 11), which are higher. Also, Alternatives 12 through 14 seem to be lower than the 10-percent cost band.

If 1 mgd of Brine A discharged from EMWD's primary RO desalter undergoes SAL-PROC™ treatment via Alternative 10, it would cost $8.7 million/yr. As a matter of interest, the revenue anticipated from the sale of salt byproducts would be about $2.0 million per year. For Alternative 11, treating 0.15 mgd (394 L/min) of Brine C by the SAL PROC™ approach, the total annual cost would be about $6.8 million. Revenue anticipated from salt sales in this case would be about $0.6 million per year. The higher costs for the SAL-PROC™ alternatives seem reasonable considering that these systems use similar technologies to the other alternatives (RO, BC, XLZR), but also include processes for washing and processing the salt by-products.

Treatment alternatives that consider RO and EDR interchangeably (Alternatives 1 to 6), in addition to brine concentrators, are estimated to be between $5.5 to $5.9 million/yr. Brine output generated by concentrators can be directed to a crystallizer or an evaporation pond, where the remaining solids will be disposed to landfill. Bypassing the thermal evaporation step and discharging directly to SARI (if considered practical) would lower the treatment costs slightly. Option 6 would be the lowest true ZLD alternative in this group, owing to its slightly lower O&M cost between EDR and RO. This alternative includes an evaporation pond (up to a maximum of

12 acres (48,562 m^2). Without the availability of such land, brine concentration followed by crystallizers would be the logical choice as a ZLD alternative.

Process combinations involving FO (Alternatives 7 to 9) were slightly higher than those of Alternatives 1 to 6 ($5.9 to $6.3 million/yr). It is challenging to price this technology due to the absence of pilot-scale demonstrations and suitable commercial membranes. 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 the assumed 70-percent recovery. Further work using FO would be warranted based on these results.

It is interesting to note that in Alternatives 12 to 14, using seeded RO to recover water from Brine B results in the lowest total costs among the 14 ZLD alternatives considered. Although the current pilot tests have demonstrated that this technology is capable of recovering secondary RO brine at about 60 to 65 percent, higher recoveries may be possible. This is thought to be the first application of this technology to treatment of RO brine, and to date there have not been long term studies on the treatment of brine to establish the robustness of the membranes to the presence of the gypsum seed crystals scouring the membrane surface. The membranes used for the pilot testing were tubular NF membranes, because tubular RO membranes were not readily available. Previous work on mine water effluent (Juby et al., 1996) using cellulose acetate RO membranes showed that high recoveries were possible with this process over long continuous operating periods (over 3,000 hours). The positive results suggest that further evaluations with this process should be made.

The annual power consumptions (kW-hr/year) for the 14 alternatives are shown in Table 5. This represents the amount of energy required in addition to the primary RO process. Processes that include brine concentration and crystallization increase the power consumption by 7-12 times over Base Case B, which considers only softening followed by secondary RO treatment. Amongst the options that include brine concentrations and crystallization, power consumption for processes that include seeded RO (Alternatives 12, 13, 14) are 30-percent less energy intensive than processes that include secondary RO (Alternatives 1,2,3), and about 38-percent less energy intensive than processes that include EDR (4, 5, 6). By reducing the amount of concentrated brine that enters the brine concentrator, the energy requirements are lowered significantly. For the proprietary SAL-PROC™ approach, the energy requirements were not available per unit process but were provided as simulated values. These values appear to be considerably lower than expected, given that the SAL-PROC™ treatment trains are similar in configuration to Alternative 2, and include additional salt extraction steps.

Bearing in mind the limitations of this study (see below) in terms of developing accurate cost estimates, the results indicate that in general the total annual cost of providing a ZLD treatment alternative is relatively insensitive to the combination of technologies selected. For most process combinations (even those that include discharge of a concentrated stream to an ocean outfall), the total annual costs are within about 10 percent of each other. The alternatives outside this range include the salt recovery alternatives (which were higher) and the alternatives involving a third concentrating step ahead of the brine concentration stage (in this case the SPARRO process). For the latter treatment combinations (Alternatives 12, 13 and 14) the impact of the third concentrating step would be to reduce the capacity of the brine concentrator. Since the brine concentrator has the largest unit capital and operating cost, reducing the capacity of these systems reduces the overall treatment costs. These results suggest that for ZLD systems, or very high recovery systems discharging to the ocean, there are benefits to concentrating the brine as much as possible before employing a very high cost, high energy brine concentrator. In this study the highest brine concentration produced was around 22,000 mg/L from the SPARRO process. In a continuous SPARRO treatment system, higher recoveries may be possible, which could increase the brine concentration to above 50,000 mg/L, and reduce the feed flow rate to the brine concentrator by about 25 percent - reducing the overall costs further. Under these conditions a reverse-osmosis type system would be nearing the typical limits of today's commercially available pressure vessels.

Table 5 Annual Estimated Power Consumption of Treatment and Disposal Alternatives

Process Combinations:												Power Consumption kW-hr/yr
A										SARI		
B	Softening	+	2^oRO	+						SARI	=	1,104,000
1	Softening	+	2^oRO	+			BC	+		SARI	=	9,376,900
2	Softening	+	2^oRO	+			BC	+		XLZR-Disposal	=	10,836,800
3	Softening	+	2^oRO	+			BC	+		Evap Pond	=	9,376,900
4	Softening	+	EDR	+			BC	+		SARI	=	10,876,600
5	Softening	+	EDR	+			BC	+		XLZR-Disposal	=	12,140,700
6	Softening	+	EDR	+			BC	+		Evap Pond-Disposal	=	10,876,600
7	FO	+					BC	+		SARI	=	11,694,000
8	FO	+					BC	+		XLZR-Disposal	=	13,153,900
9	FO	+					BC	+		Evap Pond-Disposal	=	11,694,000
10										SAL-PROC#1-Disposal	=	2,325,000
11	Softening	+	2^oRO	+	Seeded RO			+		SAL-PROC#2-Disposal	=	1,899,700
12	Softening	+	2^oRO	+	Seeded RO	+	BC	+		SARI	=	6,640,000
13	Softening	+	2^oRO	+	Seeded RO	+	BC	+		XLZR-Disposal	=	7,529,500
14	Softening	+	2^oRO	+	Seeded RO	+	BC	+		Evap Pond-Disposal	=	6,640,000

Notes:
(1) Alternative A is considered as the base case and the power consumptions for the subsequent alternatives are the excess power required in addition to the base case.
(2) Costs do not include pumping to the regional brine line (SARI).

LIMITATIONS OF STUDY

As mentioned, several of the treatment technologies considered in this study are in early phases of development (particularly MD, FO and SPARRO) and therefore estimating the capital and operating costs for full-scale systems is challenging.

Efforts to standardize the cost information simulated by commercial system providers to pilot tested processes have been illustrated in this study, but barring the intimate knowledge of all assumptions used in the simulations, not all differences can be resolved. This is apparent in the power consumption variations for proprietary processes such as SAL-PROC™.

Nonetheless, if the cost conclusions are used with the above in mind, the data should still be useful to compare different treatment train combinations.

CONCLUSION

This study has shown that:

1. There are several treatment combinations that could be used to provide a ZLD system.

2. The ZLD process combinations evaluated had similar total annual costs, given the level of cost accuracy in this study.

3. Process combinations that produced the lowest flow rate of brine to the thermal evaporation processes appeared to result in the lowest costs.

4. Emerging technologies like MD, FO and SPARRO require further development before these processes can be commercially available at more competitive pricing.

5. The cost model developed will not only benefit EMWD, but will also benefit other local, state, regional, and national agencies facing brine disposal issues from inland desalination facilities.

6. Bearing in mind the limitations of this study in terms of developing accurate cost estimates, the results indicate that in general the total annual cost of providing a ZLD treatment alternative is relatively insensitive to the combination of technologies selected. For most process combinations (even those that include discharge of a concentrated stream to an ocean outfall), the total annual costs are within about 10 percent of each other. The alternatives outside this range include the salt recovery alternatives (which were higher) and the alternatives involving a third concentrating step ahead of the brine concentration stage (in this case the SPARRO process), which were lower. For the latter treatment combinations (Alternatives 12, 13 and 14) the impact of the third concentrating step would be to reduce the capacity of the brine concentrator. Since the brine concentrator has the largest unit capital and operating cost, reducing the capacity of these systems reduces the overall treatment costs. These results suggest that for ZLD systems, or very high recovery systems discharging to the ocean, there are benefits to concentrating the brine as much as possible before employing a very high cost, high-energy brine concentrator.

7. For inland communities where access to an ocean discharge line is not a viable option, brine minimization by brine concentrators and further crystallization prior to land filling are comparable to thermal brine concentration followed by an evaporation pond. Although the capital costs of BC/XLZR are greater than evaporation ponds, the O&M costs for the BC/XLZR alternatives are slightly lower ($200,000 per year) and more water is recovered. In the long run, the use of thermal processes may be a “greener” solution than dealing with the upkeep of an evaporation pond – which can occupy as much as 12 acres of land for the treatment alternatives evaluated, particularly if a renewable source of energy (such as solar or wind) could be used to supply the large energy load of the brine concentrator.

ACKNOWLEDGEMENTS

The authors would like to acknowledge the funding provided for this project by the United States Department of the Interior, Bureau of Reclamation through the Desalination and Water

Purification Research and Development Program, and to the assistance provided by Mr. Scott Irvine. Funding was also provided by the California Department of Water Resources as part of the 2004 Water Desalination grant program (Chapter 6(a) of Proposition 50).

REFERENCES

Juby, G.J.G, Sethi, S., Li, S. Zacheis, G.A., Mulvihill, T., Standard, W., Morquecho, R., Krieger, C. and Scriven, D., (2006) Feasibility Study Evaluates Alternatives to Provide Zero-Liquid Discharge for Inland Groundwater Desalter Project, 21st Annual WateReuse Symposium, Hollywood, CA.

Juby, G.J.G., Schutte, C.F. and van Leeuwen, J., (1996) Desalination of Calcium Sulphate Scaling Mine Water: Design and Operation of the SPARRO process, Water SA, Vol. 22, No. 2, pp 161 to 172.

Mickley, M. (2001) Membrane Concentrate Disposal: Practices and Regulation. Denver, Colorado: Bureau of Reclamation.

U.S. Department of the Interior (2008) Evaluation and Selection of Available Processes for a Zero-Liquid Discharge System for the Perris, California, Ground Water Basin, Desalination and Water Purification Research and Development Program Report No. 149.

ACRYONYMS

EMWD	Eastern Municipal Water District
DWR	Department of Water Resources
USBR	United States Bureau of Reclamation
ZLD	Zero Liquid Discharge
RO	Reverse Osmosis
EDR	Electrodialysis Reversal
FO	Forward Osmosis
MD	Membrane Distillation
TDS	Total Dissolved Solids, mg/L
SARI	Santa Ana Regional Interceptor
SPARRO	Slurry Precipitation and Recycle Reverse Osmosis
UNR	University of Nevada, Reno
VEDCMD	Vacuum Enhanced Direct Contact Membrane Distillation