RO CONCENTRATE TREATMENT AND REUSE TO ACHIEVE SUSTAINABLE WATER MANAGEMENT SCHEMES FOR INDUSTRIAL FACILITIES LOCATED IN ARID, WATER SCARCE REGIONS

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Abstract

Alternative approaches were developed and costed to treat and recover RO reject generated from high purity water production within an electronics facility located in the southwestern US, where reject disposal to a surface water body or wastewater treatment plant would exceed allowable total dissolved solids (TDS) limits. Technologies utilized in the development of alternatives included those that employed a combination of precipitative softening using lime, caustic soda and soda ash implemented using pellet and conventional high-rate solids contact reactors and tubular microfiltration; granular media and membrane filtration; brackish and seawater RO using conventional and special configurations (e.g., HERO™, OPUS™, and disc tube); followed by mechanical-driven evaporation and crystallization. This arrangement was compared with an approach that used only evaporation and crystallization. The alternatives analysis revealed that the all-distillation alternative, although higher in capital cost, was estimated to have lower life cycle cost along with several non-cost benefits.

Background

As water-intensive industries such as specialty manufacturing, power generation and mining continue to locate in relatively arid and water-constrained geographies, they are faced with water supply and wastewater disposal issues that are becoming increasingly complex and expensive. Due to the both geographic and hydrogeologic conditions, arid-region water supplies have elevated levels of total dissolved solids (TDS) and inorganics, requiring the use of demineralization technologies, most typically reverse osmosis (RO), as a key step to the production of high quality process water. The disposal of the high salt waste produced from the RO, called reject (or concentrate), is a challenge because of the lack of naturally-occurring surface waters in which to discharge; or limitations on disposal to sewer because treated effluent is used for landscape irrigation or other non-potable uses. These constraints—lack and quality of water supply, and difficulties with high-salt discharges—are forcing industry to develop a more comprehensive approach to water management that incorporates water reuse and recycling. These strategies include innovative approaches to high-salt RO reject management, and in an effort to reduce overall water demands, reuse treated wastewater and limit the volume of RO concentrate requiring disposal.

This paper describes a water management study conducted in the arid southwestern U.S. to evaluate treatment options for a high salinity RO reject that is produced as a waste stream from high purity water production. Increasing RO reject flows, coupled with a fixed volume within existing solar evaporation ponds for concentrate disposal, required the development of new approaches to reducing overall reject flows. Furthermore, water supply limitations required the treatment and reuse of the clean water recovered from the reject.

Technologies considered in the development of reject treatment and reuse schemes for this study included precipitative softening using lime, caustic soda and soda ash followed by conventional and novel solids clarification processes; tubular and hollow fiber micro- and ultra-filtration, nanofiltration (NF), electrodialysis reversal (EDR) and various configurations of RO (brackish and seawater spiral wound, high efficiency RO [HERO™], and disc tube); thermally and electrically-driven evaporation; and passive and enhanced solar evaporation ponds.

Water Quality and Reject Flow

The cumulative total RO reject produced from the manufacturing campus' high purity water production facility is estimated to be approximately 2,000 gpm. This reject is essentially concentrated city (potable) water. City water is sourced from groundwater from local wells and surface water supplied from state or regional storage and conveyance systems. For the purpose of this study, historical groundwater data was assumed as the water supply since it has higher concentrations of TDS and inorganics that would result in the production of higher-salinity RO reject. To estimate the quality of the RO reject, IMSDesign™ software was used based on the RO system design summarized in Table 1. The historical groundwater and projected RO reject quality are presented in Table 2.

Table 1. RO system design criteria.

Parameter	Value
RO Membrane Type	ESPA 2
RO Average Flux	13.2 gfd
RO Recovery	88.0%
No. Stages	3 stage

Table 2. Historical groundwater quality and projected RO reject quality.

Parameter	Unit	Groundwater	Projected RO Reject
Ca	mg/L	72.0	596
Mg	mg/L	31.0	257
Na	mg/L	300	2,419
K	mg/L	4.0	32.0
NH4	mg/L	0.10	0.80
Ba	mg/L	0.066	0.55
Sr	mg/L	1.2	9.93
CO3	mg/L	2.2	0.600
HCO3	mg/L	232	1,516
SO4	mg/L	134	1,418
Cl	mg/L	454	3,702
F	mg/L	0.6	4.8
NO3	mg/L	22.0	156
B	mg/L	0.22	0.45
SiO2	mg/L	26.0	213
CO2	mg/L	2.94	35.1
TDS	mg/L	1,280	10,324
pH		8.10	7.77

Technology Overview

A variety of technologies were considered to treat the RO reject to minimize brine flows to meet the capacity of existing solar evaporation ponds, and to produce a high quality permeate suitable for reuse that would maximize overall water recovery. Membrane-based desalination was assumed to be the most effective technology. However, since the high purity RO's at the site are operated at maximum practical recovery resulting in high levels of scaling salts in the untreated RO reject, pretreatment is required to soften the water to maximize recovery of the desalination process. Furthermore, to overcome osmotic pressure (RO) or voltage (EDR) limitations, additional brine concentration technologies downstream of these processes are needed to achieve a brine flow of a suitable volume for disposal in the allotted acreage of evaporation ponds. Figure 1 shows a process flow diagram of the basic treatment train and key processes. In the subsequent sections, the various configurations considered for each major process are discussed.

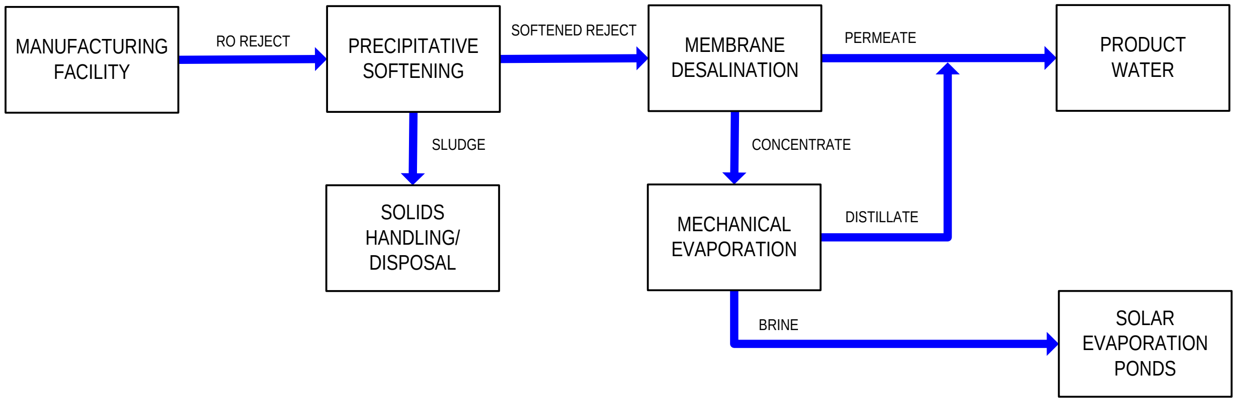

Figure 1. Process flow diagram.

Precipitative Softening

Precipitative softening was selected as the first step in the process to remove calcium, magnesium, alkalinity and silica from the reject. The softening reactions also co-precipitate barium, strontium and fluoride, which can also form precipitates and limit the recovery of the downstream desalination process. Various chemistries and equipment types can be used to perform the softening, depending on the effluent quality requirements. The following combinations of chemicals were evaluated to achieve a softened water quality that would optimize performance of the desalination processes:

- ***Lime and soda ash –*** Lime is added to increase pH and soda ash to provide additional carbonate to precipitate the desired amount of calcium, as well as strontium and barium and fluoride. Significant silica removal is also accomplished as silica co-precipitates with magnesium hydroxide. The sludge produced is easily dewatered.

- ***Caustic soda and soda ash –*** This chemistry is similar to lime softening except that sodium hydroxide is used in place of lime to raise pH. Softening performance is similar to that using lime, however less solids are generated but chemical costs are higher due to the higher cost of caustic soda when compared to lime. The resulting solids are not as easily dewatered as with lime.

Once formed, the precipitated solids must be separated from the softened water, either by gravity or by a physical solids-liquids separation process such as a membrane.

The equipment types that were considered for precipitation and solids separation processes are presented below in Table 3.

Table 3. Precipitative softening technologies.

Equipment Type	Description	Advantages	Disadvantages
Conventional Solids Contact Reactor/Clarifier (SCRC)	Traditional softening using a circular clarifier with perimeter overflow and an internal rake for removal of settled solids from the bottom of the clarifier.	Proven track record.	Large footprint.
High Rate SCRC	Modified design of a conventional clarifier with enhanced reaction and solids contact zones allowing effectively higher solids concentration in smaller footprint compared with conventional clarifier.	Smaller footprint compared to SCRC. Proven track record.	Some additional chemicals required to enhance precipitation.
Two-Stage Mixed Reaction Tanks Followed by Tubular Microfiltration	Softening reactions completed in a two-step process followed by solids separation tubular microfilter operated in cross-flow/feed-and-bleed mode with continuous solids blow down from the MF feed tank.	Provides high quality softened water quality without need for subsequent filtration. Smaller footprint than SCRCs.	Periodic system cleaning required. Higher capital costs.
Pellet Reactor (followed by SCRC)	Fluidized bed design that results in dense, pebble-like $CaCO_3$ solids that require little subsequent dewatering. Due to reaction time and solids formation mechanism, it does not effectively precipitate magnesium hydroxide and silica, requiring a subsequent SCRC of at least the same size as required for the overall lime softening.	Very small footprint. Very high percent solids generated reduces dewatering costs.	Subsequent reactor required to precipitate magnesium hydroxide and silica. Not proven for high TDS applications.

Limited bench testing was combined with process modeling of the various precipitative softening approaches to determine which would be most cost effective based on chemical doses and reaction times needed to produce the target softened water quality. Bench testing was particularly valuable as an adjunct to modeling in that it identified the need for either increased reaction times and/or increased chemical doses to achieve the desired removals of calcium and magnesium. Combining the results of the modeling, bench testing, site specific requirements (e.g., footprint), together with cost, lime and soda ash chemistry executed using a high rate SCRC was considered to be the most cost-effective and reliable softening and clarification process to achieve the desired water quality. Table 4 presents the predicted quality of the precipitative softened reject and Table 5 presents the predicted operating conditions of the SCRC.

Table 4. Predicted water quality from precipitative softening.

Parameter	Unit	Projected RO Reject	Softened Product
Ca	mg/L	596	48.4
Mg	mg/L	257	2.40
Na	mg/L	2,419	3,388
K	mg/L	32.0	32.0
NH4	mg/L	0.800	0.800
Ba	mg/L	0.546	0.100
Sr	mg/L	9.93	5.53
CO3	mg/L	0.600	295
HCO3	mg/L	1,516	4.99
SO4	mg/L	1,418	1,417
Cl	mg/L	3,702	3,701
F	mg/L	4.80	0.50
NO3	mg/L	156	156
B	mg/L	0.450	0.00
SiO2	mg/L	213	10.0
CO2	mg/L	35.1	0.0
TDS	mg/L	10,324	9,096
pH		7.77	11.30

Table 5. SCRC operating conditions.

Parameter	Value
Reactor Residence Time	12 min
Clarifier Residence Time	21 min
Surface Overflow Rate	8.9 gpm/sq ft
Lime Dose	1,450 mg/L
Soda Ash Dose	1,720 mg/L
Ferric Chloride	56 mg/L
Polymer	2 mg/L
pH	11.3

Softened Water Filtration
The clarified, softened reject must then be filtered to remove residual suspended solids prior to desalination. This can be achieved using either granular media filtration (GMF) or membrane filtration (microfiltration/ultrafiltration [MF/UF]). The former is less expensive (capital and operations), does not produce a chemical waste stream, has a lower footprint, can be housed outdoors and generates less overall wastewater flow. The primary disadvantage of GMF is that the filtered water produced will be of lower quality compared to that produced membrane filtration. This is only important if the downstream desalination process being utilized is RO. If an electrochemical process such as EDR is utilized for downstream desalination, no negative impacts will be experienced from using GMF. The better and more consistent filtered water quality (lower turbidity and SDI) produced by MF/UF is more critical when using RO for desalination and will reduce fouling and the need for chemical cleaning. Both GMF and MF/UF can be used for filtration, but the process and cost tradeoffs must be weighed. For this application, the cost and footprint advantages for GMF outweighed the filtrate quality advantages inherent to MF/UF.

Desalination and Volume Reduction
After reducing its scale-forming tendencies, the RO reject can be processed by one of two energy efficient desalination processes: RO and EDR. Two different RO configurations were evaluated based on a conventional design and the High Efficiency Reverse Osmosis (HERO™) process. A description of each, along with advantages and disadvantages, are listed in Table 6. OPUS™ (Optimized Pretreatment Unique Separation), which, like HERO™, employs weak acid cation exchange but without alkalinity removal, was also considered and found to be comparable in cost to the HERO™ process for this particular feedwater. EDR, an electro separation process that utilizes ion exchange membranes, was evaluated because of its potential to achieve a higher recovery than RO. The disadvantage of EDR is that it produces a lower quality product water than RO. The projected EDR product did not meet the re-use requirements, particularly for TOC and silica, and would require polishing using additional processes (such as RO). For these reasons, EDR was not considered any further. Table 6 provides a summary of the three desalination systems examined and the respective advantages and disadvantages of each.

In comparing HERO™ (or OPUS™) to conventional RO, the primary value of the two proprietary technologies lies in their ability to achieve maximum recovery without being limited by mineral precipitation, particularly that of silica. However, the scaling potential of the softened reject (as shown in Table 3) is sufficiently low that a similar recovery can be achieved both conventional RO and HERO™/OPUS™ because recovery is limited by feed (osmotic) pressure rather than saturation of sparingly soluble salts. In both cases, a nominal recovery of 89 percent can be achieved based on a maximum feed pressure of 1,065 psig, reducing the brine flow from 2,000 to 215 gpm. This is illustrated in Table 7. With the softened RO reject, maximum recovery is in large part dictated by the salinity of the feed. Based on achieving similar recoveries, conventional RO was considered to be the most appropriate desalination technology because it has the lowest capital and operating cost. Still, it should be noted that HERO™ and OPUS™ offer additional process and operational benefits over conventional RO when treating a surface water-based RO reject because of improved TOC removal and reduced organics fouling at high feedwater pH. Additionally, prevention of mineral precipitation with HERO™ and OPUS™ is not dependent on the advertised performance of a scale inhibitor. If the RO reject contained higher levels of silica, HERO™ or OPUS™ would have been the preferred membrane desalination process.

Table 6. Membrane-based desalination technology summary.

Technology	Description	Advantages	Disadvantages
Conventional RO	Utilizes spiral wound membranes and high pressure.	Proven technology, capable of producing high quality permeate.	Recovery is limited by scaling ions not removed by precipitative softening.
HERO™	Patented process that includes weak acid cation exchange and decarbonation to remove remaining hardness and alkalinity, with the zero hardness/alkalinity water further processes by RO operating at high pH (~10.5).	High pH operation allows for very high recoveries by increasing silica solubility. High pH operation increases organics rejection organics and reduces organic fouling of RO process.	Additional capital and operating cost for ion exchange and CO2 removal facilities and HERO™ licensing.
EDR	An electrochemical separation process in which ions are transferred through ion exchange membranes by means of an electrical potential driving force.	Can achieve higher recovery than RO on given feedwater scaling potential.	Product water of lower quality than RO permeate and less suitable for reuse (higher TOC, higher TDS, no silica removal).

Table 7. RO design feed pressures and scaling potential for conventional RO and HERO™.

Parameter	Unit	Conventional RO	HERO™
RO Design			
No. Stages		5	5
Stage 1 Feed Pressure	psi	227	180
Stage 2 Feed Pressure	psi	408	359
Stage 3 Feed Pressure	psi	651	597
Stage 4 Feed Pressure	psi	855	797
Stage 5 Feed Pressure	Psi	1,062	1,002
Scaling Potential			
$CaSO_4$	% saturation	58	0.010
$BaSO_4$	% saturation	8,290	0.0
$SrSO_4$	% saturation	322	0.0
CaF_2	% saturation	1,260	26
SiO_2	% saturation	74	18
$Mg(OH)_2$	% saturation	0.0	0.800

Further Volume Reduction Using Mechanical Evaporation

To achieve further volume reduction, the concentrate from the conventional RO process can be processed by a mechanical evaporator. Traditionally, in the power and industrial sectors, mechanical evaporators (also known as brine concentrators) and crystallizers, which utilize thermal-based distillation to separate water from the desalination process brine, are often used to reach near-zero or zero liquid discharge. Because of their high capital and operating (energy) costs, a non-conventional RO technology using a 'disc tube' configuration that can operated at feed pressures up to 1,600 psig was also considered for additional volume reduction. Table 8 provides a summary of these technologies as well as their respective advantages and disadvantages.

Table 8. Conventional RO concentrate volume reduction technologies.

Technology	Description	Advantages	Disadvantages
Brine Concentrator	Thermal evaporation provided through compressor-driven system.	Provides effective brine minimization and high quality distillate	High energy requirements Requires exotic materials of construction
Crystallizer	Provides further brine concentration through compression driven system.	Provides effective brine minimization and very high quality distillate. Capable for reducing brine flows to a ZLD slurry.	Very high energy requirements. Requires exotic materials of construction.
Disc Tube Membrane System	Disc-tube membrane system was developed to have the advantages of an open channel and spiral wound module design.	Higher operating pressure range of than conventional RO (1,600 psi). Lower fouling potential due to spacer design.	Lower recovery than thermal processes. Capital costs only slightly lower than thermal processes.

A technical and cost evaluation of these technologies indicated that the thermal evaporative processes is the best choice for RO concentrate volume reduction, primarily due to the significantly higher recovery that can be achieved. Even at 1,600 psig feed pressure, the disc-tube RO technology can only reduce the conventional RO concentrate volume by 33 percent, whereas mechanical evaporation can achieve 67 percent recovery, allowing the conventional RO concentrate flow be reduced to 70 gpm. Recovery of the former is again limited by osmotic pressure considerations inherent with the RO process whereas recovery with the mechanical evaporator is limited by sodium sulfate solubility. The TDS of the brine discharged from the mechanical evaporator is estimated to be 270,000 mg/L consisting mainly of sodium sulfate and sodium chloride.

Evaporation Ponds

In addition to evaluating softening, membrane desalination and distillation technologies for volume reduction and water reuse, an evaluation of the existing solar evaporation ponds was conducted to determine their ability to evaporate the brine from the mechanical evaporator. The existing pond system is comprised of 5 ponds with a maximum depth of 5 ft and a maximum, combined, top surface area of 27 acres. An evaporation pond model was developed by CH2M HILL to determine the evaporative and hydraulic capacity of these ponds. Even though the ponds are all slightly different in size, for modeling purposes, it was assumed that the 27 acres was divided equally over all five ponds. Other inputs to the model besides pond area and depth include side slope ratio, monthly pan evaporation rates for the location region, monthly rainfall/precipitation data, as well as quantity and salinity of the brine flow.

Based on these inputs, the model was used to calculate real-time values for water level, salt accumulation, and ultimately, the duration of pond operation until the time that either overflow occurs or the ponds must be taken out of service to remove the accumulated salt. Pond water level is directly related to the water (brine) volume in the ponds. The brine TDS concentration in the ponds is calculated based on the mass of salt inflow divided by the brine volume on any given day. The model assumes a maximum saturation of 359 g/L (primarily related to the saturation of sodium chloride). When saturation is reached, a portion of the dissolved solids in the brine will precipitate, with the resulting salt settling to the bottom of the pond. Over time some of the precipitated salt may re-dissolve (e.g. during winter months when evaporation rates are lower) but over time precipitated salt must accumulate and reduce the available brine storage volume. To illustrate these changes, Figure 2 shows the change in brine (water) and salt depth in the ponds at an incoming brine flow and TDS of 50 gpm and 50,000 mg/L, respectively.

Through model runs at varying flows and brine TDS levels, salinity of the incoming brine was identified as the most critical factor affecting pond evaporation rate. At any flow, the higher the salinity, the shorter the period before water level reached overflow. Figure 3 shows a graph of the estimated time to overflow assuming varying salinity levels in the brine inflow and a fixed flow of 70 gpm. A TDS range of 10,000 mg/L to 350,000 mg/L could be present in the pond inflow depending on the type of treatment and degree of RO reject volume reduction implemented. The data in Figure 3 shows that the ponds will overflow more quickly as brine salinity increases because evaporation rate is lower at higher salinity. Additionally, salt accumulation is more rapid at higher brine salinities effectively reducing pond volume available for liquid storage, which further reduces time to overflow.

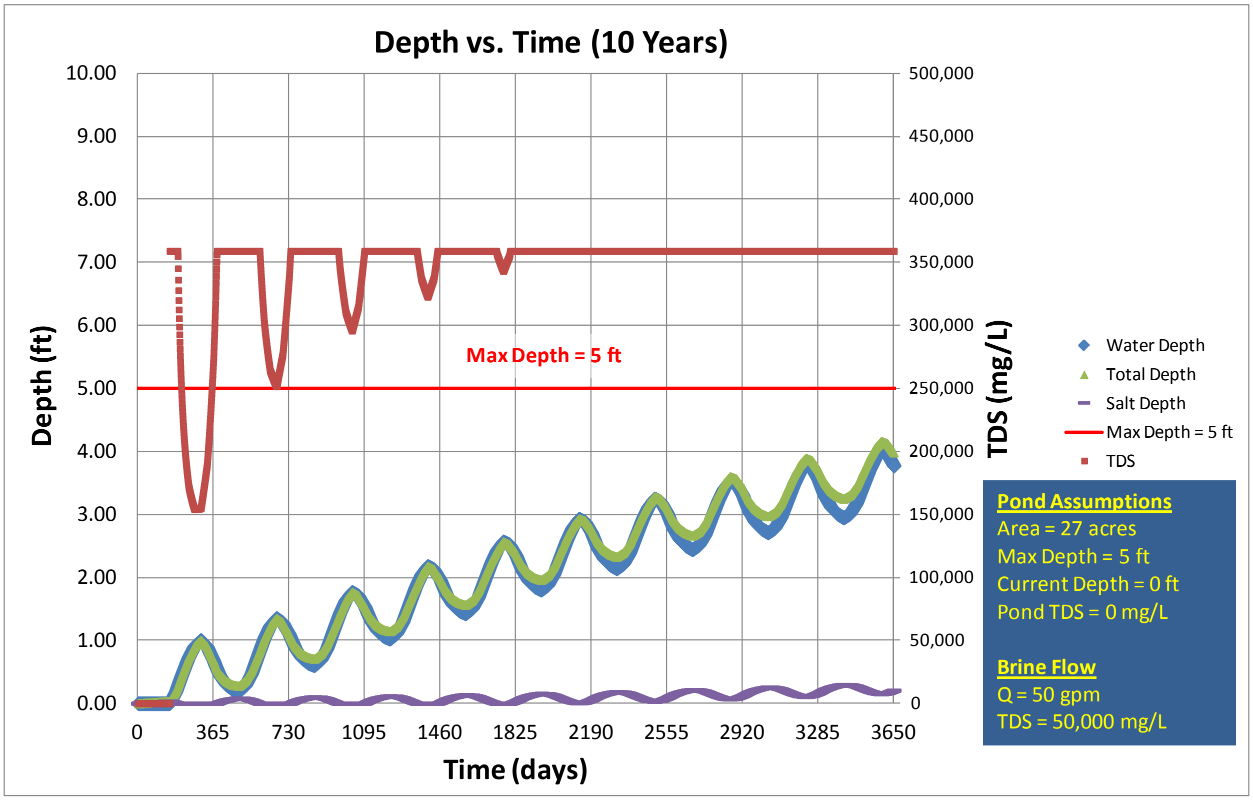

Figure 2. Pond and salt depth in evaporation ponds over time.

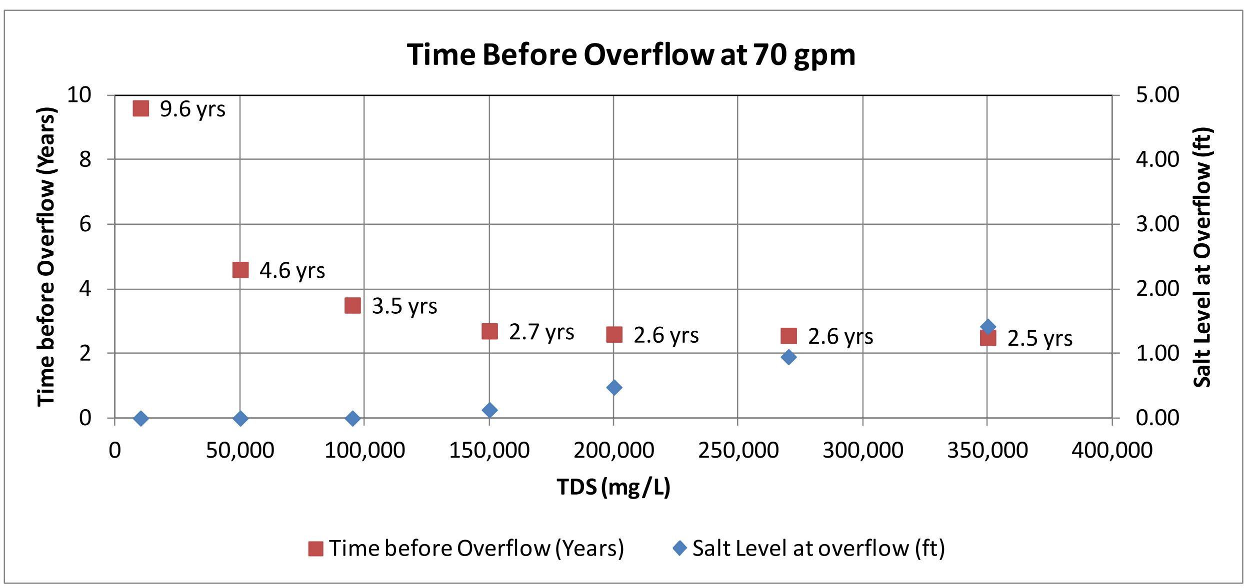

Figure 3. Time to pond overflow based on 70 gpm brine inflow and variable brine salinity.

A second analysis was conducted to determine pond capacity at various brine salinity levels based on a fixed operating period. In this analysis, 10 years was assumed as the pond "life" (after which time the pond would need to be emptied of salt and some or all of the brine). The model was used to determine the maximum flow the ponds could accept at each TDS without overflowing during ten years of operation. The results are shown in Figure 4, which further illustrate that pond capacity decreases with increasing brine salinity both because of reduced evaporation rate and more rapid accumulation of precipitated salt.

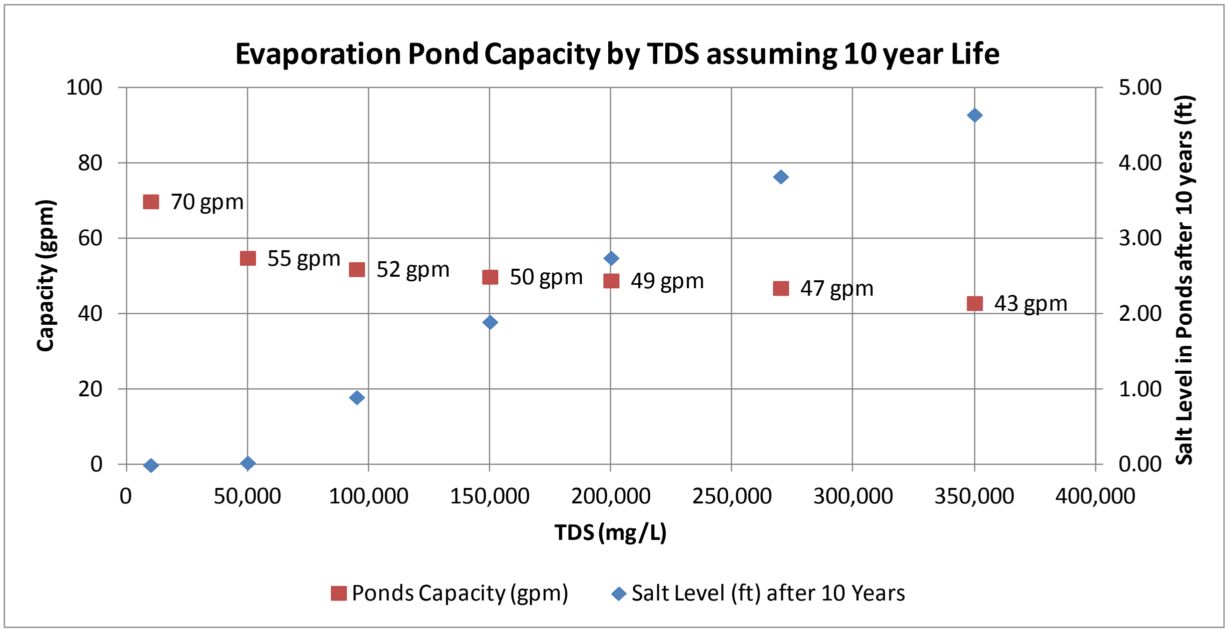

Figure 4. Pond capacity based on variable brine salinity/flow.

The largest reduction in pond flow capacity occurs between 10,000 and 50,000 mg/L brine TDS. At higher TDS levels, the brine reaches saturation relatively quickly and evaporation rate decreases to a minimum, resulting in the ponds having a relatively narrow, and low, evaporation capacity (43 – 52 gpm). Based on the RO reject flow shown in Table 2, a treatment system would be required to reduce the reject volume by between 96 and 98 percent in order to utilize the calculated capacity of the existing evaporation ponds, with a resulting brine TDS range of 250,000 to 350,000 mg/L. Such a high volume reduction and TDS concentration requires the use of not only precipitative softening, RO and mechanical evaporation, but also crystallization, as a brine concentrator would not be able to produce a brine with a salinity in the 250,000 to 350,000 mg/L range.

Alternatives Evaluation

Based on the survey of technologies and the results from evaporation pond modeling, a 'base case' solution for volume reduction of the RO reject was developed that includes precipitative softening, conventional RO, mechanical evaporation and a mechanically-driven brine crystallizer as shown in Figure 5. This treatment train would be capable of reducing the RO reject flow from 2,000 gpm to 43 gpm while recovering 1957 gpm as high quality water for reuse (combination of RO permeate and distillate). Table 9 summarizes the estimated capital and O&M costs for this alternative.

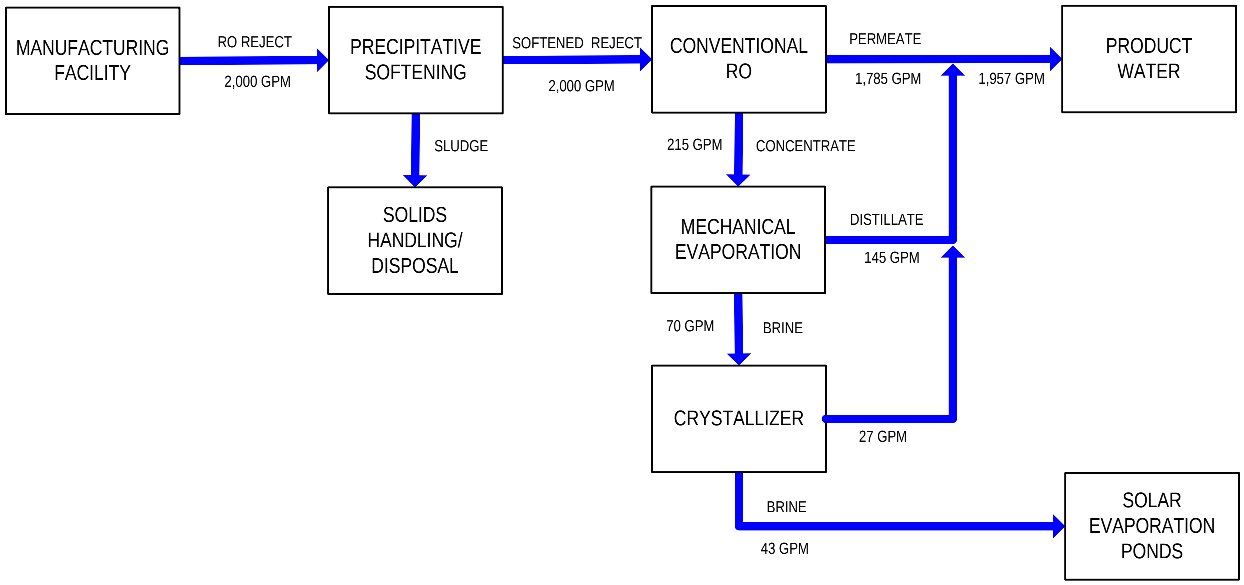

Figure 5. Base approach.

Table 9. Capital and O&M costs for base approach.

Process	Capital Cost[a]	kW	Annual Energy Cost	Annual Chemical Costs	Annual Solids Disposal Costs	Annual O&M Costs
Softening and Solids Handling	\$6.5 M	200	\$0.14 M	\$10.8 M	\$4.4 M	\$15.3 M
Conventional RO	\$10.0 M	650	\$0.46 M	\$0.70 M	-	\$1.16 M
Mechanical Evaporator and Crystallizer	\$15.0 M	1,325	\$0.93 M	\$0.03M	-	\$0.96 M
Total	\$31.5 M	2,175	\$1.53 M	\$11.5 M	\$4.40 M	\$17.4 M

[a]Cost of equipment only; assumes redundancy for all major equipment (SCRC, media filters, RO train, brine concentrator and crystallizer)

The costs presented show that the base case has a very high O&M cost, mainly because of the chemicals and solids disposal requirements associated with precipitative softening. These costs were significant enough to investigate an all-distillation approach that included multiple mechanical evaporators and a crystallizer, as shown in Figure 6. The alternative approach was developed in order to determine whether substituting the capital and O&M costs associated with a larger-capacity brine concentrator system would be less than the chemical- and solids-intensive base approach. It was also recognized that the alternative (all-distillation) approach would have a significantly reduced operations complexity because it simplified the treatment train.

Table 10 presents the capital and O&M costs for alternative approach. When contrasted to costs in Table, the alternative approach, while having a higher capital cost and annual energy cost, there is a net savings in O&M costs. A comparison of the life cycle costs (Table 11) shows that the alternative approach would result in cost savings of $30 million over 20 years (at 10 percent discount rate). In addition, this alternative would produce higher quality product water, be much simpler to operate due to the reduced number of unit processes and would eliminate truck traffic associated with delivery of lime and soda ash as well as removal of lime-based sludge. Because the operating cost of all-distillation approach is nearly all from electrical energy use (minor amounts of chemicals are used), it can also be considered a more sustainable solution if the electricity is provided from renewable sources (solar or wind or both). Based on these advantages and lower life cycle cost, the all-distillation approach was recommended for RO rejection volume reduction and water recovery (reuse).

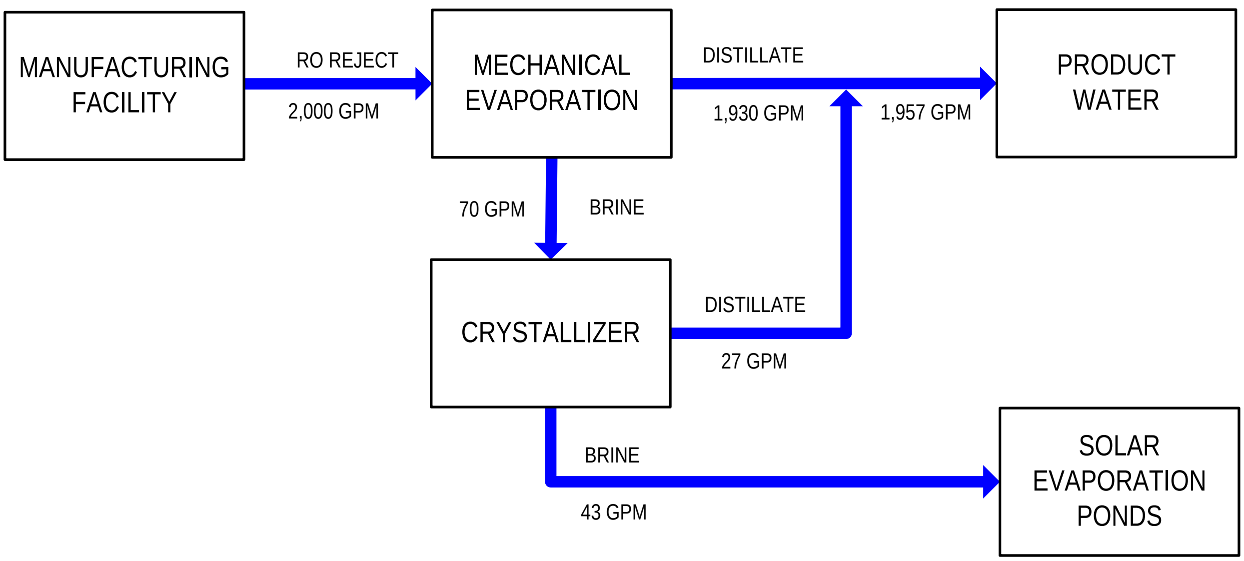

Figure 6. Alternative (distillation-only) approach.

Table 10. Capital and O&M costs for alternative approach.

Process	**Capital Cost**[a]	**MW**	**Annual Energy Cost**	**Annual Chemical Costs**	**Annual Solids Disposal Costs**	**Annual O&M Costs**
Mechanical Evaporator and Crystallizer	$44.0 M	10	$7.0 M	$0.6M	$4.9 M	$12.5 M

[a]Cost of equipment only; assumes redundancy for all major equipment (brine concentrator and crystallizer)

Table 11. Life cycle cost comparison between base and alternative approaches.

Parameter	Base Approach	Alternative Approach
Process Description	Softening, HERO, Brine Concentrator, Crystallizer	All Thermal (Distillation) Approach
Capital Cost	\$31.5 M	\$44.0 M
O&M Cost	\$17.2 M	\$12.5 M
Life Cycle Cost[a]	\$180 M	\$150 M

[a]Based on a 20-year operating life and a discount rate of 10 percent.

Conclusions

In this case study, a treatment train comprising precipitative softening, RO and mechanically-driven evaporation and crystallization was developed to reduce RO reject volume prior to discharge to the existing solar evaporation ponds. However, the operating cost of this train was significantly higher than originally anticipated, driven primarily by costs associated with precipitative softening as well as the need for further brine volume reduction using a crystallizer in order to attain a brine flow that could be adequately evaporated in the ponds. Consequently, a distillation-only treatment train was developed, which was estimated to have a lower 20-year net present value cost, despite a higher capital cost. The distillation-only alternative uses significant power (~10 MW), but the power costs are less than the cost associated with chemical use and solids disposal of the base alternative that uses precipitative softening. In addition, the distillation-only process is less complex, and therefore easier to operate, especially given the absence of lime and soda ash slurry and feed systems and the need for proper chemical dosing and precipitation pH control.

The findings from this case study illustrate that disposal of brackish waste streams, exemplified here by a RO reject produced from purification of a higher salinity municipal water supply, is both challenging and expensive. While the use of membrane-based desalination may appear to be less expensive based on its considerably lower energy use, the need for chemically intensive pretreatment can result in a higher total water cost than thermal based desalination methods, depending on what degree of volume reduction is necessary. As importantly, the use of evaporation ponds for final (zero) liquid discharge must carefully consider a number of factors to ensure that adequate pond capacity is provided, particularly when disposing of hypersaline brines, where evaporation rates are low salt accumulation is significant.

The results presented herein underpin the need for an effective industrial water management approach that moves away from the ‘use it once and discharge’ approach to one that balances high purity water production with sustainable management of saline waste steams with the objective of maximizing recovery and reuse and minimizing impact to the environment.