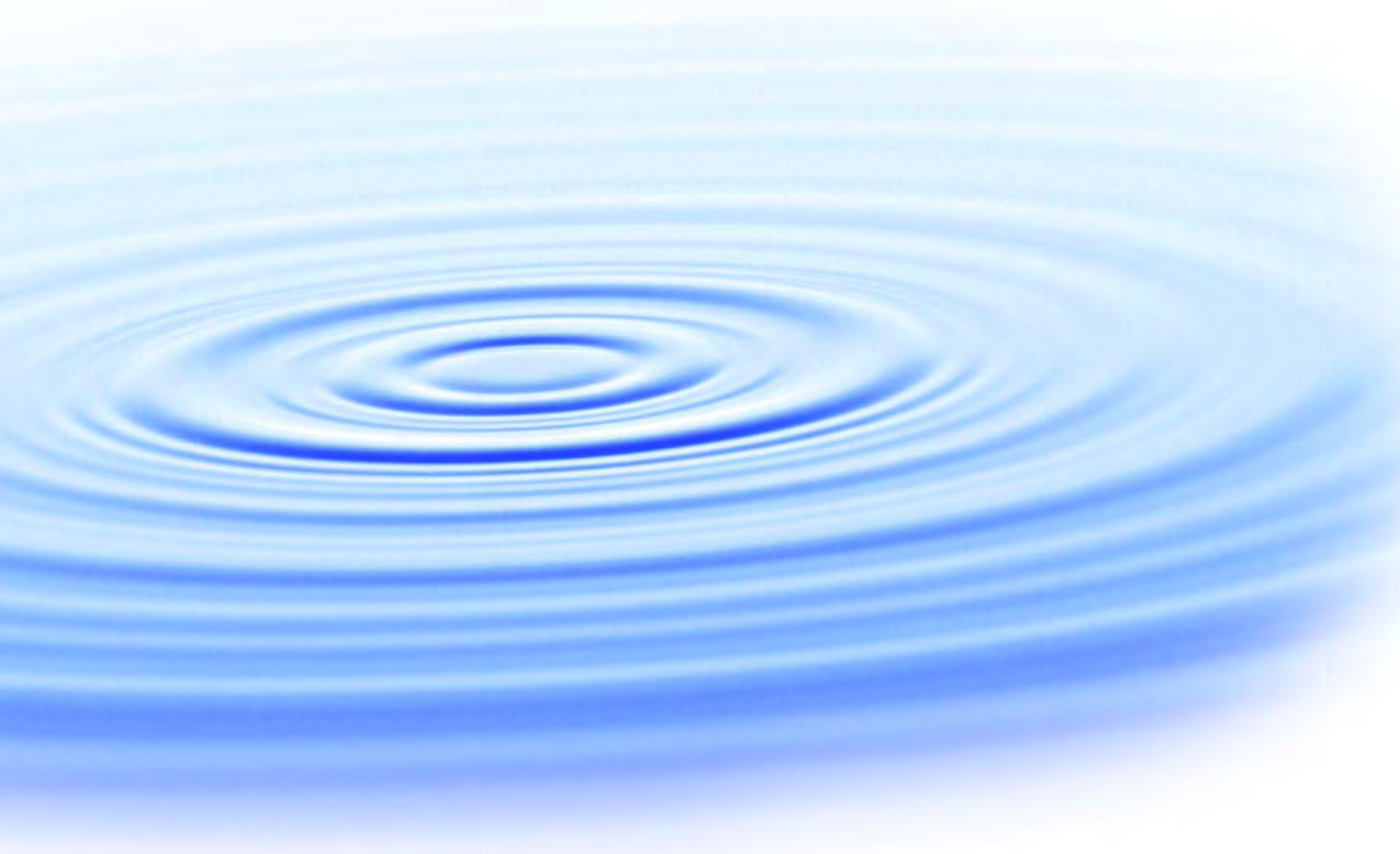
WATEREUSE
FOUNDATION

Survey of High-Recovery and Zero Liquid Discharge Technologies for Water Utilities



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About the WateReuse Foundation

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

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

The Foundation's funding partners include the U.S. Bureau of Reclamation, the California State Water Resources Control Board, the Southwest Florida Water Management District, and the California Department of Water Resources. Funding is also provided by the Foundation's Subscribers, water and wastewater agencies, and other interested organizations. The Foundation also conducts research in cooperation with the Global Water Research Coalition.

Survey of High-Recovery and Zero Liquid Discharge Technologies for Water Utilities

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ACRONYMS

AWWA	American Water Works Association
ED	electrodialysis
EDR	electrodialysis reversal
EPA	Environmental Protection Agency
GMH	gypsum-magnesium hydroxide
gpd	gallons per day
gpm	gallons per minute
IX	ion exchange
HEEP™	High-Efficiency Electro-Pressure (ED system)
HEEPM™	High-Efficiency Electro-Pressure Membrane
M&A	Mickley and Associates
MF	microfiltration
MG	million gallons
MGD	million gallons per day
MH	magnesium hydroxide
MTBE	methyl tert-butyl ether
NF	nanofiltration
NORM	naturally occurring radioactive material
PCC	precipitated calcium carbonate
RCRA	Resource Conservation and Recovery Act
RO	reverse osmosis
TCLP	toxic characteristic leaching procedure
TDS	total dissolved solids
TENORM	technologically enhanced naturally occurring radioactive material
UF	ultrafiltration
WWTP	wastewater treatment plant
ZLD	zero liquid discharge

FOREWORD

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

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

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

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

The Foundation's primary funding partner is the U.S. Bureau of Reclamation. Other funding partners include the California State Water Resources Control Board, the California Department of Water Resources, the Southwest Florida Water Management District, Foundation Subscribers, water and wastewater agencies, and other interested organizations. The Foundation leverages its financial and intellectual capital through these partnerships and funding relationships. The Foundation is also a member of the Global Water Research Coalition.

This publication is the result of a study sponsored by the Foundation and is intended to communicate the results of this research project. The principal goal of this project was to develop realistic performance and costs for high-recovery and zero liquid discharge processing schemes over a range of size, salinity, and water composition variables.

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- Michael Dellavecchia, O’Brien & Gere (ARROW^{TM} technology)
- Brad Culkin and Peter Corboy, New Logic Research, Inc. (VSEP^{TM} technology)
- Ernst Schmidt and Louis Sferrazza, EET Corporation (HEEPM^{TM} technology)

The Project Advisory Committee (listed below) was helpful in reviewing the draft final report and offering good, constructive, and appreciated suggestions.

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EXECUTIVE SUMMARY

Due to their high cost, high-recovery and zero liquid discharge (ZLD) systems used in various industries are not currently used at any municipal sites. Because of increasing challenges to disposal of concentrate, these types of systems will, by necessity, eventually be applied in municipal settings. There is thus a need to develop realistic performance and cost analyses for high-recovery and ZLD processing schemes. The purpose of this research was to evaluate the costs of commercially available volume minimization and ZLD processing schemes over a range of size, salinity, and water composition variables. In addition to identifying and characterizing processes and developing performance and cost projections, regulatory issues associated with use of the technologies have been identified and are discussed.

The report provides insights into the performance and economic aspects of commercial high-recovery and ZLD systems, including specific areas where cost reductions might be possible. The report also provides useful background and reference information for further consideration of such systems.

PROJECT BACKGROUND

Five conventional concentrate disposal methods account for disposal at over 98% of the municipal membrane desalination sites built in the United States (Mickley, 2007a). These options are:

- Surface water discharge
- Discharge to wastewater treatment
- Subsurface injection
- Evaporation ponds
- Land application

These disposal options, however, are not widely available, due to climate, hydrogeology, land, and other requirements. While one of these options may be available in most locations, there are locations where options are either not available or are not cost-effective. This is particularly true in the arid Southwest. New, alternative solutions are required for these areas for anything but small desalination plants.

The consideration of alternative or new concentrate disposal options is also driven by other factors:

- Growing challenges in disposing of concentrate due to:
  - The growing number and size of membrane plants and resultant concentrate volume
  - Increasing regulatory pressures
  - Growing public awareness and concern regarding environmental issues
- Increased valuing of concentrate as “lost water”
- An ultimate goal of sustainable technologies

One consideration is for new uses or reuse of concentrate. This is a worthy consideration that may help a limited number of sites. In general, the disposal of concentrate is site specific, and any use or reuse of concentrate will need to take this variability into account. Most beneficial uses are either unproven or do not provide a final disposal solution (or both). In addition, they are not broadly applicable and, as with traditional disposal options, they are dependent on climate, hydrogeology, nearby industry, regulations, and other factors.

Another consideration is further treatment of concentrate to facilitate disposal, use, or reuse. This includes reducing the volume of concentrate by high-recovery and ZLD processes.

ZLD processes are high-recovery processes where either the final brine is disposed of within the plant boundary (such as in an evaporation pond) or the process produces solids for disposal. Thus, “high-recovery processes” is a more general term that includes ZLD processes. High recovery may offer a solution to the disposal of concentrate, but this is not guaranteed. It is currently a very expensive alternative, but one that with time and reduced costs may be mandated by the factors mentioned above. Except for those schemes that send substantial portions of the brine to large evaporation ponds, high-recovery options typically recover more than 90% of the water from the wastewater stream. The value of this recovered water may somewhat offset the disposal cost.

An analysis of costs for high-recovery processing of a hypothetical 20 million gallons per day (MGD) of concentrate from the Phoenix area was done by the report author in 2003 (Mickley, 2007a). A previous analysis by the U.S. Bureau of Reclamation (USBR, 2000) had considered the options of massive evaporation ponds and a long pipeline to the Sea of Cortez. The 2003 analysis revealed that high-recovery processing could significantly reduce the capital costs but would greatly increase operating costs, such that the annualized costs of the high-recovery processing alternatives were similar to those for the evaporation pond and pipeline alternatives. This simplified the Phoenix analysis which, while suitable and correct at the very general level of analysis conducted, raised many questions about how individual process steps would perform and interact in complex high-recovery processing systems utilizing multiple processing steps.

The present study was undertaken to address such questions and to examine more broadly the sensitivities of process performance and costs to feedwater salinity and the chemical composition of the water.

In the present study, the five conventional ZLD processing schemes, three patented high-recovery technologies, and a patented selective salt recovery technique were evaluated in greater detail.

ZLD STUDY

The individual processing steps (and their abbreviations used in the illustration) of the five conventional ZLD processing schemes for treating wastewater include:

- Reverse osmosis (RO)
- Lime softening (LS)
- Thermal brine concentrator (BC)
- Thermal crystallizer (CRYST)
- Spray dryer (SD) (used only for low-volume flows)

- Evaporation pond (EP)
- Landfill (LF)

Product water is produced by the RO, BC, and CRYST process steps. Processing steps for the wastewater treatment, beginning with concentrate (conc), are as follows:

- Scheme 1A: conc → BC → EP
- Scheme 1B: conc → BC → CRYST → EP
- Scheme 2A: conc → LS → RO → BC → EP
- Scheme 2B: conc → LS → RO → BC → CRYST → EP
- Scheme 3: conc → LS → RO → EP

Solids produced from the lime softening and crystallization steps go to a landfill.

Capital, operating, and annualized costs were developed for these five processing schemes for each of 12 different sets of conditions. The 12 cases included variable process size (1, 10, and 20 MGD of concentrate to be treated), concentrate salinity (4000, 8000, and 12,000 mg/L total dissolved solids), and concentrate composition (seven different compositions). The water qualities correspond to actual concentrates from seven locations. Salinities of the different concentrates were normalized to a salinity of 8000 mg/L. This allowed separation of salinity as a variable from the varied compositions of the concentrates. Five of the 12 cases were used to study the size and salinity variables (at a constant composition), and 7 of the 12 cases were used to study the effects of chemical composition (at a constant size and salinity).

SUMMARY: INDUSTRIAL ZLD SYSTEMS

- Both the salinity and composition of the concentrate (feedwater to the ZLD processing systems) have significant effects on the performance of the individual process steps.

  The effects of salinity and composition variables on RO system performance are well known. The effects of these variables on brine concentrator performance are not available in the literature and therefore require interaction with the manufacturer(s) of the equipment to determine and understand the effects. One of the major effects is how composition determines the limit to which the feedwater can be concentrated. Over the range of compositions considered, the exiting brine concentration from the brine concentrator varied from 167,000–358,000 mg/L. The brine at the upper level is twice as concentrated (reduced volume) as brine at the lower level. This variability in the brine exiting the brine concentrator will affect the flow and composition (and consequently process step size and cost) to the crystallizer or evaporation pond that follows.

  The salinity of the concentrate has a major effect on the performance of the processing steps. For example, for a given limit on the exiting brine concentration from the brine concentrator, a feed of 4000 mg/L salinity would be concentrated twice as much as a feed salinity of 8000 mg/L. Flow to the process step following the brine concentrator would be twice as large in the higher-salinity situation.

- Salinity and composition of the concentrate (feedwater to the ZLD processing systems) have significant effects on capital, operating, and annualized costs.

The effects on cost were due to the previously mentioned effects on individual process step performance, which affect process step size and cost.

- ZLD costs are lowest for low-salinity feedwater and, in general, for low-hardness feedwater (but also are dependent on the alkalinity level).

- System size had a relatively small effect on unit system costs (dollars per MGD).

  This is primarily due to the large plant sizes considered. Concentrate flows of 1, 10, and 20 MGD resulted in the use of multiple equipment modules where savings of economies of scale were minimized. In addition, evaporation ponds have little, if any, economy of scale.

- The sensitivities of performance and cost to salinity and composition suggest that simplified analyses of ZLD processing costs are subject to large errors.

  A detailed analysis based on understanding how each processing step is affected by salinity and composition is required for meaningful cost projections on these complicated, complex processing systems.

- Volume reduction by membrane processing prior to use of a brine concentrator is, in general, the recommended processing scheme.

  Brine concentrators have high capital and high operating costs, the latter due to high energy requirements. Use of volume reduction by a second-stage membrane process prior to the brine concentrator reduces the capital and operating costs. The one exception in the 12 cases was case 6, where high calcium levels and low bicarbonate levels resulted in very high softening costs due to significant lime and soda ash addition. This in turn resulted in less volume reduction at the second RO step and larger volumes going to the brine concentrator and evaporator steps. This exception again points out the need for a detailed process analysis when making decisions, even at a fairly preliminary level, for such complex systems.

- The ultimate disposal challenge is what to do with sodium-dominated brines. The dissolved solid content (total dissolved solids) of the final brine resulting from ZLD (high-recovery) processing is sodium dominated, containing one or more dissolved sodium chloride, sodium sulfate, and sodium carbonate salts.

  Most salts with lower solubilities than sodium salts are precipitated in processing steps such as softening and brine concentration. In the latter case, most lower-solubility salts are allowed to precipitate and become suspended solids in the brine. The variability in the final brine composition after ZLD processing is much less than the variability in composition of the starting concentrates (feed to the ZLD processes) as a result of removal of lower-solubility salts. This final brine is most typically dominated by sodium salts.

SELECTIVE SALT RECOVERY STUDY

A sixth commercial ZLD processing scheme is that of Geo-Processors USA, Inc. This technology consists of one or more series of volume reduction steps followed by a salt recovery step. Less-expensive salts may be added to bring about precipitation and removal of more valuable salts. The technology was evaluated by determining performance and costs upon treatment of several of the 12 cases defined for the ZLD study. Although this technology is commercial, as of October 2007 it has not been piloted or demonstrated in the United States. Due to the nature and geographical origin of the technology (Australia), the analysis required significant input from Geo-Processors. Thus, the performance and costs developed were done on a different and less controlled basis than for the ZLD study. For this reason, the results of the Geo-Processors study presented in Appendix 5 are not directly comparable to those of the ZLD study but are still highly informative. Some of the more important findings of the Geo-Processors study were:

- Technology to accomplish selective salt recovery has been patented, developed, and licensed outside of the United States (Geo-Processors, 2007).
- The processing approach, steps, performance, and costs were defined by analysis of concentrate treatment by this technology.
- The costs analysis demonstrates how salt recovery and marketing can beneficially impact total plant costs.
- In general, for the concentrates evaluated, the most cost-effective salt to recover is precipitated calcium carbonate, a high-value salt used in specialty paper manufacturing.
- The potential benefits are both economic and environmental and also represent a move toward the goal of sustainability.
- Selective salt recovery needs to be pilot tested in the United States to determine its applicability and feasibility for treating concentrates and other waters.

COMMERCIAL HIGH-RECOVERY PROCESS STUDY

Three patented, commercial high-recovery technologies were characterized and evaluated by developing performance and costs for two or three feedwaters each. As with the Geo-Processors selective salt recovery technology, this information was developed with significant input from the process manufacturers. The technologies are considered emerging in the sense of not having been applied, other than piloting studies, in the municipal setting. The technologies are:

- HEEP^{TM} by EET Corporation
- VSEP^{TM} by New Logic Research
- ARROW^{TM} by O’Brien & Gere

As with the Geo-Processors technology, the results cannot be compared directly with the results of the industrial ZLD process study. The results are informative and helpful in describing the technologies and their potential to be applied in municipal settings. Some of the more important findings from the study of these three technologies were:

- There are patented, commercial high-recovery systems that offer advantages over the conventional high-recovery approach of interstage treatment between two RO stages.
- These technologies may each have applications in municipal settings.

- A preliminary cost evaluation suggests that these technologies are all more cost-effective than direct use of brine concentrators, consistent with the results of the ZLD process scheme analysis.

REGULATORY STUDY

The primary regulatory issue is whether further concentration of concentrate will render the resulting brine or solid to be hazardous or radioactive in nature due to contaminants present in the original concentrate. The regulatory frameworks for hazardous wastes and radioactive wastes are reviewed, and means of ascertaining the nature of brine and solids are discussed. The key issues associated with this study were:

- The hazardous question can be addressed early in the planning process by (1) obtaining a broad characterization of the concentrate (or raw water) constituents that includes those of high concern, (2) estimating the resulting brine or solid concentration of the constituents, and (3) comparing the estimated concentrations with hazardous waste standards.
- The radionuclide question can be addressed in the same manner but is complicated by the fact that the standards are to be developed by individual states and in many cases do not yet exist. For this situation, state regulatory agencies handle such questions on a case-by-case basis.
- Because levels of contaminants are greater in the concentrate, a detailed water quality analysis of the concentrate may be better suited to determine the levels of concern that contaminants may pose, rather than a detailed analysis of the raw water. It is possible, for example, for a level of contaminant to be undetectable at feedwater concentrations but present at problematic levels in the concentrate or subsequent brine or solids.

SUMMARY

- The only distinction between ZLD processes and high-recovery processes, both of which produce brine, is whether the brine will be disposed of within the plant boundary.
- High-recovery and ZLD processing schemes are technically feasible but, in general, not economically feasible for municipal applications.
- Economic feasibility in the municipal industry requires cost reductions.
- High capital costs are associated with evaporative processing steps and final disposal steps, such as evaporation ponds and landfill.
- High operating costs are associated with energy and chemical needs.
- Cost-effective high-recovery and ZLD treatment steps (as opposed to final disposal steps) do not guarantee cost-effective final waste disposal steps.
- Cost reductions are needed for both the treatment steps (volume reduction steps) and the final disposal steps.
- Promising areas of consideration for cost reduction include:
    - Second-stage membrane processing before or in place of evaporative volume reduction steps
    - New Logic's VSEP^{TM} technology for reducing chemical costs
    - Geo-Processors' selective salt recovery processing for reducing disposal costs (and reducing environmental footprints, including the CO_2 footprint)

RECOMMENDATIONS

- Cost reductions are necessary for application of most high-recovery and ZLD processing schemes. Research should be conducted with this goal in mind. In particular, the goals should include a reduction in capital cost and a reduction of operating costs due to energy and chemical requirements.
- It is important that the newer commercial technologies (Geo-Processors' SAL-PROC, EET Corporation's HEEPM, New Logic's VSEP, and O'Brien & Gere's ARROW) be piloted for municipal applications and benchmarked against the more traditional high-recovery and ZLD approaches.
- Due to the significant effects of salinity and composition on system performance and cost, simplified analyses of high-recovery and ZLD systems costs may be subject to considerable error. Future analyses need to include considerations of the effect of water quality on each processing step to ensure good cost projections.
- Detailed water quality analyses need to be done at the concentrate level to determine whether contaminants present at low levels in feedwater (or perhaps undetectable there) will result in brine or solids being hazardous or containing problematic levels of radionuclides.
-

There is a need to develop a knowledge base for high-recovery and ZLD processing and to acknowledge the real cost and environmental consequences of large-scale concentrate disposal for inland desalination plants.

CHAPTER 1

INTRODUCTION

1.1 REASONS FOR THE STUDY

Although approximately 120 commercial zero liquid discharge (ZLD) systems are in use to treat wastewaters in various industries (Mickley, 2007a), at present there are no high-recovery or ZLD desalination processing systems operating at municipal sites. This is due to the high capital and operating costs associated with these processing systems.

Increasing challenges to concentrate management (see Chapter 2) have resulted in high-recovery and ZLD systems recently being considered for municipal applications (Mickley, 2007b). Indeed, in various locations (such as Las Vegas, Phoenix, and Denver), conventional concentrate management options are not cost-effective and thus desalination plants are not being built. While costs remain prohibitive for most municipal situations, it is now timely to identify and review commercial high-recovery and ZLD processing schemes and to characterize their performance and costs.

1.2 PROJECT OBJECTIVE, GOALS, AND KEY QUESTIONS

The objective of this study was to identify, characterize, and evaluate commercially available high-recovery and ZLD processing schemes. The understanding gained from this effort will help to identify areas where cost reductions may be possible and determine the future applicability of these systems to municipalities.

Specifically, the goals of the project were to:

- More fully understand these systems, including:
    - their performance and costs
    - the dependence of performance and costs on size and water quality (salinity and composition)
    - performance and costs of individual processing steps that make up the processing schemes
- More fully understand the potential future applicability of these systems to municipalities
- Identify areas where cost reductions may be possible
- Identify promising commercial technologies and approaches that may be used in future high-recovery and ZLD systems

Key questions to be answered in the study included:

- How can clarity be brought to definitions of high recovery and ZLD?
- What processing schemes are used at the approximately 120 commercial ZLD industrial facilities in the United States?
- How are the performance characteristics of individual processing steps and the composite processing schemes of these commercial ZLD systems affected by variations in water quality (salinity and composition)?
- How do the costs of these commercial ZLD systems vary with size and water quality (salinity and composition)?

- What are the contributions of the individual process steps to the total costs for commercial ZLD systems?
- Do any of the commercial processing schemes have consistently lower costs?
- What are the major cost factors, regarding both capital and operating costs, for these commercial ZLD processing schemes?
- What are the different technical approaches to achieving high recovery?
- What commercial high-recovery technologies, not yet widely applied, exist and what are their performance and cost characteristics?
- What new (to concentrate management) regulatory issues are raised by high-recovery processing?
- What are the limiting cost factors that lead to high capital and operating costs for the high-recovery and ZLD processing systems?

1.3 PROJECT TASKS

Individual project tasks included:

- Characterizing the commercially available volume minimization and ZLD technologies
- Evaluating the economics of these technologies
- Inventorying and preliminarily evaluating promising technologies that may be part of future commercial volume minimization and ZLD processing schemes
- Summarizing regulatory issues associated with volume minimization and ZLD technologies
- Identifying areas of needed research

The primary purpose of the study was to evaluate the costs of commercially available volume minimization and ZLD processing schemes over a range of size, salinity, and water composition variables. In addition to characterizing these processes and their capital and operating costs, regulatory issues were identified and will be discussed.

1.4 PROJECT APPROACH TO DEVELOP PERFORMANCE AND COST PROJECTIONS

There is confusion within the municipal community in both the literature and general usage with regards to definitions of high-recovery and ZLD systems. Distinctions between high-recovery and ZLD processes are discussed in Chapter 3 and are then used to categorize the processing schemes that are considered for analyses. Although ZLD processing schemes are high-recovery processes, the term ZLD as used in other industries refers to processes where no liquid leaves the plant boundary (the original definition of ZLD). Many ZLD processing schemes include evaporation ponds as a final processing step, where the ponds are within the plant boundary. High recovery in the membrane desalination industry is taken to mean a recovery above that usually possible with a single conventional membrane stage. Since recovery is dependent on the water quality (salinity and composition), the numerical definition of high recovery is arbitrary. In this report, “high recovery” is taken to mean a recovery of 92% or higher.

For project evaluation, high-recovery and ZLD processing schemes were divided into two groups:

- Existing commercial ZLD processing schemes found in other industries
- Emerging commercial high-recovery and ZLD processing schemes used in other industries

1.4.1 Existing Processing Schemes (Commercial ZLD)

The approximately 120 commercial industrial wastewater ZLD systems in the United States involve processing schemes that include some or all of the following process steps:

- Reverse osmosis (RO)
- Lime softening
- Thermal brine concentrator
- Thermal crystallizer
- Evaporation pond
- Landfill

Determination of the effects of water quality (salinity and composition) on process performance and costs means determination of these effects on each processing step of these multistep processing schemes. This in turn dictates a certain level of process step understanding, characterization, and evaluation. This level of description is difficult for the thermal processes, where the technologies are somewhat proprietary and where commercial installations are in private industries that do not openly share performance and cost information.

Due to the complexity of chemistry-dependent performance, it was necessary to obtain performance estimates for the thermal process steps directly from manufacturers. For this study, information was obtained from GE-Ionics-RCC, the company involved in a majority of the industrial ZLD sites in the United States. The information obtained was used to define feed conditions (flow and water quality) for the process step following the thermal step. Costs for thermal processes were developed using in-house updated cost curves.

Since the Principal Investigator for this study conducted the detailed evaluation process (with the exception of defining the performance of the thermal process steps), the performance and costs for the existing commercial processing schemes were developed on a consistent basis, allowing comparisons among the different processing schemes.

1.4.2 Geo-Processors Selective Salt Recovery Technology

Another commercial ZLD technology exists that has not yet been piloted or demonstrated in the United States. The Geo-Processors technology is largely proprietary and involves processing steps and details not easily evaluated by outside parties. Geo-Processors was asked to provide performance and cost information in response to various water quality cases. The information on the different cases is self-consistent, as it was developed by Geo-Processors. However, the cost information cannot be compared directly with the costs of the other existing commercial processing schemes discussed above.

The information, however, is of value in providing a detailed description of the nature of Geo-Processors' processing sequence, of how the technology performance and cost varies with water quality, and of important cost factors. The Geo-Processors technology is discussed in Appendix 5.

1.4.3 Emerging Processing Schemes (Patented and Commercial High Recovery)

As a prelude to the identification and analysis of these technologies, it was useful to examine the different technical approaches that could be used to achieve high recovery. These are discussed in Chapter 3.

The criteria for selecting processing schemes were:

- The processing schemes were different from existing ones previously discussed.
- The technologies were patented.
- The technologies had been commercially used or were being piloted or demonstrated in industry (nonmunicipal settings).

Since the processing schemes were patented and proprietary (to varying degrees), it was not possible to develop an independent evaluation of performance and cost. Technology owners were asked to provide performance and cost estimates for various cases, defined by size and water quality. The information sought from the manufacturers was considerable in amount and detail. While several iterations were involved in asking for and receiving information, the information obtained does not provide a consistent basis for comparison. The inconsistencies are also in part due to the different standard procedures for developing and reporting costs that were used by the process manufacturers. Thus, the cost information obtained is not on a standard and consistent basis.

As a result, the performance and cost estimates for the different emerging processing schemes cannot be directly compared with those for other emerging schemes, nor can they be directly compared with the cost evaluations for existing processing schemes. The information, however, is useful in providing detailed descriptions of the various processes, in reflecting how the processing schemes perform for different waters, and in identifying important cost factors.

1.5 OUTLINE OF REPORT

Following this introductory chapter, Chapter 2 provides background information on municipal desalination concentrate management and on high-recovery and ZLD processing systems. Chapter 3 discusses volume minimization and ZLD technologies. Commercially available technologies for further evaluation are identified and characterized in terms of applications, performance capabilities, and commercial use.

In Chapter 4, the methodology used to evaluate the economics of the existing commercial ZLD processing schemes is defined. Size, salinity, and composition variables are discussed, and the specific sets of variables used to evaluate the processing schemes are presented. Models used in sizing, performance prediction, and costing are documented.

Chapter 5 presents the results of the evaluation of the existing commercial ZLD processing schemes that have been applied in the United States. Performance and economic results are provided for the five processing schemes and for 12 different cases in which process size, salinity, and composition varied. Appendix 5 characterizes and discusses Geo-Processors' selective salt recovery process, which was developed, patented, and licensed outside of the

United States. This technology has potential to bring concentrate management into a more sustainable mode with the recovery of commercial-grade salts.

Chapter 6 presents economic evaluation results for three emerging high-recovery processes. Results are discussed in terms of sensitivity to size, salinity, and composition variables. Insights gained from the analysis are discussed.

Chapter 7 discusses regulatory issues associated with the highly concentrated residuals produced in volume minimization and ZLD processing schemes. This includes both brines and solids.

In Chapter 8, design, cost, and operating considerations for brine concentrator ZLD systems are discussed from the perspective of field experience and resulting lessons learned from the power industry.

Chapter 9 presents the findings and conclusions stemming from the study results. Also included is a section on recommendations for future research.

Supportive information is provided in several appendices. Appendix 1 provides process flow diagrams for the commercial ZLD processing schemes in use in the United States. Appendix 2 contains the master cost tables resulting from analysis and evaluation of these ZLD processing schemes. Appendix 3 contains bar charts showing individual process step contributions to capital and operating costs for these ZLD processes. Appendix 4 provides a list and discussion of detailed observations resulting from the process scheme performance and economic evaluations. Finally, Appendix 5 provides a characterization and evaluation of the Geo-Processors selective salt recovery technology.

CHAPTER 2

BACKGROUND

2.1 INTRODUCTION

The purpose of this study was to evaluate the costs of commercially available volume minimization and ZLD processing schemes over a range of size and water qualities (salinity and composition). The focus is restricted to technologies used in wastewater treatment, thus eliminating those technologies used in process water treatment. New technologies were also identified that might become part of more cost-effective processing sequences. In addition to characterizing the processes and developing capital and operating costs, regulatory issues were identified and discussed.

2.2 RELEVANT ISSUES

In the United States, membrane technology is the technology of choice for addressing the growing need to treat lower-quality water resources. Consequently, the number of desalination plants has been steadily increasing. This growth and the concentrate disposal practices from these advanced water treatment processes have been well-documented (Mickley, 2001a, 2006a; Mickley et al., 1993). There were, as of the end of 2004, over 500 municipal water and wastewater treatment plants of size 25,000 gallons per day (gpd) or larger utilizing membrane technology (desalination and low pressure) in the United States (Mickley, 2006a). Of the more than 260 desalination plants, over 95% are inland plants.

While concentrate disposal options are site specific, until the late 1990s one or more conventional options (surface water discharge, disposal to wastewater treatment lines, deep-well injection, land application, evaporation ponds, or recycling) were available to nearly every site or location. Historically, about 75% of concentrates are disposed to surface water or to the front end of a wastewater treatment plant. While disposal to surface water has been used with concentrates of all sizes, disposal to the wastewater is not typically used for larger volume concentrates, where concerns have been raised about the effects of the concentrate total dissolved solids (TDS) on process microorganisms and on the effluent TDS.

The number and size of both desalting and low-pressure membrane plants have been increasing (Mickley, 2001a, 2006a). In addition, discharge and general disposal regulations have become more stringent, and public awareness of environmental issues and resultant concerns have increased. All these factors have resulted in concentrate disposal being a growing challenge, regardless of location. Concentrate disposal is particularly problematic in the inland arid southwestern portion of the United States, where both surface water disposal and disposal to publicly owned treatment works are limited. In addition, the application of the other three traditional concentrate disposal options has been and will continue to be limited. Deep-well injection is unproven or prohibited in these arid states, and the costs of determining feasibility and implementing the wells are huge. Land application is limited by the lack of the large quantities of water needed to dilute the concentrate and make it more compatible with the groundwater. Finally, although the climate is ideal for evaporation ponds, the high cost of ponds makes this solution prohibitive for large volumes of concentrate. At present there are several inland sites in the arid Southwest where membrane desalting plants could provide needed drinking water if concentrate disposal solutions were available.

The immediate challenge is to find a cost-effective concentrate management option that will support the feasibility of inland desalination plants. The longer-term and ultimate challenge is to find a concentrate management solution that does not simply shift a water quality problem from one area (or one user) to another.

The prolonged recent droughts in the western United States have generated awareness and interest in recovering and/or recycling all potential sources of fresh water, including membrane concentrate. A growing issue raised by some water resource management groups is how to value resource water lost through concentrate disposal. The issue is one of growing focus but as yet of limited study and clarity. In several recent feasibility evaluations, the value of lost water has been taken into consideration when comparing disposal options.

Another growing issue but one that is still in its infancy is the goal of sustainable technologies. While in many situations this is not possible, it is a desirable and ultimately necessary direction. Disposal of concentrate to surface water and groundwater results in salt load buildup. Eventually this can reach a level that will limit additional dischargers. Such a situation exists in the Denver, CO, area, where complex modeling of entire stream segments has been used to determine whether discharge is possible (i.e., whether or not receiving water standards can be met) and what effect this will have on disposal permits of current dischargers. Even though the discharged concentration of a given constituent may be below the receiving water limit, it may be above the ambient level and thus increase the receiving water concentration. This can limit the possibilities for future dischargers and increases in discharge volumes by existing dischargers. The situation is one of salt loading of a constituent. Discharges of greater salinity than the receiving water result in total salt loading.

The consideration of alternative or new concentrate disposal options is thus driven by several factors:

- Growing challenges of disposing of concentrate, including:
  - the growing number and size of membrane plants and resultant concentrate volumes
  - increasing regulatory pressures
  - growing public awareness and concern for environmental issues
- Increased valuing of “lost water”
- The ultimate goal of sustainable technologies

One consideration is for new uses or reuse of concentrate, which may help a limited number of sites. In general, concentrate is site specific, and any general use or reuse of concentrate will need to take this variability into account. Most beneficial uses are either unproven or do not provide a final disposal solution (or both). In addition, they are not broadly applicable and, as with traditional disposal options, they are dependent on climate, hydrogeology, nearby industry, regulations, and other factors.

Another consideration is further treatment of the concentrate to facilitate disposal, use, or reuse. This includes reducing the volume of concentrate by high-recovery and ZLD processing.

High-recovery and ZLD processing systems (definitions for these are discussed further in Section 3.1) offer an alternative to the disposal of concentrate to surface and groundwater. It is currently a very expensive alternative but one that with time and reduced costs may be mandated by the factors mentioned above. Except for schemes based on large evaporation ponds, typically these processes need to recover most of the water from the wastewater stream. The value of this recovered water can help offset the disposal cost.

Since ZLD processing systems are high-recovery systems, the discussion in the rest of this chapter uses the term “high recovery” to mean both high-recovery and ZLD systems.

2.3 FURTHER TREATMENT OF CONCENTRATE

A recent research project provided some foundation for the present effort (Mickley, 2007a). More specifically, a soon-to-be-completed U.S. Bureau of Reclamation project addresses this topic in a broad fashion. That project, which began in 2002, is entitled “Treatment of Concentrate.” It looks at two general topics: (1) what treatments may be required to remove a specific contaminant from a concentrate to enable disposal via conventional means, and (2) what are the means for and issues involved with further concentration of concentrate. Some of the early findings of the U.S. Bureau of Reclamation project are discussed next.

2.3.1 Increasing Recovery

Increasing recovery beyond that typically achievable in a single-pass RO process can help or hinder disposal, depending on the particular disposal option. Increases in recovery reduce concentrate volume, increase its salinity, and typically make the concentrate less compatible (in terms of salinity) with the receiving water.

Increasing recovery may help other disposal options, such as evaporation ponds (a smaller volume would require evaporation), deep-well injection (disposal of a smaller volume), and evaporative-based ZLD (resulting in a smaller volume going to expensive thermal evaporative systems).

Unless the disposal options of evaporation ponds or deep-well injection are available, there is usually nothing gained by increasing recovery unless the resulting recovery is very high. One exception to this involved a very long pipeline to a point of discharge into the ocean. In that case, volume reduction reduced the pumping costs by reducing the volume required for pumping to the ocean. However, this is not a very likely scenario for most installations, and it is unlikely the reduced energy costs associated with the pumping would be sufficient to justify the additional treatment costs.

2.3.2 Reducing Size and Energy Costs Associated with Brine Concentrators

Evaporative-based conventional ZLD technologies are energy-intensive, which results in high annualized costs. These costs can be offset somewhat by increasing membrane system recovery prior to these thermal evaporative systems.

The various means of increasing membrane system recovery are mostly variants of extensive pretreatment of the feed to a two-stage membrane system or interstage treatment prior to the second membrane stage.

Such increased treatment has its costs, and for situations of high-hardness waters, such treatment can result in high chemical costs and high solids disposal costs. In a conventional evaporative-based ZLD scheme, the increased recovery and reduced volume of the new concentrate (brine) results in a much smaller thermal, energy-intensive, evaporative system. With the smaller size comes a much-reduced energy cost. The significant chemical and solids disposal costs (associated with the pretreatment), however, substantially replace the saved energy cost, frequently resulting in an equivalently high operating cost. It is important to note that the increased labor typically associated with increased treatment can be a significant portion of the higher operating cost. For instance, an additional full-time operator to support additional treatment can translate into as many as four additional personnel on the payroll.

There are limited uses for mixed solids that may result from a ZLD process. In some cases, where the feedwater contains predominantly one salt (NaCl dominated, $NaHCO_3$ dominated, Na_2SO_4 dominated, etc.), the final salt may be of sufficient purity for some uses; however, this is not typical.

The U.S. Bureau of Reclamation study (Mickley, 2007a) considered evaporative-based ZLD disposal options based on a hypothetical situation in the Phoenix area. Various regional brackish RO sites produced a total of 20 million gal/day (MGD) of concentrate of a specific water quality. This basis was used in a 2000 U.S. Bureau of Reclamation report (USBR, 2000) that considered two disposal options: transport of the concentrate via a long distance pipeline to the Sea of Cortez and a multi-square-mile area system of evaporation ponds. The 2007 study looked at two additional scenarios:

- Treating the concentrate with a thermal brine concentrator followed by evaporation ponds
- Sending the concentrate to a second-stage RO system whose concentrate then went to a brine concentrator followed by evaporation ponds

Capital, operating, and annualized costs are given for these various scenarios in Table 2.1 and Table 2.2.

The costs in Table 2.1 were figured at $0.05/kWh, sludge disposal at $30/ton, and an annualized cost over 40 years and 7.125% interest (the basis used in the 2000 U.S. Bureau of Reclamation report). The capital cost savings of the high-recovery technology schemes (options 3 and 4) are evident in the capital cost row. The much higher operating costs are evident in the operating cost row. Table 2.2. shows details of the operating costs for options 3 and 4.

The inclusion of high-recovery RO dramatically reduces the size of the thermal brine concentrator that follows, and in so doing reduces the energy costs significantly. The decreased energy costs, however, are made up for by the increased costs of chemicals and sludge disposal. These results reflect that the reduction of high-recovery costs is not simply a matter of reducing energy consumption.

In the above analysis, the thermal options were not credited with the water recovered.

Table 2.1. Costs Associated with Different Disposal Options (20 MGD of Concentrate)

Parameter	Option 1 Pipeline	Option 2 Evaporation Ponds	Option 3 Thermal Evaporation + Evaporation Ponds	Option 4 HERO + Thermal Evaporation + Evaporation Ponds
Capital Cost, M$	310	410	136	76
Operating Cost, M$/year	0.8	1.6	32.9	28.7
Annualized Cost, M$/year	24	33	43	35
Water Lost, MGD	20	20	0.8	0.8

Table 2.2. Operating Costs for Different Disposal Options (20 MGD of Concentrate)

	Operating Cost, M$			
Parameter	Option 1 Pipeline	Option 2 Evaporation Ponds	Option 3 Thermal Evaporation + Evaporation Ponds	Option 4 HERO + Thermal Evaporation + Evaporation Ponds
Labor	---	---	1.1	3.1
Energy	0.8	---	31	3.9
Chemicals	---	---	---	6.2
Sludge Disposal	---	---	---	14.7
Evaporation Pond	---	1.6	0.8	0.8
Total	0.8	1.6	32.9	28.7

2.3.3 Need for Further Study

This background provided the impetus for the present study. Early work in the U.S. Bureau of Reclamation study (Mickley, 2007a) demonstrated the general issues, costs, and potential of high-recovery and ZLD technologies. The potential includes:

- Achieving a sustainable solution
- Enabling the building of desalting plants in areas where conventional disposal options are not possible
- Reducing lost water

The use of high-recovery and ZLD processing in municipal sectors is limited primarily by its high cost. Since there have been no prior applications within the municipal sector, there are also unknowns concerning the treatment of the varying concentrate water qualities, the different natures of the solids produced, and the environmental issues associated with application of these processing schemes.

Since the early findings of the U.S. Bureau of Reclamation study, there have been significant interest levels and research initiated in volume minimization (high-recovery processing). These have been mainly membrane-based processing systems. Distinctions between high-recovery processing and ZLD processing have been blurred by the lack of consistent working definitions, resulting in an unclear framework from which to evaluate various studies.

The findings of the U.S. Bureau of Reclamation study (Mickley, 2007a) were of a preliminary nature, laying the basis for and justifying the more-detailed and in-depth present study. There is a need to further characterize and to define the potential of high-recovery and ZLD processing systems. This includes documenting costs, cost factors, and cost issues that are critical in reducing the costs of high-recovery processing schemes and identifying promising directions to achieve cost-effectiveness.

2.3.4 Selective Salt Recovery

A variation of high-recovery and ZLD processing, not currently applied in the United States, is selective salt recovery. Selective salt recovery and subsequent use (sale) of the salt(s) may offset the high operating costs typically associated with high-recovery situations. Reducing the amount of disposal reduces the environmental impact. Further, removal of carbonate species can reduce the CO_2 footprint of a desalination process. Selective salt recovery represents a step toward greater sustainability, and it may be an important new direction for concentrate management.

As a result of the above and other studies, the U.S. Bureau of Reclamation project (Mickley, 2007a) looked at the selective removal of individual salts from concentrate. Based on salt solubility and the ionic composition of a concentrate, a general sequence of salt precipitation may be inferred. During investigation of the possibility and issues of selective salt recovery, we became aware of an Australian company, Geo-Processors Pty. Ltd. (now a U.S. company, Geo-Processors USA, Inc.) that commercially recovers salts from virtually any effluent, including membrane concentrates and seawater (www.geo-processors.com). Subsequent communication with Geo-Processors provided information that led to a preliminary evaluation of their technology and its applicability to treatment of membrane concentrate. Examples of commercial and pilot projects provided by Geo-Processors showed a variety of applications, with some having a net operating income due to the sale of salts produced.

Table 2.3 shows applications for various salts obtainable from effluents or concentrates via selective salt removal processes. Values for these salts vary widely, depending on the grade and volume. A form of calcium carbonate called precipitated calcium carbonate (PCC) can be valued as high as $350/ton. Commercial forms of sodium chloride typically have a value around $15/ton.

2.3.5 Other High-Recovery Technologies

As well as the need to define and characterize commercially available and widely used ZLD systems, it is important to look at other high-recovery commercial technologies that are relatively new in the treatment of industrial wastewaters. Indeed, the reason some of the new technologies were developed was to overcome limitations of older technologies.

Table 2.3. Major Salts and Application Areas

Chemical Formula	Name	Application Areas
$CaCO_3$	Calcium carbonate	Paper coating pigment, filler for plastics and rubbers, special inks, paints, and sealants
$CaSO_4 \cdot 2H_2O$	Gypsum	Remediation of sodic soils, manufacture of building products
$CaSO_4 \cdot 2H_2O + Mg(OH)_2$ Slurry	Gypsum magnesium hydroxide	Wastewater treatment, pH buffering, soil conditioner for sodic soil
$CaCl_2$ (liquor)	Calcium chloride	Dust suppression, road base stabilization, sodic soil remediation, cement and concrete stabilizer, construction industry
$KNaSO_4$	Glacerite	Potassium fertilizer
$Mg(OH)_2$ slurry	Magnesium hydroxide	Water and wastewater treatment, environmental, animal stock feed, feedstock for magnesium metal production, fire retardants and refractories, acid neutralization
$xMgCO_3 \cdot yMg(OH)_2 \cdot zH_2O$	Magnesium carbonate light	Fire retardant, feedstock for magnesium metal production, filler for paper manufacturing, rubber, and paint
NaOH	Caustic soda	Many applications industrially, including basic feedstock for chemical processes, pH adjustment
NaCl	Halite	Food and industrial processes, chloralkali production, bulk salt supply
Na_2CO_3	Soda ash	Water treatment, chemical industry
Na_2SO_4	Thenardite	Surfactants manufacture, detergent manufacture, glass manufacture, remediation of calcareous soil
NaOCl	Sodium hypochlorite	Disinfection, chemical industries, pool chlorine
$NaClO_4$	Sodium chlorate	Paper bleaching, chemical industries

2.4 SUMMARY

While there are currently no high-recovery or ZLD systems operating at municipal desalination plants because of the high costs involved with these systems, increased challenges in finding cost-effective concentrate management solutions have led to the consideration of high-recovery and ZLD systems. Of the few available studies, most have focused on a site-specific application and thus have not considered the effects of plant size and water quality on process performance and costs. The present study is the first to look at the full spectrum of commercial high-recovery and ZLD technologies and how performance and costs of these technologies depend on plant size and water quality (salinity and composition).

CHAPTER 3

VOLUME MINIMIZATION AND ZLD TECHNOLOGIES

3.1 INTRODUCTION

In discussing the high-recovery and ZLD processing systems used in industry, it is important to keep in mind that due to their high cost they are not in use in any municipal systems.

ZLD processing and the associated terminology evolved from the 1970s mandate for ZLD processing at power plants near the Colorado River, which was necessary in order to reduce the salinity of the river, which eventually crosses into Mexico. The term ZLD means that no effluent (liquid) leaves the ground-level plant boundary. All liquid (water) is either recovered and reused or it is evaporated from evaporation ponds located within the plant boundary. While done for environmental reasons, it also resulted in simpler and quicker permitting, an important time-saving benefit to the power industry.

All of the early ZLD systems included evaporative processing steps, such as a brine concentrator and in some instances a crystallizer.

Several definitions of ZLD have been used within the municipal desalination community, largely due to the consideration only recently of ZLD processing in the industry and the nonfamiliarity with the technologies. These definitions include the following aspects:

1. No water leaving the plant boundary
2. Processing that involves a brine concentrator and/or crystallizer
3. Processing of wastewater all the way to solids

More recently, ZLD systems used in the power and other industries have used membrane processing as a means either to reduce the feed volume going to the evaporative processes or to entirely replace the evaporative-based ZLD processing (Aquatech, 2007). The systems that include membrane processing meet definition 1 above.

The author suggests that this first definition is the most useful to avoid confusion when discussing volume minimization (high-recovery processing) and ZLD processing.

This definition leads to the following points:

- High-recovery processing may be ZLD processing *if* no water leaves the plant boundary.
- ZLD leads to high recovery, *but* high recovery does not imply that ZLD is achieved.

It is further helpful to discuss ZLD processing as either evaporative-based, membrane-based, or a combination of the two.

To date, the various high-recovery and ZLD processing schemes have all been expensive and have not yet been applied in a municipal setting.

Figure 3.1 depicts a range of recoveries, starting with more conventional first-stage RO recoveries (on the left) and proceeding (to the right) with increasing recovery efficiencies. At the extreme right are both the high-recovery and ZLD situations.

First-stage RO processing recoveries are typically limited to the range of 60–85% (concentrate from 15–40%). They produce a concentrate of from 15–40% of the feed volume. Volume minimization in municipal desalination means reducing the concentrate volume relative to that produced by this single-pass, first-stage membrane processing. In many situations, if recovery were increased from 70–85%, it is arguable that little would be gained from the perspective of concentrate disposal. While the volume of concentrate would be reduced by a factor of 2 (from 30% of the feed volume to 15%), the concentrate salinity would increase by a factor of 2. This increase in salinity may result in a greater incompatibility between the concentrate and any potential receiving water, making it more difficult to dispose of the concentrate by these options. Disposal by the other two conventional disposal options, deep-well injection and evaporation ponds, is frequently not possible at the location in question. Unless the reduced-volume brine is further concentrated so as to produce a smaller-volume brine or solids (or a sludge), there may be no options.

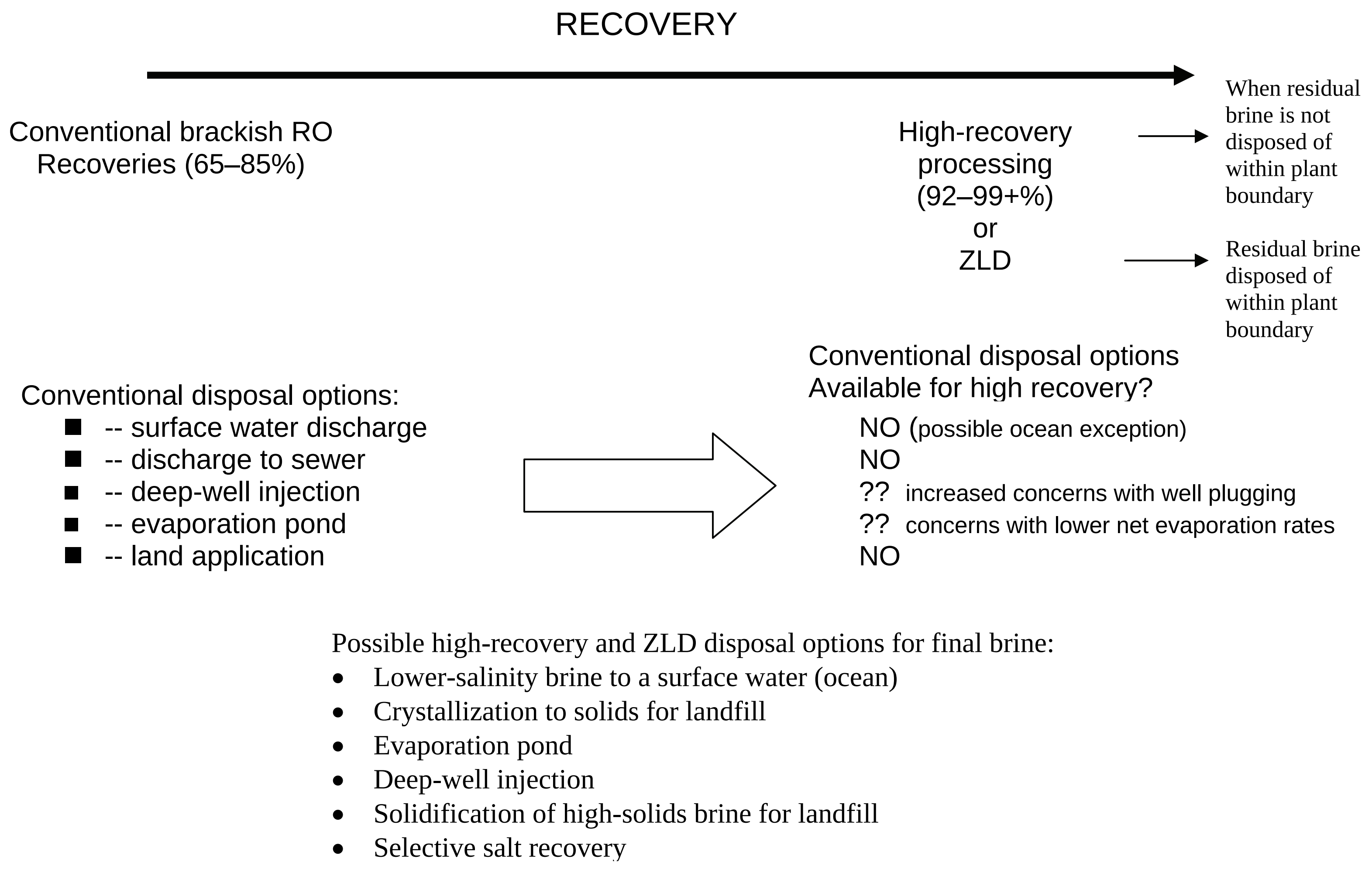
From this perspective, we have designated recoveries of ≥92% as high recovery. Volume minimization technologies of interest in this report are those capable of achieving recoveries in this range. The available disposal options for brine or solids resulting from the high-recovery and ZLD processing schemes are limited and typically have an end step of disposing nonleachable salts and/or solids to a landfill.

3.2 REVIEW OF CONVENTIONAL ZLD PROCESSING SCHEMES

There are approximately 120 industrial wastewater (nonmunicipal) ZLD facilities in the United States. The industrial ZLD technologies are in reality a sequence of individual processing steps or individual technologies. Technologies that are routinely used in various ZLD processing schemes include:

- Thermal (evaporative) brine concentrators
- Thermal (evaporative) crystallizers
- RO
- Evaporation ponds
- Spray dryers

All of these technologies have been used in ZLD processing schemes in the United States. In addition to these, one might add Geo-Processors' sequential salt precipitation technology. This technology has been commercially used in Australia and other countries and has been identified as having the potential to impact ZLD costs in the United States (Mickley, 2006b, 2007a).



RECOVERY

Conventional brackish RO
Recoveries (65–85%)

High-recovery
processing
(92–99+%)
or
ZLD

When residual brine is not disposed of within plant boundary

Residual brine disposed of within plant boundary

Conventional disposal options:

- ■ -- surface water discharge
- ■ -- discharge to sewer
- ■ -- deep-well injection
- ■ -- evaporation pond
- ■ -- land application

Conventional disposal options
Available for high recovery?

NO (possible ocean exception)
NO
?? increased concerns with well plugging
?? concerns with lower net evaporation rates
NO

Possible high-recovery and ZLD disposal options for final brine:

- Lower-salinity brine to a surface water (ocean)
- Crystallization to solids for landfill
- Evaporation pond
- Deep-well injection
- Solidification of high-solids brine for landfill
- Selective salt recovery

Figure 3.1. Range of recoveries and possible disposal options.

Conceptually, the ZLD processing schemes involve a series of process steps, primarily of two types:

1. *Pretreatment or treatment steps* to permit more efficient subsequent desalination of the stream or to selectively remove specific salts; these steps produce either waste solids or product solids and slurries
2. *Desalination steps* to recover water

The treatment and desalination steps may be linked in a number of ways, ranging from simple to complex. The most typical processing schemes are depicted in Figure 3.2. These processing schemes represent the general treatment approaches used in the approximately 120 industrial wastewater ZLD facilities in the United States. Processing Schemes 1A, 1B, 2A, 2B, and 3 are evaluated in Chapters 4 and 5. Processing Scheme 4 for selective salt recovery is evaluated in Appendix 5.

GENERAL PROCESSING SCHEME 1

Conventional ZLD plants almost exclusively involve the use of thermal brine concentrators in one of the following three schemes:

A: concentrate → pretreatment (minimal) → thermal brine concentrator → evaporation pond (standard/enhanced)

B: concentrate → pretreatment (minimal) → thermal brine concentrator → crystallizer → solids (with possible very small blowdown stream)

GENERAL PROCESSING SCHEME 2

More recently, *high-recovery RO systems* have been used to reduce the volume of solution going to the thermal brine concentrators. The processing schemes include:

A: concentrate → treatment → second-stage RO → thermal brine concentrator → evaporation pond
B: concentrate → treatment → second-stage RO → thermal brine concentrator → crystallizer → solids

In general, the treatment used in this high-recovery RO scheme is significantly more intensive than in Processing Scheme 1.

GENERAL PROCESSING SCHEME 3

In some cases, the two-stage membrane system is used alone without any thermal processing:

concentrate → treatment → second-stage RO → evaporation pond

GENERAL PROCESSING SCHEME 4

Another more recent processing scheme (Geo-Processors, 2007), not yet used in the United States, entails selective or sequential recovery of individual salts from the effluent. The processing scheme used depends on the particular effluent being treated:

concentrate → salt removal → second-stage RO → salt removal → brine concentrator → mixed salts recovery

Figure 3.2. Frequently used ZLD processing schemes.

3.2.1 Characteristics of Conventional ZLD Technologies

Most conventional ZLD plants utilize mechanical vapor recompression evaporator technology for the primary desalination step. This technology can take low-salinity effluent or waters up to approximately 160,000–360,000 mg/L TDS, depending on the specific initial water quality involved. The degree to which a feedwater may be concentrated may be limited by (1) the onset of precipitation of salts, such as glauberite (sodium-calcium sulfate), sodium sulfate, or sodium chloride, or (2) the production of an unacceptably high level of suspended solids. The theoretical range of concentration factors possible is considerable, from about 1.67 (for a feed salinity of 60,000 mg/L and a limitation of 160,000 mg/L) to 90 (for a feed salinity of 4000 mg/L and a limitation of 360,000 mg/L).

In most cases, the evaporator surface area is selected to produce a power demand of 75–95 kWh/1000 gal of feedwater flow. Note that a volume feed to a mechanical vapor recompression evaporator preconcentrated to 60,000 mg/L is about 15 times less than for a feed of 4000 mg/L and thus would require significantly less energy. Preconcentration may be done by a second membrane stage when sufficient pretreatment is done to remove potential scalants. With this theoretical limit of extensive pretreatment, the concentration factor of the RO unit is limited by osmotic pressure, as in the case of seawater RO. This limit is typically reached at a TDS level of 60,000–70,000 mg/L for a sodium chloride-dominated water. The energy requirement for RO systems is considerably less than that for thermal evaporation, and thus the energy savings possible with preconcentrating the feedwater to the evaporator system are evident.

The pretreatment necessary to remove potential scalants to enable a two-stage RO treatment (high-recovery RO system) can be considerable. This is especially true of high-hardness waters where softening is required. In this situation, lime softening can require large amounts of chemicals and produce large amounts of solids.

The brine produced by the brine concentrators for a 1-MGD system is typically from 2–10% of the feed level and thus in the range of 20,000–100,000 gpd. Water recovery relative to the feed to the thermal evaporators is in the range of 90–99%. This brine is typically treated either by evaporation ponds, crystallizers, or spray dryers. A single crystallizer may be used if the volume is 50 gal/min (gpm) or less. When treated by a thermal evaporative crystallizer, most if not all of the remaining water can be recovered. The very high power consumption for vapor compression crystallizers falls in the range of 200–250 kWh/1000 gal of feedwater. In cases of high levels of the very soluble calcium or magnesium chlorides, a small blowdown stream from the crystallizer may be necessary to prevent excessive buildup of these species in the vapor body, which can lead to excessive boiling point elevation and prevent production of dry cake. If the volume of brine is in the range of 2–6 gpm, steam-driven crystallizers are more economical. Steam can be supplied by a package boiler or from a process source, if one is available. For larger systems, electrically driven vapor compressors are normally used to supply heat for evaporation. Small volumes of less than 10 gpm may also be treated by spray dryers.

The evaporative processes produce a very pure product water, typically 10 mg/L or less of TDS, depending on the salinity of the feed.

Many of the ZLD plants are associated with power plants, and final disposal may be via evaporation ponds in single-use ponds on the plant property. Proper design of an evaporation pond associated with a zero discharge plant dictates that the pond has sufficient depth to accommodate the deposition of precipitated solids over the life of the plant. Even then, depending on the actual operating conditions, it may be necessary at some point to dredge the pond or build new ponds.

A relatively new technology is that of enhanced evaporation ponds, such as the Israeli WAIV system (Lesico, 2007). This system employs a floating surface that has as much as 33 times the wetted surface area as that of the footprint. It covers part of the pond in strips and can result in a land requirement of only one-fifth that of conventional ponds.

With the exception of enhanced evaporation, these conventional technologies comprise most of the ZLD processing schemes in the United States and therefore are the processing schemes that have been evaluated in this study.

Since the Geo-Processors selective salt recovery technology is not yet commercial in the United States, the author has chosen to discuss that technology in Appendix 5. Emerging technologies that might impact high-recovery and ZLD processing in the near future are discussed in Chapter 6.

3.2.2 Installed Wastewater ZLD Plants in the United States

Table 3.1 lists 82 of the installed wastewater ZLD plants in the United States. It lists plants from two of the three major system producers: GE-Ionics-RCC and HPD. The vast majority of ZLD plants in the United States are RCC systems. The third system producer is Aquatech, which is estimated to have less than 10 wastewater ZLD facilities in the United States.

3.3 REVIEW OF VOLUME REDUCTION TECHNOLOGIES

There are some commercially available technologies that have been used in other industries. There have also been several noncommercial research studies funded by municipal desalination-oriented groups (Awwa Research Foundation, U.S. Bureau of Reclamation, Desalination Research and Innovation Partnership [DRIP], etc.).

Table 3.1. Domestic Wastewater Brine Concentrator Installations by State

Client	Plant	Location	Startup	Vol, gpm	Industry	Provider
Asarco	Hayden	AZ	1984	200	Copper smelt	RCC
IBM	Tucson	AZ	1985	100	Electronics	RCC
Salt River Project	Phoenix	AZ	1986		Power	HPD
Apache Nitrogen	Benson	AZ	1994	90	Fertilizer	RCC
Calpine	South Point	AZ	2000	400	Power	RCC
Intel	Fab 22	AZ	2001	155	Manufacturing	RCC
Pinnacle West Energy	West Phoenix	AZ	2001	100	Power	RCC
Pinnacle West Energy	Phoenix	AZ	2001		Power	HPD
Pinnacle West Energy	Tonopah	AZ	2001		Power	HPD
NEPCO	Gila River	AZ	2003	900	Power	RCC
Salt River Project	Navajo	AZ	1974-1980	1400	Power	RCC
Los Angeles County	Pitchess	CA	1988	150	Power	RCC
Thermo Electron Corp.	Mendota	CA	1988	38	Power	RCC
National Energy Constructors	Hanford	CA	1989	75	Power	RCC
Aerojet	Sacramento	CA	1990	15	Chemical	RCC
Harbert International	Tracy	CA	1990	50	Power	RCC
Walsh Construction	Mecca	CA	1991	50	Power	RCC
Simpson Paper	Ripon	CA	1993	11	Paper	HPD
La Paloma Generating	La Palmoa	CA	2001	600	Power	RCC
Constellation Energy	Victorville	CA	2001		Power	HPD
FP&L	Blythe	CA	2003	420	Power	RCC
SMUD	Cosumnes	CA	2004	280	Power	RCC
Con Agra	King City	CA	2005	50	Food	RCC
Public Service of CO	Hayden	CO	1976	250	Power	RCC
Tri-State Generation & Trans.	Craig	CO	1978	700	Power	RCC
Public Service of CO	Pawnee	CO	1980	450	Power	RCC
City of Colorado Springs	RD Nixon	CO	1980	350	Power	RCC
Calpine	Rocky Mountain	CO	2003		Power	HPD
City of Gainesville	Deerhaven	FL	1981	300	Power	RCC
Orlando Utilities Commission	Orlando	FL	1986			
Orlando Utilities Commission	Stanton	FL	1993	600	Power	RCC
Northern Canadian Power	Lake	FL	1993	3	Power	RCC
Pasco Cogen/Zurn/Nepco	Pasco	FL	1993	3	Power	RCC
Misson Energy	Auburndale	FL	1993	130	Power	RCC
Bechtel/US Generating	Cedar Bay	FL	1993	300	Power	RCC
Bechtel/US Generating	Indiantown	FL	1994	580	Power	RCC
Pasco County	N Port Richey	FL	1997	30	Ash landfill	RCC
U.S. Navy	Jacksonville	FL	1997	3.5	Mfg cleanup	RCC
Tallahassee Electric	St. Marks	FL	2000	230	Power	RCC
Orlando Utilities Commission	Stanton CC	FL	2004	500	Power	RCC
Cogentrix Rathdrum	Rathdrum	ID	2001	2	Power	RCC
Global Energy	Wabash River	IN	2002	100	Power	RCC
Brown & Root	Shreveport	LA	1990			HPD
CRS Sirrine	Stratton	ME	1988	20	Power	RCC
Westinghouse	Bellingham	MA	1991	30	Power	RCC
Harquahala Generating	Foxboro	MA	2001		Power	HPD
Tractabel/Parsons	Choctaw	Missouri	2005	106	Power	RCC
Resources Technology Group	Weldon Springs	MO	1995	50	Pit cleanup	RCC
Aquila Utilicorp	Aries Power	MO	2000	280	Power	RCC
Pennsylvania Power & Light	Colstrip	MT	1977	350	Power	RCC
Nevada Power	Clark	NV	1982	600	Power	RCC
Southern California Edison	Mojave	NV	1985	600	Power	RCC
Kerr-McGee Chemical Corp.	Henderson	NV	1989	100	Chemical	RCC
Stone & Webster	Apex	NV	2002	82	Power	RCC
Arizona Public Service	Four Corners	NM	1979	400	Power	RCC
Phelps Dodge	Hidalgo	NM	1991	200	Copper smelt	RCC
Public Service of NM	San Juan	NM	1974-1984	3300	Power	RCC
New York State Gas & Elec	Milliken	NY	1993	30	Power	RCC
Panda	Rosemary	NC	2002	50	Power	RCC
NC Power	E-Town	NC	2003	23	Power	RCC
East Penn	Lyon Station	PA	1995	100	Battery mfg	RCC
AES Ironwood	Ironwood	PA	2001	200	Power	RCC
Ocean State Power	Harrisville	RI	1990		Power	HPD
Ocean State Power	Harrisville	RI	1997	150	Power	RCC
EMI/Stone & Webster	Tiverton	RI	1998	20	Power	RCC
Otter Tail Power	Big Stone	SD	1980	600	Power	RCC
Texas Utilities	Monticello	TX	1982	250	Power	RCC
Texas Utilities	Martin Lake	TX	1983	500	Power	RCC
Texas New Mexico Power	Bremond	TX	1989	322	Power	RCC
Asarco	El Paso	TX	1991	150	Copper smelt	RCC
EI Dupont	Orange	TX	1992			HPD
Texas Independent Energy	Guadalupe	TX	2000	400	Power	RCC
Hays Energy	San Marcos	TX	2001	1,000	Power	RCC
Parsons/Tractebel	Wise County	TX	2002	60	Power	RCC
Utah Power & Light	Huntington	UT	1974	200	Power	RCC
US Army	Tooule	UT	1990			HPD
PacifiCorp	Huntington	UT	2002	200	Power	RCC
Fluor Daniel/Doswell	Doswell	VA	1991	90	Power	RCC
Northwest Alloys	Addy	WA	1981	80	Magnesium	RCC
Japanese Gas Corp	Hanford	WA	1994	20	Nuclear power	RCC
USKDK Corporation	Moses Lake	WA	1997			HPD
Coastal Chemical	Cheyenne	WY	1991	150	Chemical	RCC

Four of the commercial high-recovery systems that have patent protection and have had some degree of industry success are listed in Table 3.1.

- ARROW^{TM} by O’Brien & Gere
- HEEPM^{TM} (high-efficiency electro-pressure membrane) by EET Corporation
- HERO^{TM} (high-efficiency reverse osmosis) by Aquatech, Inc.
- VSEP^{TM} (vibratory shear enhanced processing) by New Logic

With the exception of the HERO^{TM} technology, these systems are considered emerging technologies.

3.3.1 ARROW^{TM} System by O’Brien & Gere (O’Brien & Gere, 2007)

The patents for the ARROW technology are process patents that use unique flow configurations of standard technologies to achieve high recovery. Treatment is after the second membrane step, with the treated water recycled to the interstage site. Silica is removed by high-pH caustic softening, which also removes carbonate species. In the absence of silica, ion exchange (IX) may be used. Water recovered from the IX or the precipitation step may be returned to the process at several locations. High recoveries are achieved, and pilot studies are under way in the pharmaceutical industry.

3.3.2 HEEPM^{TM} System by EET Corporation (EET, 2007)

EET Corporation has patents for both a high-efficiency electrodialysis (ED) system and a combined ED–RO system for high-recovery applications. While the ED system alone can achieve high recoveries, the combination system is usually less expensive. Feedwater is sent to a holding tank, from which water is fed to both an ED system and an RO system. A waste product is taken off the ED system, with the product water being returned to the feed tank. Product water is taken off the RO system, with the concentrate being returned to the feed tank. This unique processing arrangement can lead to unusually high recoveries. Such recoveries have been obtained in various nonmunicipal pilot and demonstration tests. Costs for treating low-salinity streams are significantly less than for treating higher-salinity streams. In each case, high recoveries may be obtained. A key is maintaining the TDS level in the feed tank at a relatively low level, which is accomplished by the ED system.

3.3.3 HERO^{TM} System by Aquatech (Aquatech, 2007)

The HERO^{TM} (high-efficiency reverse osmosis) system has been used successfully in an estimated 20 nonmunicipal applications around the world. The scaling potentials of calcium and carbonate species are drastically reduced by pretreatment or interstage treatment (typically lime softening or IX, depending on water quality), and if silica is present the second-stage RO is operated at a high pH. The HERO^{TM} system can produce brine suitable for further processing by a brine concentrator or disposal directly to evaporation ponds. The technology usually reduces the capital costs relative to the brine concentrator system. While also substantially reducing energy costs, these costs are replaced in part by chemical costs and solids disposal costs.

3.3.4 VSEP^{TM} System by New Logic Research, Inc.

VSEP^{TM} stands for vibratory shear enhanced process, where the entire flat sheet membrane device is vibrated to provide a high sheer force at the membrane solution interface. This force keeps scalants away from the membrane surface and even allows precipitates to be formed without substantially reducing membrane flux. The VSEP^{TM} systems have been used mainly for industrial applications involving difficult-to-treat waters, such as produced waters containing oil, food-processing wastewater, etc. Recently, New Logic Research has been considering “cleaner” water applications, including municipal ones. A successful potable water pilot test was recently conducted at the Big Bear Ski Resort in California (Lozier, 2007). The VSEP^{TM} system can incorporate any membrane (RO, nanofiltration [NF], ultrafiltration [UF], or microfiltration [MF]) and can typically achieve much higher recoveries than spiral-wound membrane systems.

As depicted in Figure 3.3, the processes discussed represent the various volume reduction (high-recovery) approaches. More specifically, four general approaches have been taken in these commercial and research efforts. The most common approach is treatment followed by a standard membrane step or two standard membrane steps with interstage treatment. The treatment in both cases minimizes the recovery limitations due to sparingly soluble salts and silica. This approach with some variation is used by the commercial HERO^{TM} technology of Aquatech and has been the subject of nearly all of the funded research studies. This approach is similar to processing Scheme 3 of Figure 3.2.

A second approach uses seeded slurry brine concentrators, the SPARRO^{TM} technology (Juby et al., 2000), which is a seeded slurry RO system, and the VSEP^{TM} system. The common element is that these processes allow precipitation, either by providing a site for the precipitates (the seeded slurry) or by creating a high shear environment where the precipitate’s effect on membrane transport is minimized (the VSEP^{TM} system).

A third approach is the HEEPM^{TM} technology, where two membrane processes take feed from a common feed tank. The fourth approach is the ARROW^{TM} technology, where the treatment step to remove sparingly soluble solids and/or silica is placed after two membrane stages, with treated water recycled back prior to one of the membrane stages.

The first approach, that of treatment of concentrate to reduce sparingly soluble salt and silica scaling potential, is addressed in the ZLD study, where the concentrate undergoes lime softening treatment, allowing second-stage membrane processing. The other three approaches, specifically the VSEP^{TM}, HEEPM^{TM}, and ARROW^{TM} systems, are characterized in Chapter 6.

3.4 REVIEW OF SELECTIVE SALT RECOVERY PROCESSING

3.4.1 Introduction

Geo-Processors (now Geo-Processors USA, Inc.) has successfully piloted and licensed their technology outside the United States to recover commercial-grade salts from many different waters. Documentation independent of Geo-Processors’ literature is generally lacking, as Geo-Processors’ results have been kept proprietary. The commercial viability in site-specific applications has not been demonstrated in the United States.

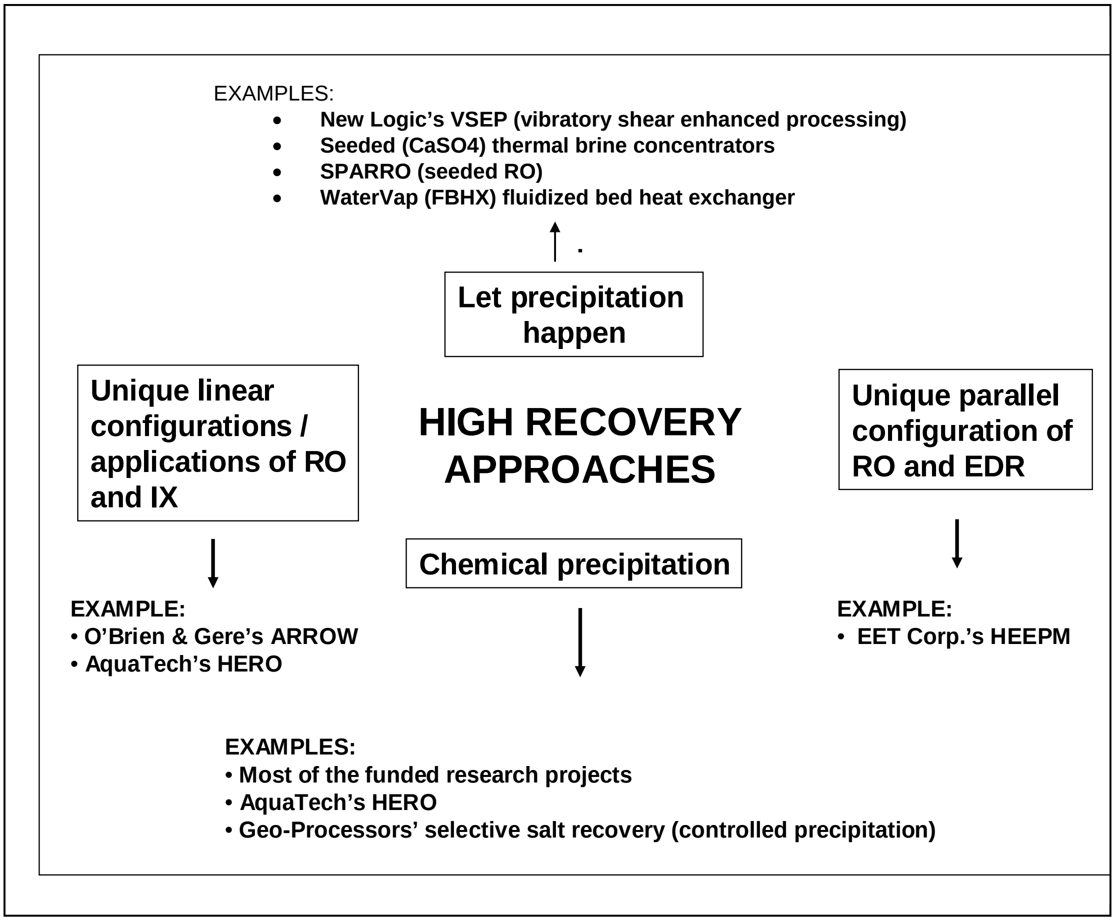
A recent analysis (Mickley, 2006b) showed that there are many applications for the major salts obtainable from concentrates and that many of the salts have sufficient value to make their sale economically attractive. The feasibility of a site-specific operation to recover and market salts, however, depends on several factors, including:

- Volume of concentrate
- Water quality (salts obtainable from the concentrate)
- Quality (form and purity) of salts obtained
- Reliability and consistency of salt quality
- Types of applications for the obtainable salts (types of markets)
- Existence of local market
- Size of local market
- Reliability of local market
- Combined income from sale of the different salts

Each site-specific consideration of the concept will require a feasibility analysis phase to address these and other issues prior to commitment to the concept. It is also important to note that market value is not directly related to economic feasibility. A sufficient mass of salts must be available to make processing and recovery feasible. There is likely a fundamental conflict that must be resolved between the economic structures of the function of producing water as a utility and producing salt or other by-products as price-variable commodities. There is a need to develop value-added products that utilize salts removed from concentrate to uncouple feasibility from dependence on existing markets, and this likely remains a significant challenge.

In general, salt separation and marketing of salts hold considerable promise to provide concentrate disposal solutions for many locations, including locations in the arid southwestern United States, where desalination plants are not being built due to the lack of a cost-effective concentrate disposal solution.

Of importance beyond providing cost-effective concentrate disposal solutions, the separation of salts, and their marketing, is a strong step toward achieving a sustainable, environment-supporting solution where water recovery is maximized and salts are recycled.



EXAMPLES:

- **New Logic's VSEP (vibratory shear enhanced processing)**
- **Seeded (CaSO4) thermal brine concentrators**
- **SPARRO (seeded RO)**
- **WaterVap (FBHX) fluidized bed heat exchanger**

Let precipitation happen

Unique linear configurations / applications of RO and IX

HIGH RECOVERY APPROACHES

Unique parallel configuration of RO and EDR

Chemical precipitation

EXAMPLE:

- **O'Brien & Gere's ARROW**
- **AquaTech's HERO**

EXAMPLE:

- **EET Corp.'s HEEPM**

EXAMPLES:

- **Most of the funded research projects**
- **AquaTech's HERO**
- **Geo-Processors' selective salt recovery (controlled precipitation)**

Figure 3.3. High-recovery, volume reduction approaches.

3.4.2 Background

3.4.2.1 The Reason for Consideration of Salt Separation

There are locations, particularly in the arid southwestern United States, where desalination plants have not been built due to the lack of cost-effective concentrate disposal. The consideration of selective and sequential salt removal from concentrate and marketing of the salts has resulted from the logical consideration and elimination of other concentrate disposal options for these locations. It is also an approach that maximizes water recovery and ultimately represents a sustainable solution—an important goal.

An analysis of disposal options (Mickley, 2007a) for the Phoenix area (discussed in Chapter 2) suggested the following:

- Conventional disposal options are neither available nor cost-effective for the 20 MGD of concentrate proposed. This includes surface water disposal, disposal to wastewater treatment systems, deep-well injection, land application, and evaporation ponds.
- Concentrate, in general, does not have any use that also serves as a means of disposal. Beneficial uses of concentrate (as concentrate) do not necessarily solve the concentrate disposal challenge.
- If concentrate is processed in a conventional ZLD scheme to obtain solids, the mixed salts obtained from concentrate, in general, do not have any use and thus must be landfilled at considerable cost.
- Commercial ZLD technologies (thermal brine concentrators, not presently used for treating municipal concentrates) are energy-intensive and have high operating costs.
- The use of commercial volume reduction technologies (such as high-recovery RO systems) prior to brine concentrators can lower capital costs significantly but still result in high operating costs, due to high chemical usage and high solids disposal costs.

In the long term, reducing ZLD processing costs will lower the water production costs associated with further processing concentrate and disposing of the solids, i.e., achieving a disposal solution. Improvements in desalination technologies that can be incorporated into ZLD processing schemes are the subject of research and may in time have this impact. In the short term, recovery and sale of individual salts may be the only option to significantly reduce operating costs and thus impact the total cost associated with disposal. The practicality of this possibility has been given considerable support by the identification of the Australian (and now a U.S.) company Geo-Processors, which reports to have successfully done this in several commercial ventures on a wide range of waters outside the United States. It is recognized that water production is in the realm of a critical utility, whereas salt production and sales rate as a commodity. It is inevitable that at some point water production will need to continue with no significant market for the salt produced. Consequently, there will always be a need for brine or salt disposal capabilities, if only as a backup to commercial distribution of recovered products.

There is also a larger need served by consideration of individual salt recovery. Landfills, even if they are a cost-effective means of disposing of salts, may at some point become sources of pollution. The counterargument in terms of sustainability is that if lined cells are used, landfilling is a sustainable practice for at least the near future. The mass per volume of salt disposed in landfills is considerably less than that of typical solid waste, for which we have no alternative. The only truly sustainable solution to concentrate disposal over the long term is recovery of most of the water and recovery and use of the salts. The recovery of most of the water is highly desirable in regions with limited water resources.

3.4.2.2 Salts Recoverable from Concentrate

As water is evaporated from a mixed salt solution, salts will precipitate (or be removed from solution) in a sequence according to their solubility and propensity for coprecipitation and adsorption. Precipitation is dependent on the particular salt solution, temperature, pH, residence time, agitation, presence of other species, such as antiscalants, and other variables. Control of salt in terms of crystal size, morphology, and purity, for example, is a complex function of such variables and an important consideration in defining a technical approach to produce the salts.

A general sequence of common salt precipitations is obvious from experience with RO, thermal brine concentrators, and crystallizers (Mickley, 2007a). RO is limited by the precipitation of sparingly soluble salts that include calcium carbonate, silica, calcium sulfate, and others. Use of antiscalants (and, historically, acid) has allowed operation past the saturation level of such sparingly soluble salts. With brackish sources, a limitation due to sparingly soluble salts may still exist even with extensive pretreatment. With higher-quality waters, when extensive pretreatment has removed the limitation due to sparingly soluble salts, the osmotic force becomes the limiting factor for second-stage RO recovery. Practical limits on pressure to overcome osmotic forces result in second-stage (or seawater) RO concentrates in the range of 65,000–75,000 mg/L for NaCl-dominated waters. Thermal brine concentrators, which are frequently used to process concentrate and other wastewater, are often limited by the formation of sodium sulfate, sodium carbonate, and eventually sodium chloride precipitates. They typically produce brine in the range of 160,000–360,000 mg/L of TDS. Crystallizers operating on the brine from thermal brine concentrators will precipitate Na_2CO_3, Na_2SO_4, and NaCl but require a blowdown stream for the highly soluble salts $CaCl_2$ and $MgCl_2$.

From this general consideration one can see the promise of sequential and thus selective removal of salts from solution, with a likely sequence being (from top to bottom):

General solubility level	**Salt examples**
Sparingly soluble salts	Calcium carbonate, calcium sulfate
Moderately soluble salts	Sodium carbonate, sodium sulfate
Soluble salts	Sodium chloride
Highly soluble salts	Calcium chloride, magnesium chloride

Prediction of the salt precipitation sequence in terms of amounts corresponding to physical and chemical conditions is difficult. Most software programs are limited in one or more ways, in part due to the facts that they were not designed to perform these calculations and many double salts are poorly characterized and seldom encountered. Software models used for estimating possible membrane system recovery with antiscalant use are limited in terms of the different salts included, the salinity range, accuracy (due to the inclusion of safety factors), and difficulty of incorporation into an iterative calculation necessary for defining the precipitation path.

A more accurate precipitation path calculation is made possible by using various geochemical speciation programs that determine how a given water will separate into liquid and solid phases. These programs are limited, however, in that they (with few exceptions) do not predict the pH change that takes place upon precipitation involving carbonate species, are limited in terms of salinity range, are generally difficult to use, and are not suited for sequential application to predict the precipitation path of a solution as it becomes more concentrated. There have been few published studies of the predictive capabilities of the software programs (Bourcier et al., 1996; Huff, 2004) to predict precipitation pathways, including amounts precipitated, effects of pH, and other factors.

While there may be minor salts of high value (Dirach et al., 2005) that would shift the economics of concentrate disposal through their recovery, the present review is focused on the removal of bulk salts that offer the opportunity for improving the cost-effectiveness and lessening the environmental impact.

Table 2.3 presented a list of major salts. The individual salts are listed along with several application areas for each salt. Some of these salts are discussed further in Appendix 5.

3.4.3 General Processing Schemes

In selective salt recovery, the solution is concentrated and treated in a series of steps to obtain the individual salts in their desired form (which is dependent on the marketable use in question). Concentration steps bring the solution near the point of precipitation for the salt. Treatment steps cause the salt in question to precipitate so that it can be recovered in the desired form and purity.

Where multiple salts are recovered and where they have a wide range of solubility, the processing may involve a series of alternating concentration and treatment steps that recover the salts sequentially from the lowest to the highest solubility. The concentration (desalination) steps include RO, NF, ED–EDR, thermal evaporation, crystallization, evaporation ponds (including enhanced evaporation ponds), and solar ponds. The treatment steps include such operations as pH adjustment, chemical addition, temperature control, thickening, and washing.

Some salts whose market values are low (e.g., some forms of NaCl and $CaSO_4$) cannot be cost-effectively processed by equipment-intensive processing schemes. These salts are more typically recovered by solar pond treatment of specialized water or wastewater concentrated in the salt of interest. This becomes a climate-dependent and land-intensive process that is not suitable for most locations.

While there are some applications of crude salts of lower quality, many applications require salts to meet quality specifications that may include form, size, and purity. Salts obtained from the initial precipitation may need to be washed to remove surface impurities and even redissolved and then recrystallized or reformed to remove “bubbles” of impurities of highly soluble salts. This processing also allows control over crystal size. Reformed NaCl is produced from crushing, grinding, and dissolving NaCl crystals and then crystallizing under very controlled conditions, usually indoors, away from any influence of climate. This “refinement” of salt quality adds value to the salt at the expense of additional production cost.

In general, there is a need for equipment and processing independent of climatic changes (temperature or rainfall) to allow accurate control of salt characteristics to meet product specifications. This is typically not possible with solar ponds.

Geo-Processors USA, Inc., has such technology and has applied it to a variety of situations. Some details and insights into the Geo-Processors technology are available in their patents. While the exact processing conditions and treatment sequence for a given salt recovery operation are not evident from the patents, it is apparent that the key to their success is a detailed and in-depth understanding of the many possible chemical reactions that can take place, including how the reactions are affected by temperature, pressure, pH, and other salts and chemicals present. This understanding allows for precise control and tailoring of processing conditions for a wide range of water qualities and salts.

For any new technology to be suitable for broad application to salt production, it must allow considerable control over processing conditions. As with all processes, improvements can be achieved via a reduction in processing throughput (decreased residence time) of each processing step. This reduces the equipment size and likely the footprint of the technology.

The consideration of selective salt recovery marks an important milestone in concentrate management in which some degree of beneficial use may be made of any concentrate. In recovering salts, environmental impacts are also reduced. The Geo-Processors technology is characterized further in Appendix 5.

Dow Chemical recently licensed a salt recovery technology with the intent of commercializing it (Dow, 2006). The technology was developed by Dr. Tom Smith of the University of South Carolina and is owned by ZDD, Inc. The ZDD process uses ED to reduce the salinity of concentrate from an RO stage so that the salt-depleted stream can be recycled to the RO stage to improve recovery of potable water. Residual solutions can be evaporated to dryness to produce various salts. The technology is not yet commercial.

CHAPTER 4

METHODOLOGY FOR ECONOMIC EVALUATION

4.1 INTRODUCTION

The key milestones in defining the methodology used in providing an economic evaluation of commercial ZLD systems include the following:

- Identification of process schemes
- Identification of variables
- Identification of meaningful values and ranges of variables
- Determination of the approach to study these variables
- Definition of the approach to predict sizes of process components
- Definition of the approach to predict process performance
- Definition of the approach to estimate costs of process and processing

Each of these areas is discussed in this chapter.

4.2 SYSTEM VARIABLES AND STUDY CASES

Many variables affect the design and resulting costs of ZLD systems. These include:

- Concentrate flow rate (million gallons per day) to be treated
- Concentrate salinity (milligrams per liter) to be treated
- Concentrate chemical composition
- Location (climate, local utility costs, local solids disposal costs, etc.)

4.2.1 Flow Rate and Salinity Variables

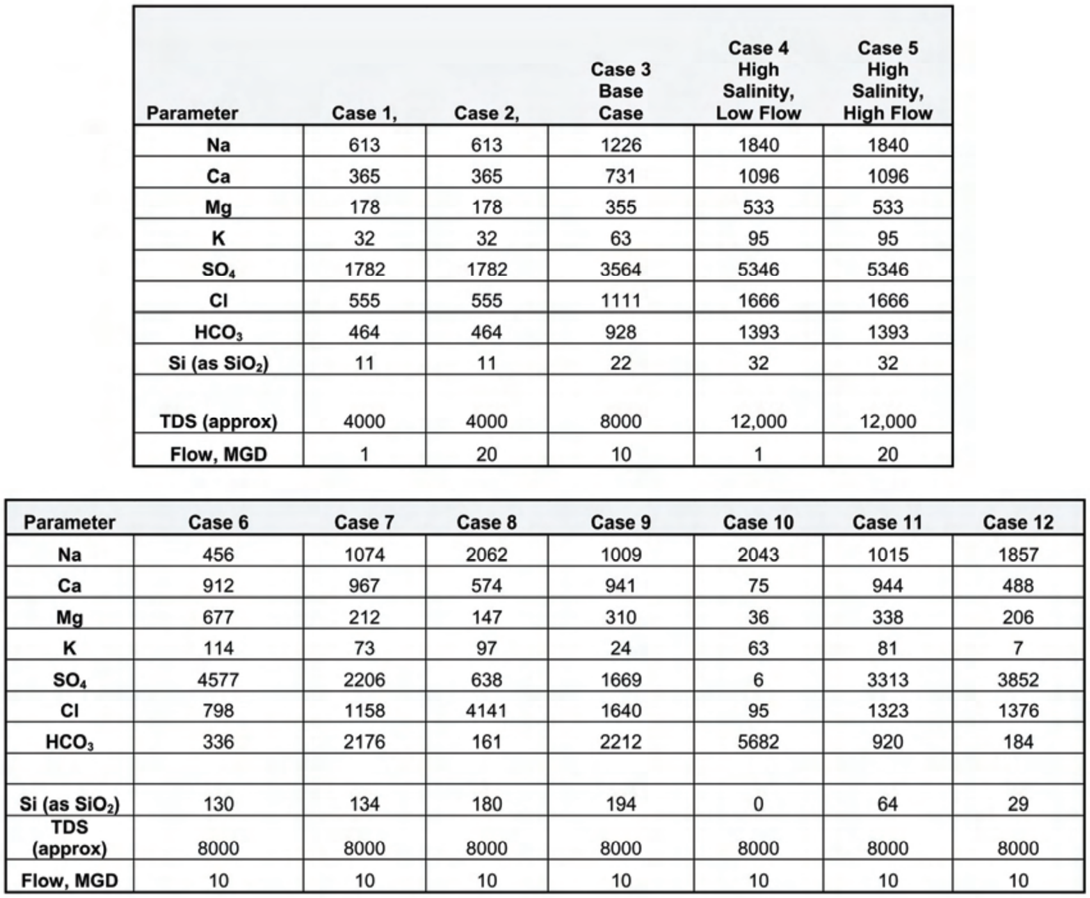
A project challenge was to cost-effectively evaluate the influence of these variables on ZLD system costs. For this purpose, it was convenient to study the variables in two groups. The first group includes flow rate and salinity and evaluates their influence on cost at a base case chemical composition. The base case composition chosen is a projected concentrate water quality from Phoenix (USBR, 2000) adjusted to a reference salinity of 8000 mg/L, as shown in Table 4.1. This base case is identified as Case 3.

Flow rates are set at 1, 10, and 20 MGD to cover a wide range of realistic current and future concentrate flows being considered. The largest present inland projected concentrate flow is 5 MGD (El Paso, 2007). A 20-MGD flow was used in a U.S. Bureau of Reclamation study for the Phoenix area conducted in 2000 (USBR, 2000), and large concentrate flows were considered in a recent Southern Nevada Water Authority study (SNWA, 2006).

The salinities are set at 4000, 8000, and 12,000 mg/L, which covers the range of brackish water concentrate salinities found in the United States (Mickley, 2006a).

The compositions at 4000 and 12,000 mg/L were determined by straight dilution or concentration of the 8000-mg/L concentrate of Case 3. Every resulting case was checked for anion–cation balance and adjusted as necessary. Each final water quality case (salinity and composition) was simulated using RO membrane software to make sure it corresponded to an achievable concentrate. Cases 1–5 of Table 4.1 show the specific parameters used for studying the salinity and flow rate variables.

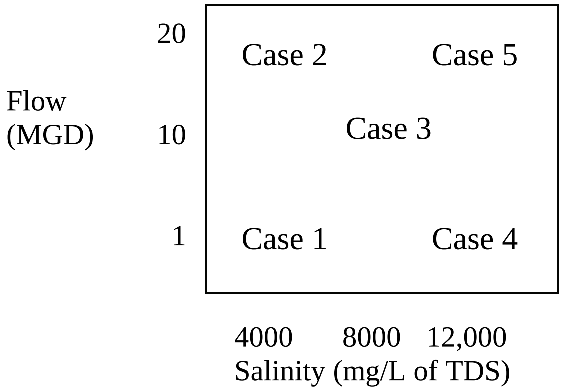
Table 4.1. Chemical Composition, Salinity, and Flow for the 12 Cases



Parameter	Case 1,	Case 2,	Case 3 Base Case	Case 4 High Salinity, Low Flow	Case 5 High Salinity, High Flow
Na	613	613	1226	1840	1840
Ca	365	365	731	1096	1096
Mg	178	178	355	533	533
K	32	32	63	95	95
SO_4	1782	1782	3564	5346	5346
Cl	555	555	1111	1666	1666
HCO_3	464	464	928	1393	1393
Si (as SiO_2)	11	11	22	32	32
TDS (approx)	4000	4000	8000	12,000	12,000
Flow, MGD	1	20	10	1	20

Parameter	Case 6	Case 7	Case 8	Case 9	Case 10	Case 11	Case 12
Na	456	1074	2062	1009	2043	1015	1857
Ca	912	967	574	941	75	944	488
Mg	677	212	147	310	36	338	206
K	114	73	97	24	63	81	7
SO_4	4577	2206	638	1669	6	3313	3852
Cl	798	1158	4141	1640	95	1323	1376
HCO_3	336	2176	161	2212	5682	920	184
Si (as SiO_2)	130	134	180	194	0	64	29
TDS (approx)	8000	8000	8000	8000	8000	8000	8000
Flow, MGD	10	10	10	10	10	10	10

Schematically, the five cases represent the corner and center points of the variable space shown below:



4.2.2 Chemical Composition Variables

The starting point for consideration of the chemical composition variables are the major ions and silica:

- Cations: Na^+, Ca^+, Mg^+, and K^+
- Anions: SO_4^-, Cl^-, and HCO_3^-
- Other: Silica

Together, these ions typically constitute nearly all of the brackish water ionic composition and, together with silica, most of the dissolved solids composition (TDS). It is of interest to consider the variability that these constituents have on processing costs. In general, waters may be considered to be sodium chloride dominated or not. Sodium chloride-dominated waters are typically the least problematic for brackish desalination processes. Potassium, typically present in lower amounts, is similarly not generally a processing problem. From the perspective of RO processing, the constituents of most frequent concern are Ca, SO_4, HCO_3, and silica, all of which can contribute to scaling of the membrane through solubility or precipitation. This is true of most desalination processes, including thermal brine concentrators. While magnesium can present hardness challenges similar to calcium, in most brackish waters in the western United States magnesium levels are lower than and follow the calcium levels. While these distinctions are not absolute, they are strong enough to allow meaningful simplification of the number of water quality constituents to be studied. The species chosen for study were thus Ca, SO_4, HCO_3 and silica.

In Table 4.1, Cases 6–12 list the various site-specific chemical compositions chosen for this study. The cases are treated generically but in total they represent sites that include Las Vegas, Phoenix, Denver, El Paso, Southern California, and Wyoming.

To decouple the study of chemical composition from the study of salinity and concentrate flow, the salinity of Cases 6–12 was standardized to a base case salinity of 8000 mg/L. Some of the actual salinities were in effect diluted to achieve the 8000-mg/L level, and some were concentrated. Evident from Table 4.1 is the wide range of conditions represented by these 12 cases. Variable ranges are as follows:

- Salinity: 4000–12,000 mg/L
- Flow: 1–20 MGD
- Calcium: 75–1096 mg/L
- Sulfate: 6–5346 mg/L
- Bicarbonate: 161–2212 mg/L
- Silica: 11–194 mg/L

To summarize: the variables of Cases 1–5 are salinity and flow, with all five cases having the same chemical composition; the variable of Cases 6–12 is the chemical composition, with all seven cases having the same salinity and flow. This approach allows the salinity and flow variables to be uncoupled from the chemical composition.

4.3 ZLD PROCESS CONSIDERATIONS

4.3.1 Process Schemes Considered

Based on the information summarized in Figure 3.2, the process schemes chosen for evaluation are shown in Table 4.2. The choices, as stated in Chapter 2, reflect process steps presently used in ZLD schemes in other industries. The first of five process schematics and the water quality at each processing step for the 12 variable cases are provided in Figure 4.1. Corresponding schematics for the other four process schemes are provided in Appendix 1.

4.3.2 Process Sizing, Performance, and Cost Calculations

Table 4.3 provides an overview of the approach to sizing equipment, to determining process step performance, and to developing costs for the processing scheme. Table 4.4 lists the various assumptions, based on typical values, made to assess process performance and to estimate costs.

4.3.2.1 Brine Concentrator and Crystallizer

Sizing for a brine concentrator and crystallizer is based on the feed flow rate. Performance, however, is not so simple, being dependent on the salinity and chemical composition of the feedwater. Performance of the brine concentrator and crystallizer are critical to the determination of process scheme costs. GE-Ionic-RCC, manufacturer of perhaps 80% or more of the evaporative ZLD systems in U.S. industry, was helpful in providing support on estimating performance, and some of the information in Table 4.5 reflects their input. Table 4.5 is a listing of brine concentrator and crystallizer parameters for the 12 water quality cases and the processing schemes. The data in Table 4.5 also reflect several characteristics of brine concentrators and crystallizers. These include:

- The degree of concentration taking place in the brine concentrator and crystallizer is dependent on the chemical compositions in different effluent salinities (column 4 for the brine concentrator and column 13 for the crystallizer) and effluent flow rates (column 10 for the brine concentrator).
- Performance of the brine concentrator is typically limited (column 7) by the formation of glauberite, $Na_2Ca(SO_4)_2$, the precipitation of NaCl, or excessive suspended solids levels.
- The chemical composition determines the energy requirement (a function of the boiling point rise) and is reflected in column 8.
- Materials of construction are also a function of chemical composition (column 9).
- The effluent from the brine concentrator is the feed flow to the crystallizer (column 11).
- Crystallizer performance is dependent on chemical composition of the brine from the brine concentrator. High levels of very soluble salts (such as $CaCl_2$ and $MgCl_2$) require a purge stream (column 12).
- Salinity and volume of the purge stream are given in columns 13 and 14, respectively.

Table 4.2. Commercial ZLD Process Schemes Chosen for Evaluation

Scheme	Processing Step Sequence
1A	Conc. → BC → EP
1B	Conc. → BC → Cryst. → EP and LF
2A	Conc. → LS → RO2 → BC → EP and LF
2B	Conc. → LS → RO2 → BC → Cryst. → EP and LF
3	Conc. → LS → RO2 → EP and LF

Note: Conc. = concentrate; BC = brine concentrator; EP = evaporation pond; Cryst. = crystallizer; LF = landfill; LS = lime softener; RO2 = second-stage RO.

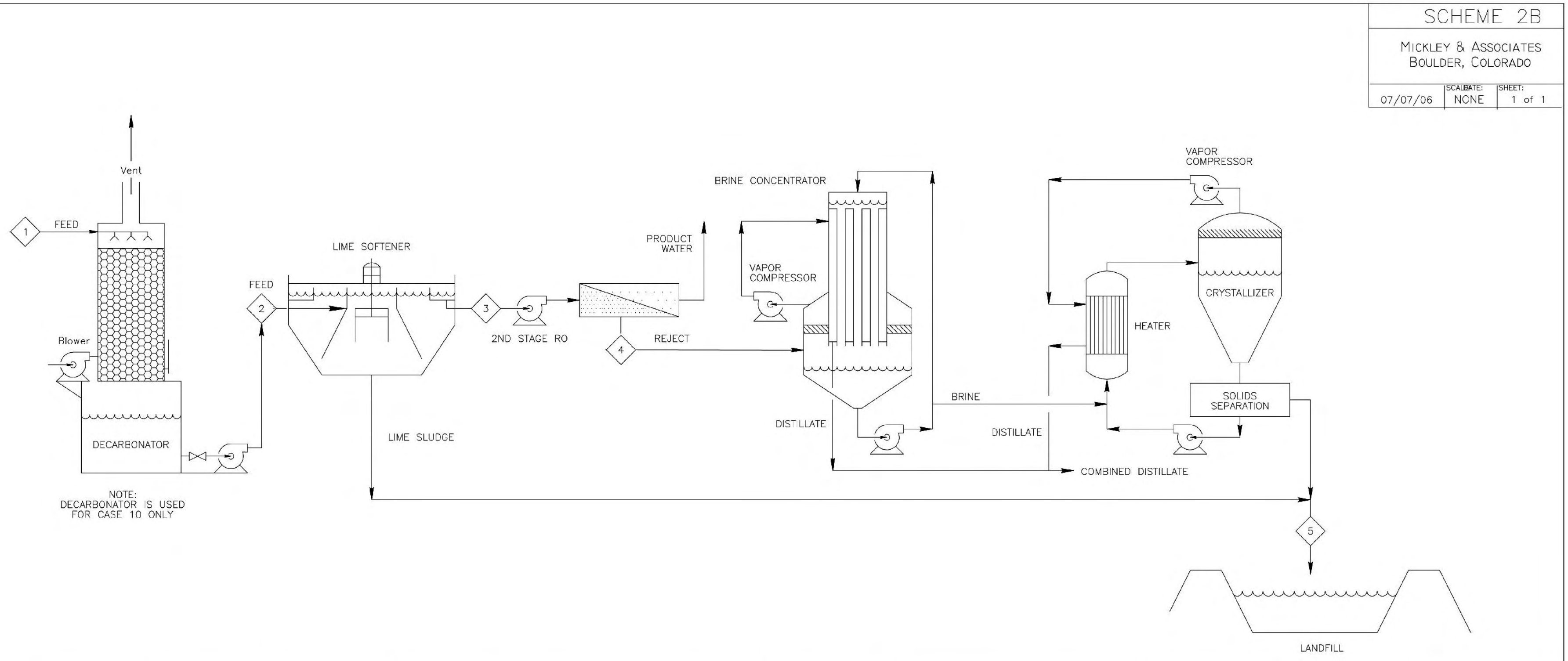
4.3.2.2 Lime Softener and Second-Stage RO

Sizing for a lime softener and second-stage RO system is based on the incoming flow rate. It is assumed that dewatering the lime softening solids to 50% solids results in negligible water loss at the lime softening step. Thus, the feed flow to the second-stage RO is assumed to be the same as the feed flow to the lime softener. Lime softener performance was modeled by using the *AWWA Chemical Water Softening Model,* a program developed by Rothberg, Tamburini, and Windsor for the American Water Works Association which calculates dose and sludge produced for lime and soda ash softening. RO performance was modeled using the Hyd-RO-dose software for predictive modeling of chemistry for membrane systems. It was developed and is sold by French Creek Software, Inc. It serves the same function as programs available from membrane manufacturers in predicting possible recovery levels for a given water analysis. It has the advantage of being more comprehensive in the number of sparingly soluble salts included and in being more accurate than other estimation methods.

4.3.2.3 Evaporation Pond

Sizing or determination of the required acreage for an evaporation pond is based on the flow rate and the net evaporation rate assumed. A base net evaporation rate of 3.6 gpm/acre of fresh water was used. The net evaporation rate used in the calculations is less than the fresh water rate and is dependent on the salinity of the incoming flow. The rate of buildup of solids in the evaporation pond depends on the salinity and density of the incoming flow. High-salinity influent may result in ponds becoming full during the 20-year life of a desalination plant. In this situation, the ponds would be covered and new ponds would be constructed to replace them.

Regarding landfill, high solids streams are produced in the lime softening step (50% solids) and the crystallization step (85% solids).



	CASE 1					CASE 2					CASE 3				
	1	2	3	4	5	1	2	3	4	5	1	2	3	4	5
Flow- MGD	10	10	10	1.210	522 (Dry TPD	1	1	1	0.060	26 (Dry TPD)	20	20	1	0.060	521 (Dry TPD
Na	1,226	1,226	2,451	19,695		613	613	1,245	19,364		613	613	1,245	19,364	
Ca	731	731	0	0.0		365	365	0	0.0		365	365	0	0.0	
Mg	355	355	0	0.0		178	178	0	0.0		178	178	0	0.0	
K	63	63	63	524.0		32	32	32	507.0		32	32	32	507.0	
SO_4	3,564	3,564	3,564	30,661		1,782	1,782	1,782	29,702		1,782	1,782	1,782	29,702	
Cl	1,111	1,111	1,111	9,166		555	555	555	8,788		555	555	555	8,788	
HCO_3	926	926	165	602.0		464	464	133	2302.0		464	464	133	2302.0	
Si (As SiO_2)	22	22	0	0.0		11	11	0	0.0		11	11	0	0.0	
TDS-Approx	8,000	8,000	7,350	60,803		4,000	4,000	3,743	61,010		4,000	4,000	3,743	61,010	
Landfill, Ac					230					12					231

	CASE 4					CASE 5					CASE 6				
	1	2	3	4	5	1	2	3	4	5	1	2	3	4	5
Flow- MGD	1	1	1	0.190	81 (Dry TPD	20	20	1	0.190	1593 (Dry TPD)	10	10	10	1.450	667 (Dry TPD)
Na	1,840	1,840	3,656	19,179		1,840	1,840	3,656	19,179		456	456	2,721	18,315	
Ca	1096	1096	0	0.0		1096	1096	0	0.0		912	912	0	0.0	
Mg	533	533	0	0.0		533	533	0	11.0		677	677	0	0.0	
K	95	95	95	501.0		95	95	95	501.0		114	114	114	774.0	
SO_4	5,346	5,346	5,346	28,943		5,346	5,346	5,346	28,943		4,577	4,577	4,577	24,634	
Cl	1,666	1,666	1,666	8,793		1,666	1,666	1,666	8,793		798	798	798	5,415	
HCO_3	1393	1393	189	989.0		1393	1393	189	989.0		336	336	163	0.4	
Si (As SiO_2)	32	32	0	0.0		32	32	0	0.0		130	130	0	0.0	
TDS-Approx	12,000	12,000	10,949	58,506		12,000	12,000	10,949	58,506		8,000	8,000	8,370	10	
Landfill, Ac					35					699					304

	CASE 7					CASE 8					CASE 9				
	1	2	3	4	5	1	2	3	4	5	1	2	3	4	5
Flow- MGD	10	10	10	0.920	495 (Dry TPD	10	10	10	1.300	361 (Dry TPD	10	10	10	0.920	493 (Dry TPD
Na	1,074	1,074	1,832	19106		2,062	2,062	2959	22,488		1,009	1,009	1,908	19,908	
Ca	967	967	0	0		574	574	16	122		941	941	0	0	
Mg	212	212	0	21		147	147	2	15		310	310	0	0	
K	73	73	73	771		97	97	97	737		24	24	24	253	
SO_4	2,206	2,206	2,206	24,634		638	638	638	5,431		1,669	1,669	1,669	18,933	
Cl	1,158	1,158	1,158	12,223		4,141	4,141	4141	31,472		1,640	1,640	1,640	17,311	
HCO_3	2176	2176	176	2,000		161	161	111	882		2210	2210	172	1,953	
Si (As SiO_2)	134	134	0	0.0		180	180	0	0.0		194	194	0	0	
TDS	8,000	8,000	5,441	58,953		8,000	8,000	7,964	61,156		8,000	8,000	5,409	58,678	
Pond Acres					234					146					242

	CASE 10					CASE 11					CASE 12				
	1	2	3	4	5	1	2	3	4	5	1	2	3	4	5
Flow- MGD	10	10	10	1.260	561 (Dry TPD)	10	10	10	1.240	548 (Dry TPD)	10	10	10	1.350	438 (Dry TPD)
Na	2,043	2,043	2,043	23,320		1,015	1,015	2,455	19,261		1,857	1,857	2,796	20,272	
Ca	75	75	16	127.0		944	944	0	0.0		488	488	0	0.0	
Mg	36	36	2	16.0		338	338	0	16.0		206	206	0	15.0	
K	63	63	63	499.0		81	81	81	641.0		7	7	7	51.2	
SO_4	6	6	6	1,550		3,313	3,313	3,313	27,152		3,852	3,852	3,852	28,956	
Cl	95	3,013	3,013	34,224		1,323	1,323	1,323	10,474		1,376	1,376	1,376	10,055	
HCO_3	5682	650	255	2,115.0		920	920	162	1347.0		184	184	154	1,171.0	
Si (As SiO_2)	0	0	0	0.0		64	64		0.0		29	29		0.0	
TDS	8,000	5,886	5,398	61,851		8,000	8,000	7,331	59,153		8,000	8,000	8,181	60,766	
Pond Acres					251					244					172

Figure 4.1. Process schematic for Scheme 2B.

Table 4.3. Sizing, Performance, and Costing Approaches to Different ZLD Processing Steps

Item	Sizing Approach	Performance Approach	Costing Approach
Thermal Brine Concentrator	Sizing is based on flow from previous process step	GE-Ionics-RCC modeling support	Cost model previously developed by M&A
Crystallizer	↓	GE-Ionics-RCC modeling support	Cost model previously developed by M&A
Lime Softener		AWWA chemical water softening model	Cost model previously developed by M&A
Second-Stage RO		Hyd-RO-dose software	Cost model previously developed by M&A
Evaporation Pond		Mickley and Associates (M&A) model of evaporation rate as a function of salinity; assumption of solids density, which along with salinity determines pond buildup of solids	Cost model previously developed by M&A
Landfill (dedicated)		Assumption of solids density, which along with salinity determines pond buildup of solids	Cost model previously developed by M&A

Table 4.4. Assumptions Used in Sizing, Performance, and Cost Analyses

Item	Assumptions
Lime Softener	
• Treatment level	To 10 mg/L Ca, 40 mg/L Mg; both are $CaCO_3$
• Silica removal (ppm)	Due to magnesium levels equal to or exceeding silica (SiO_2) levels
• Chemical cost	$75/ton for hydrated lime; $135/ton for 58% soda ash
• Dewatering	By filter press
• Capital cost	$700,000/MGD feed
• Operating cost	Chemical cost plus 2% of capital cost
Second-Stage RO	
• Capital cost	$2,500,000/MGD, installed
• Operating cost	$1.00/kgal product
Brine Concentrators	
• Energy use	Depends on chemical make-up of water; can be 75–95 kWh/kgal
• Cost of electricity	$0.08/kWh
• Capital cost	Depends on feed flow and chemical make-up: For cases limited by glauberite, 75 kWh/kgal For cases limited by NaCl, 95 kWh/kgal Other cases, 85 kWh/kgal
• Operating cost	90% of operating cost is energy
Crystallizer	
• Energy use	225 kWh/kgal
• Cost of electricity	$0.08/kWh
• Capital cost	Depends on feed flow
• Operating cost	95% of operating cost is energy
Evaporation Ponds	
• Net pan evaporation rate	3.6 gpm/acre
• Effect of salinity	Curve, 3.6 gpm/acre (fresh water) to 2.4 gpm/acre; 350,000 mg/L
• Maximum pond depth	12 ft
• Density of solids	74.9 lb/ft^3
• Capital cost	$300,000/acre
• Operating cost	1% of capital cost
• Salinity situations	~60,000 when concentrate is from second-stage RO; 145,000–358,000 mg/L when brine is from brine concentrator; 374,000–450,000 mg/L when brine is from crystallizer
• Action when ponds fill	Cover over pond, construct new pond
Landfill (dedicated)	
• Solids received	85% solids from crystallizer, 50% solids from lime softener
• Maximum landfill depth	10 ft
• Density of solids	143.6 lb/ft^3
• Capital cost	$250,000/acre
• Operating cost	1% of capital cost plus $10/ton for hauling
Miscellaneous Cost Factors	
• Administrative	Taken as 25% of capital cost
• Project contingency	Taken as 25% of capital cost
• Process contingency	Taken as 5% of capital cost
• Annualized cost basis	20 years, 8% interest

Table 4.5. Brine Concentrator and Crystallizer Performance Estimates

Situation 1: Brine Concentrator and Crystallizer Treatment, Schemes 1A and 1B

1	2	3	4	5	6	7	8	9	10	11	12	13	14
	Brine Concentrator									Crystallizer			
Case No.	Feed TDS, mg/L	Feed Flow, MGD	Effluent TDS, mg/L	Type of Evaporator	Chemical(s) Added	Limiting Parameter	Energy, kWh/kgal	Relative Material Cost	Effluent Flow, gpm	Crystallizer Feed, gpm	Crystallizer Purge?	Purge Salinity, mg/L	Purge Volume, gpm
1	8000	10	257,000	Seeded slurry	H2SO4	Glauberite	75	Low	210	210	Y	450,000	11
2	4000	1	257,000	Seeded slurry	H2SO4	Glauberite	75	Low	21	21	Y	450,000	0.6
3	4000	20	257,000	Seeded slurry	H2SO4	Glauberite	75	Low	421	421	Y	450,000	11
4	12,000	1	263,000	Seeded slurry	H2SO4	Glauberite	75	Low	30	30	Y	384,000	3
5	12,000	20	263,000	Seeded slurry	H2SO4	Glauberite	75	Low	604	604	Y	384,000	62
6	8000	10	261,000	Seeded slurry	H2SO4	Glauberite	75	Low	210	210	N	---	0
7	8000	10	226,000	Seeded slurry	H2SO4	Glauberite	75	Low	231	231	Y	411,000	7
8	8000	10	358,000	Seeded slurry	H2SO4	NaCl	95	High	154	154	Y	400,000*	80*
9	8000	10	226,000	Seeded slurry	H2SO4	TSS	85	Low	231	231	Y	374,000	29
10	8000	10	167,000	Seeded slurry	H2SO4	Glauberite	75	Low	302	302	N	---	0
11	8000	10	258,000	Seeded slurry	H2SO4	TSS	85	Low	210	210	Y	410,000	14
12	8000	10	199,000	Seeded slurry	H2SO4	Glauberite	75	Low	278	278	N	---	0

* = assumed

Situation 2: Brine Crystallizer or Crystallizer Treatment after Lime Softening and Second-Stage RO, Schemes 2A and 2B

	Brine Concentrator									Crystallizer			
Case No.	Feed TDS, mg/L	Feed Flow, MGD	effluent TDS, mg/L	Type Evaporator	Chemical(s) Added	Limiting Parameter	Energy, kWh/kgal	Relative Material Cost	Effluent Flow, gpm	Crystallizer Feed, gpm	Crystallizer Purge?	Purge Salinity, mg/L	Purge Volume, gpm
1	60,803	1.21	145,000	Seeded slurry	H2SO4/CaCl2	Glauberite	75	Low	352	352	N	---	---
2	61,010	0.06	145,000	Seeded slurry	H2SO4/CaCl2	Glauberite	75	Low	17.5	17.5	N	---	---
3	61,010	1.2	145,000	Seeded slurry	H2SO4/CaCl2	Glauberite	75	Low	350	350	N	---	---
4	58,506	0.19	145,000	Seeded slurry	H2SO4/CaCl2	Glauberite	75	Low	53.2	53.2	N	---	---
5	58,506	3.74	145,000	Seeded slurry	H2SO4/CaCl2	Glauberite	75	Low	1047	1047	N	---	---
6	57,623	1.45	145,000	Seeded slurry	H2SO4/CaCl2	Glauberite	75	Low	400	400	N	---	---
7	58,953	0.92	145,000	Seeded slurry	H2SO4/CaCl2	Glauberite	75	Low	282	282	N	---	---
8	61,156	1.3	263,000	Seeded slurry	H2SO4/CaCl2	NaCl	95	High	161	161	N	---	---
9	58,578	0.92	143,000	Seeded slurry	H2SO4/CaCl2	Glauberite	75	Low	284	284	N	---	---
10	61,851	1.26	247,000	Falling film	H2SO4	NaCl	95	High	219	219	N	---	---
11	59,153	1.24	145,000	Seeded slurry	H2SO4/CaCl2	Glauberite	75	Low	283	283	N	---	---
12	60,766	1.25	145,000	Seeded slurry	H2SO4/CaCl2	Glauberite	75	Low	291	291	N	---	---

Note: The other assumptions and parameters used in determination of size, performance, and cost are listed in Table 4.4. Costs of the process steps were estimated using in-house cost models and preliminary-level quotes from equipment vendors.

CHAPTER 5

ECONOMIC EVALUATION RESULTS FOR ZLD PROCESSING SCHEMES

5.1 INTRODUCTION

The methodologies of Chapter 4 were followed to produce the results presented and discussed in this chapter. Background calculations are contained in several tables. Example tables are provided in this chapter; the remainder of the tables can be found in Appendices 2 and 3.

5.2 REPRESENTATION OF RESULTS

5.2.1 Tabular Representation of Design, Performance, and Cost Parameters and Calculated Costs for Each Scheme

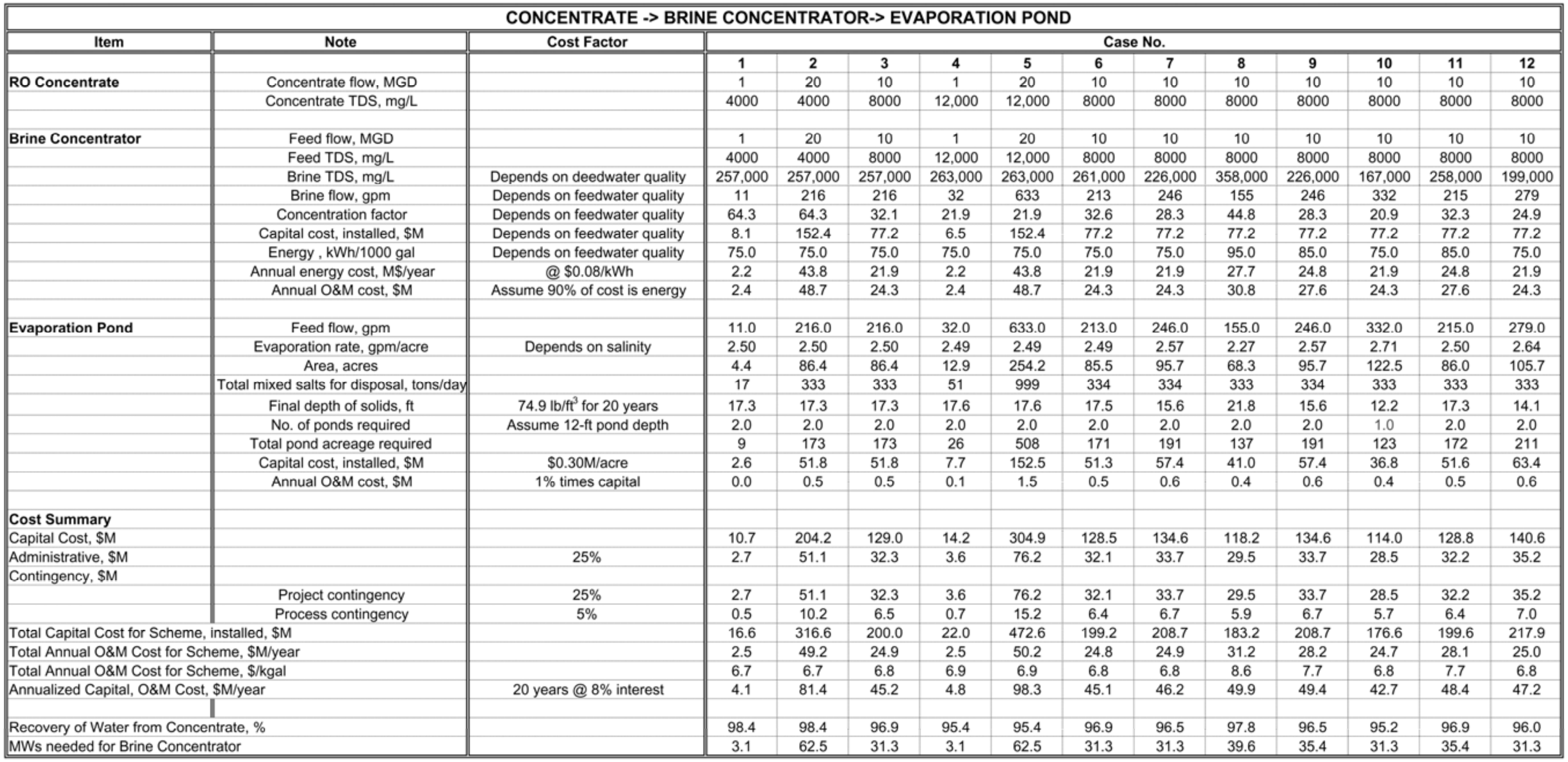
There are five master cost tables, one for each processing scheme. An example corresponding to processing Scheme 1A is given in Table 5.1. Tables for the other four processing schemes are in Appendix 2. These tables show the design, performance, and cost parameters for each processing step and for each of the 12 variable cases, with one table for each processing scheme. The tables represent the master tabulation of capital and operating costs for each process step and for the entire processing scheme. In a cost summary at the bottom of each table, administrative and contingency costs are added to the installed equipment costs to determine the total capital cost. An annualized cost is then calculated by amortizing the total capital cost at 8% interest over 20 years and adding the result to the operating cost. The percentage of concentrate recovered is given in the next-to-last row. Finally, the bottom row provides the energy need for the brine concentrator in terms of megawatts. All other tables and figures were generated using information from these tables.

Table 5.2 is a summary table for all capital, operating, and annualized costs for all processing schemes and variable cases. In addition to including information from the master cost tables, Table 5.2 introduces three additional cost parameters:

- *Unit capital cost,* defined as the capital cost divided by the gallons per day of feed. This parameter allows direct comparison of capital costs of different-sized processes.
- *Unit operating cost,* defined as the operating cost per 1000 gal of feed. This parameter allows direct comparison of operating costs for different-sized processes.
- *Unit annualized cost,* defined as the annualized cost divided by the gallons per day of feed. As with the unit capital cost and unit operating cost, this parameter allows direct comparison of annualized costs for different-sized processes.

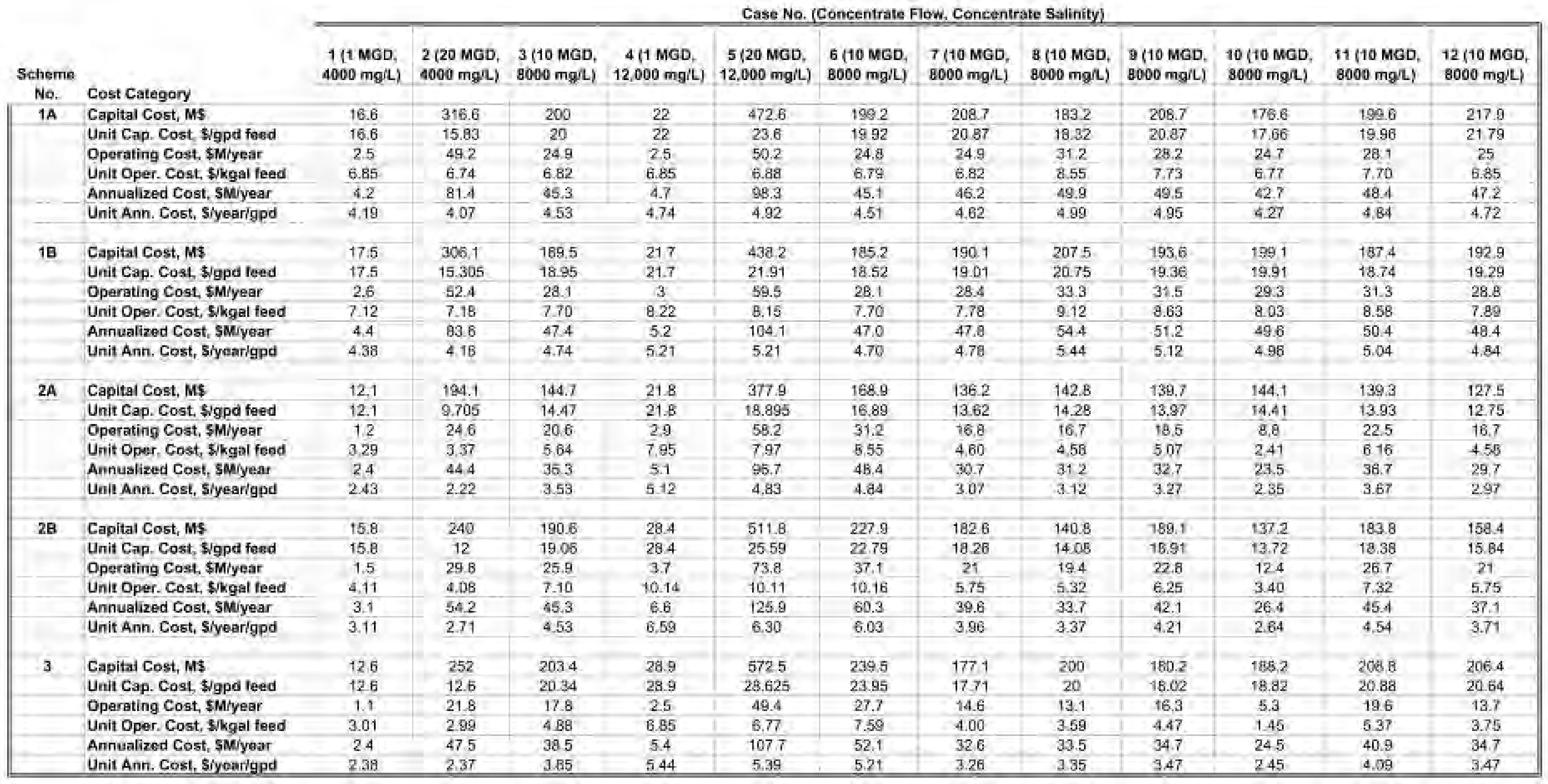
Graphical representation of the data in the master cost tables is provided in the following four sections, along with discussions of the graphs. Sections 5.2.2 and 5.2.3 deal with Cases 1–5. The independent variables affecting process costs are the salinity (TDS) and the plant size (in terms of the feed flow to the process). Sections 5.2.4 and 5.2.5 deal with Cases 6–12, where the only independent variable affecting process costs is the composition of the water.

Table 5.1. Performance, Design, and Cost Parameters for Process Scheme 1A



CONCENTRATE -> BRINE CONCENTRATOR-> EVAPORATION POND														
Item	**Note**	**Cost Factor**	**Case No.**											
			1	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**
RO Concentrate	Concentrate flow, MGD		1	20	10	1	20	10	10	10	10	10	10	10
	Concentrate TDS, mg/L		4000	4000	8000	12,000	12,000	8000	8000	8000	8000	8000	8000	8000
Brine Concentrator	Feed flow, MGD		1	20	10	1	20	10	10	10	10	10	10	10
	Feed TDS, mg/L		4000	4000	8000	12,000	12,000	8000	8000	8000	8000	8000	8000	8000
	Brine TDS, mg/L	Depends on deedwater quality	257,000	257,000	257,000	263,000	263,000	261,000	226,000	358,000	226,000	167,000	258,000	199,000
	Brine flow, gpm	Depends on feedwater quality	11	216	216	32	633	213	246	155	246	332	215	279
	Concentration factor	Depends on feedwater quality	64.3	64.3	32.1	21.9	21.9	32.6	28.3	44.8	28.3	20.9	32.3	24.9
	Capital cost, installed, $M	Depends on feedwater quality	8.1	152.4	77.2	6.5	152.4	77.2	77.2	77.2	77.2	77.2	77.2	77.2
	Energy , kWh/1000 gal	Depends on feedwater quality	75.0	75.0	75.0	75.0	75.0	75.0	75.0	95.0	85.0	75.0	85.0	75.0
	Annual energy cost, M$/year	@ $0.08/kWh	2.2	43.8	21.9	2.2	43.8	21.9	21.9	27.7	24.8	21.9	24.8	21.9
	Annual O&M cost, $M	Assume 90% of cost is energy	2.4	48.7	24.3	2.4	48.7	24.3	24.3	30.8	27.6	24.3	27.6	24.3
Evaporation Pond	Feed flow, gpm		11.0	216.0	216.0	32.0	633.0	213.0	246.0	155.0	246.0	332.0	215.0	279.0
	Evaporation rate, gpm/acre	Depends on salinity	2.50	2.50	2.50	2.49	2.49	2.49	2.57	2.27	2.57	2.71	2.50	2.64
	Area, acres		4.4	86.4	86.4	12.9	254.2	85.5	95.7	68.3	95.7	122.5	86.0	105.7
	Total mixed salts for disposal, tons/day		17	333	333	51	999	334	334	333	334	333	333	333
	Final depth of solids, ft	74.9 lb/ft^3 for 20 years	17.3	17.3	17.3	17.6	17.6	17.5	15.6	21.8	15.6	12.2	17.3	14.1
	No. of ponds required	Assume 12-ft pond depth	2.0	2.0	2.0	2.0	2.0	2.0	2.0	2.0	2.0	1.0	2.0	2.0
	Total pond acreage required		9	173	173	26	508	171	191	137	191	123	172	211
	Capital cost, installed, $M	$0.30M/acre	2.6	51.8	51.8	7.7	152.5	51.3	57.4	41.0	57.4	36.8	51.6	63.4
	Annual O&M cost, $M	1% times capital	0.0	0.5	0.5	0.1	1.5	0.5	0.6	0.4	0.6	0.4	0.5	0.6
Cost Summary														
Capital Cost, $M			10.7	204.2	129.0	14.2	304.9	128.5	134.6	118.2	134.6	114.0	128.8	140.6
Administrative, $M		25%	2.7	51.1	32.3	3.6	76.2	32.1	33.7	29.5	33.7	28.5	32.2	35.2
Contingency, $M														
	Project contingency	25%	2.7	51.1	32.3	3.6	76.2	32.1	33.7	29.5	33.7	28.5	32.2	35.2
	Process contingency	5%	0.5	10.2	6.5	0.7	15.2	6.4	6.7	5.9	6.7	5.7	6.4	7.0
Total Capital Cost for Scheme, installed, $M			16.6	316.6	200.0	22.0	472.6	199.2	208.7	183.2	208.7	176.6	199.6	217.9
Total Annual O&M Cost for Scheme, $M/year			2.5	49.2	24.9	2.5	50.2	24.8	24.9	31.2	28.2	24.7	28.1	25.0
Total Annual O&M Cost for Scheme, $/kgal			6.7	6.7	6.8	6.9	6.9	6.8	6.8	8.6	7.7	6.8	7.7	6.8
Annualized Capital, O&M Cost, $M/year		20 years @ 8% interest	4.1	81.4	45.2	4.8	98.3	45.1	46.2	49.9	49.4	42.7	48.4	47.2
Recovery of Water from Concentrate, %			98.4	98.4	96.9	95.4	95.4	96.9	96.5	97.8	96.5	95.2	96.9	96.0
MWs needed for Brine Concentrator			3.1	62.5	31.3	3.1	62.5	31.3	31.3	39.6	35.4	31.3	35.4	31.3

Table 5.2. Total Capital, Operating, and Annualized Costs for All Process Schemes and Cases



Scheme No.	Cost Category	Case No. (Concentrate Flow, Concentrate Salinity) 1 (1 MGD, 4000 mg/L)	2 (20 MGD, 4000 mg/L)	3 (10 MGD, 8000 mg/L)	4 (1 MGD, 12,000 mg/L)	5 (20 MGD, 12,000 mg/L)	6 (10 MGD, 8000 mg/L)	7 (10 MGD, 8000 mg/L)	8 (10 MGD, 8000 mg/L)	9 (10 MGD, 8000 mg/L)	10 (10 MGD, 8000 mg/L)	11 (10 MGD, 8000 mg/L)	12 (10 MGD, 8000 mg/L)
1A	Capital Cost, M$	16.6	316.6	200	22	472.6	199.2	208.7	183.2	206.7	176.6	199.6	217.9
	Unit Cap. Cost, $/gpd feed	16.6	15.83	20	22	23.6	19.92	20.87	18.32	20.87	17.66	19.96	21.79
	Operating Cost, $M/year	2.5	49.2	24.9	2.5	50.2	24.8	24.9	31.2	28.2	24.7	28.1	25
	Unit Oper. Cost, $/kgal feed	6.85	6.74	6.82	6.85	6.88	6.79	6.82	8.55	7.73	6.77	7.70	6.85
	Annualized Cost, $M/year	4.2	81.4	45.3	4.7	98.3	45.1	46.2	49.9	49.5	42.7	48.4	47.2
	Unit Ann. Cost, $/year/gpd	4.19	4.07	4.53	4.74	4.92	4.51	4.62	4.99	4.95	4.27	4.84	4.72
1B	Capital Cost, M$	17.5	306.1	189.5	21.7	438.2	185.2	190.1	207.5	193.6	199.1	187.4	192.9
	Unit Cap. Cost, $/gpd feed	17.5	15.305	18.95	21.7	21.91	18.52	19.01	20.75	19.36	19.91	18.74	19.29
	Operating Cost, $M/year	2.6	52.4	28.1	3	59.5	28.1	28.4	33.3	31.5	29.3	31.3	28.8
	Unit Oper. Cost, $/kgal feed	7.12	7.18	7.70	8.22	8.15	7.70	7.78	9.12	8.63	8.03	8.58	7.89
	Annualized Cost, $M/year	4.4	83.6	47.4	5.2	104.1	47.0	47.8	54.4	51.2	49.6	50.4	48.4
	Unit Ann. Cost, $/year/gpd	4.38	4.18	4.74	5.21	5.21	4.70	4.78	5.44	5.12	4.96	5.04	4.84
2A	Capital Cost, M$	12.1	194.1	144.7	21.8	377.9	168.9	136.2	142.8	139.7	144.1	139.3	127.5
	Unit Cap. Cost, $/gpd feed	12.1	9.705	14.47	21.8	18.895	16.89	13.62	14.28	13.97	14.41	13.93	12.75
	Operating Cost, $M/year	1.2	24.6	20.6	2.9	58.2	31.2	16.8	16.7	18.5	8.8	22.5	16.7
	Unit Oper. Cost, $/kgal feed	3.29	3.37	5.64	7.95	7.97	8.55	4.60	4.58	5.07	2.41	6.16	4.58
	Annualized Cost, $M/year	2.4	44.4	35.3	5.1	96.7	48.4	30.7	31.2	32.7	23.5	36.7	29.7
	Unit Ann. Cost, $/year/gpd	2.43	2.22	3.53	5.12	4.83	4.84	3.07	3.12	3.27	2.35	3.67	2.97
2B	Capital Cost, M$	15.8	240	190.6	28.4	511.8	227.9	182.6	140.6	189.1	137.2	183.8	158.4
	Unit Cap. Cost, $/gpd feed	15.8	12	19.06	28.4	25.59	22.79	18.26	14.06	18.91	13.72	18.38	15.84
	Operating Cost, $M/year	1.5	29.8	25.9	3.7	73.8	37.1	21	19.4	22.8	12.4	26.7	21
	Unit Oper. Cost, $/kgal feed	4.11	4.08	7.10	10.14	10.11	10.16	5.75	5.32	6.25	3.40	7.32	5.75
	Annualized Cost, $M/year	3.1	54.2	45.3	6.6	125.9	60.3	39.6	33.7	42.1	26.4	45.4	37.1
	Unit Ann. Cost, $/year/gpd	3.11	2.71	4.53	6.59	6.30	6.03	3.96	3.37	4.21	2.64	4.54	3.71
3	Capital Cost, M$	12.6	252	203.4	28.9	572.5	239.5	177.1	200	180.2	188.2	208.8	206.4
	Unit Cap. Cost, $/gpd feed	12.6	12.6	20.34	28.9	28.625	23.95	17.71	20	18.02	18.82	20.88	20.64
	Operating Cost, $M/year	1.1	21.8	17.8	2.5	49.4	27.7	14.6	13.1	16.3	5.3	19.6	13.7
	Unit Oper. Cost, $/kgal feed	3.01	2.99	4.88	6.85	6.77	7.59	4.00	3.59	4.47	1.45	5.37	3.75
	Annualized Cost, $M/year	2.4	47.5	38.5	5.4	107.7	52.1	32.6	33.5	34.7	24.5	40.9	34.7
	Unit Ann. Cost, $/year/gpd	2.38	2.37	3.85	5.44	5.39	5.21	3.26	3.35	3.47	2.45	4.09	3.47

5.2.2 Graphical Representation of Costs for Cases 1–5

Salinity levels and plant sizes for the five cases are summarized in Table 5.3.

Table 5.3. Salinity Levels and Plant Sizes for Cases 1–5

Case No.	TDS, mg/L	Plant Size, MGD
1	4000	1
2	4000	20
3	8000	10
4	12,000	1
5	12,000	20

Figure 5.1 shows the capital, operating, and annualized costs for the five cases. The costs vary greatly due to the range of plant sizes. A useful comparison of costs can be facilitated by examining unit costs. Figure 5.2 presents the costs shown in Figure 5.1 in terms of unit costs.

Unit costs for the five processing schemes are made up of contributions from each processing step of the processing scheme. The process step contributions are depicted in the bar charts. Figure 5.3 presents costs for Case 1 and each of the five processing schemes. Bar charts for Cases 2–5 are contained in Appendix 3, as Figures A3.1–A3.4. The unit capital costs represented in these figures refer only to the installed equipment costs. Administrative and contingency costs, such as those listed in Table 5.1 for Scheme 1A, are not included in the unit capital costs, as they do not apply to specific process steps.

5.2.3 Observations and Explanations for Cases 1–5

Cases 1–5 entail the same chemical composition but with three different salinities and three different plant sizes. The effects of the salinity and size variables, as represented by the different cases, result in the wide range of costs apparent in Figure 5.1. This range, in terms of capital cost, goes from less than $10M to almost $600M. The range in operating cost goes from about $2M/year to almost $75M/year.

The effect of salinity (TDS) on the costs is evident from a comparison of the unit costs shown in Figure 5.2. It is apparent that the unit capital costs of Cases 4 and 5 (salinity of 12,000 mg/L) are higher than the unit capital cost of Cases 1 and 2 (salinity of 4000 mg/L). For some of the process schemes, the effect is even more apparent on the unit operating costs. The unit annualized cost is more heavily influenced by the unit operating cost, and it also shows increased costs for each process scheme as a function of salinity.

The small effect of plant size on unit costs is shown by the slightly lower unit capital cost and unit annualized cost for Case 2 (20 MGD) relative to Case 1 (1 MGD) and for Case 5 (20 MGD) relative to Case 4 (1 MGD). While economies of scale can reduce costs for larger plants, the sizes considered are so large that multiples of each equipment step are required for the larger sizes. Consequently, most of any economy of scale is lost. Negotiations of costs with equipment and materials suppliers for large plants would likely reduce costs somewhat. This factor could not be quantified, however, and was not included in the analysis.

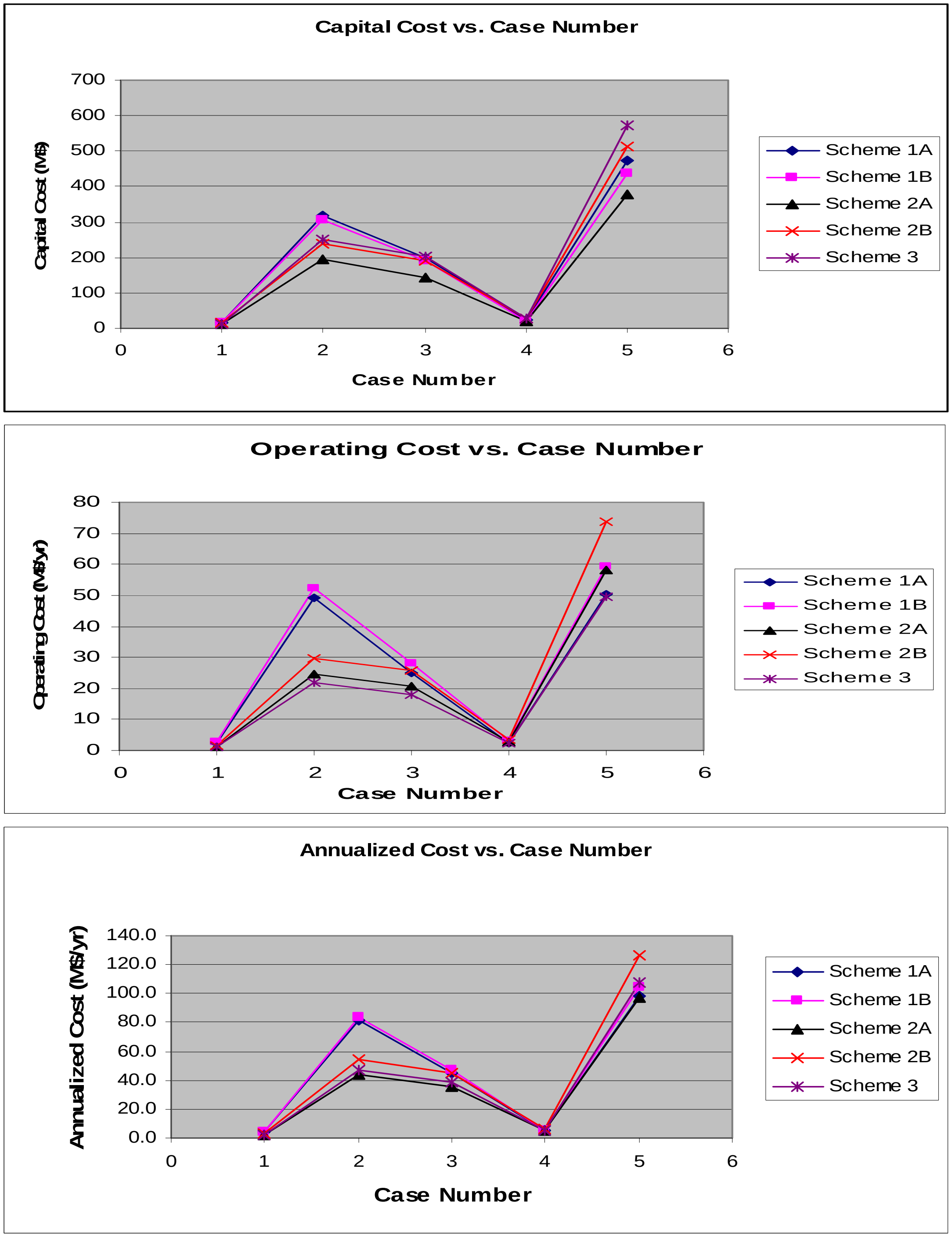


Figure 5.1. Capital, operating, and annualized costs for Cases 1–5.

Case Number:	1	2	3	4	5
Salinity, mg/L:	4000	4000	8000	12,000	12,000
Size, MGD:	1	20	10	1	20

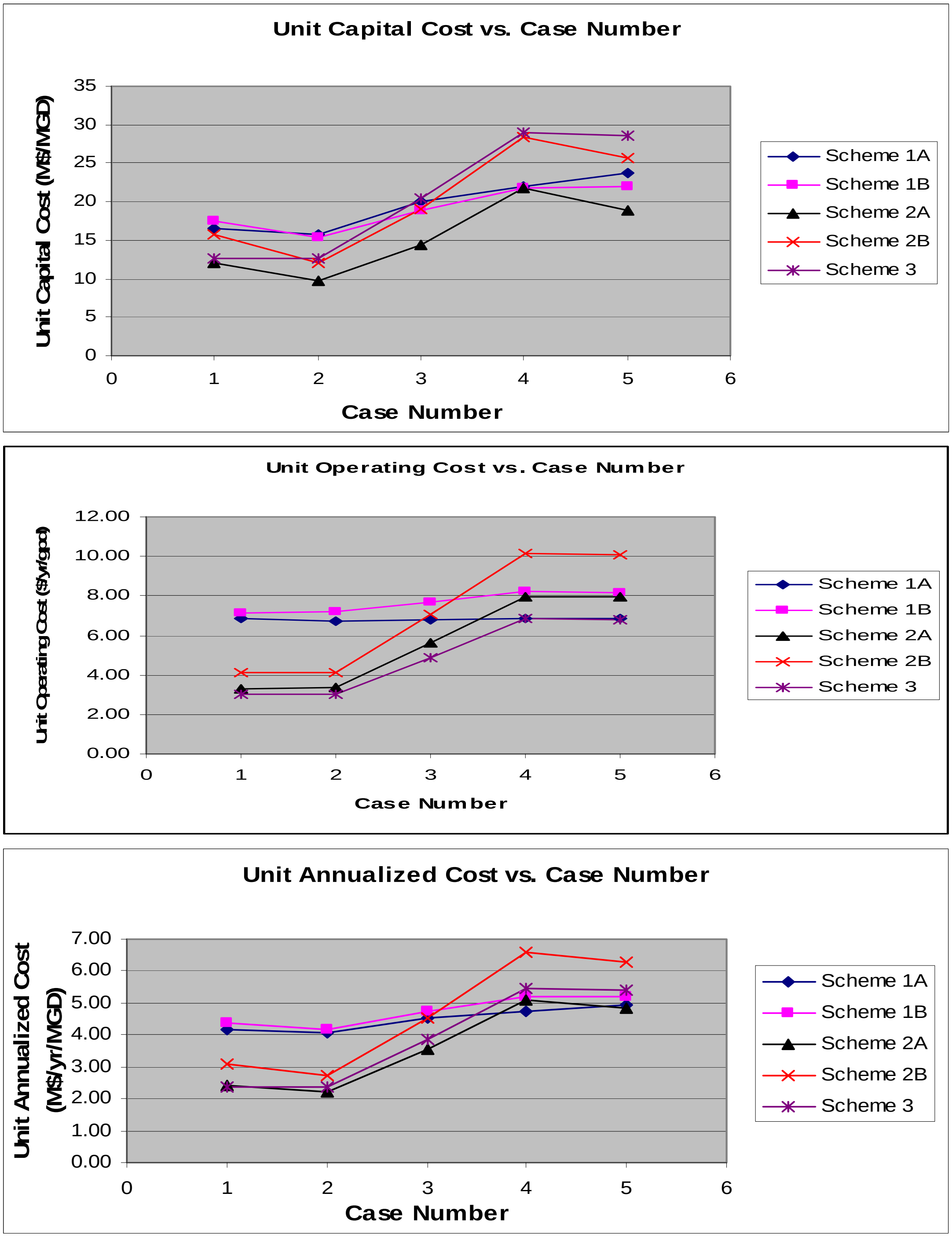
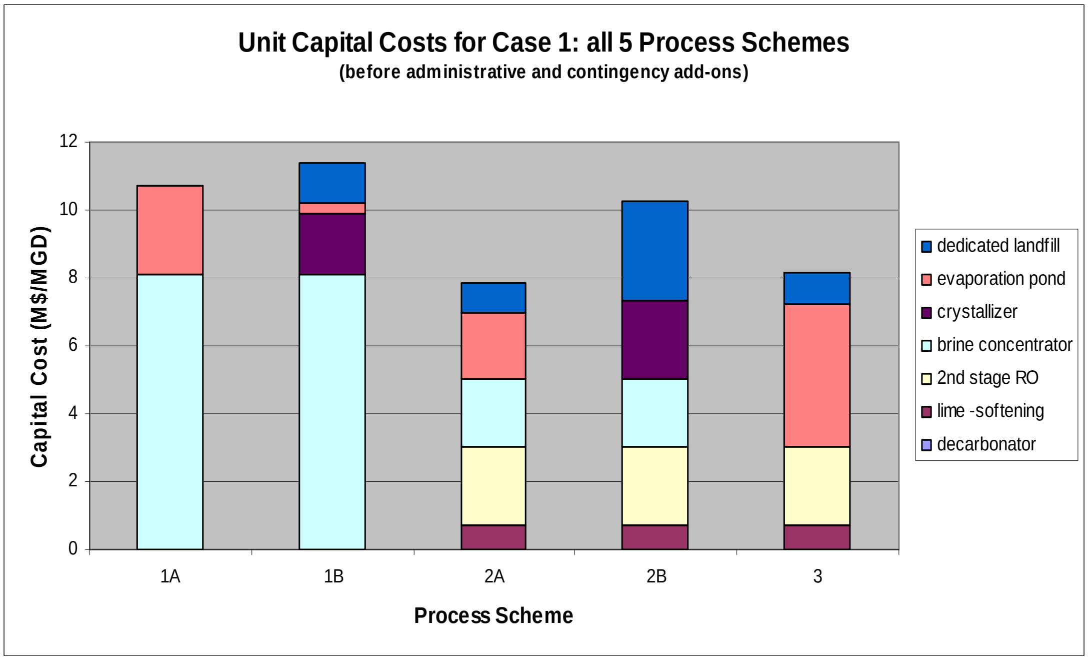


Figure 5.2. Unit capital, operating, and annualized costs for Cases 1–5.

Case Number:	1	2	3	4	5
Salinity, mg/L:	4000	4000	8000	12,000	12,000
Size, MGD:	1	20	10	1	20



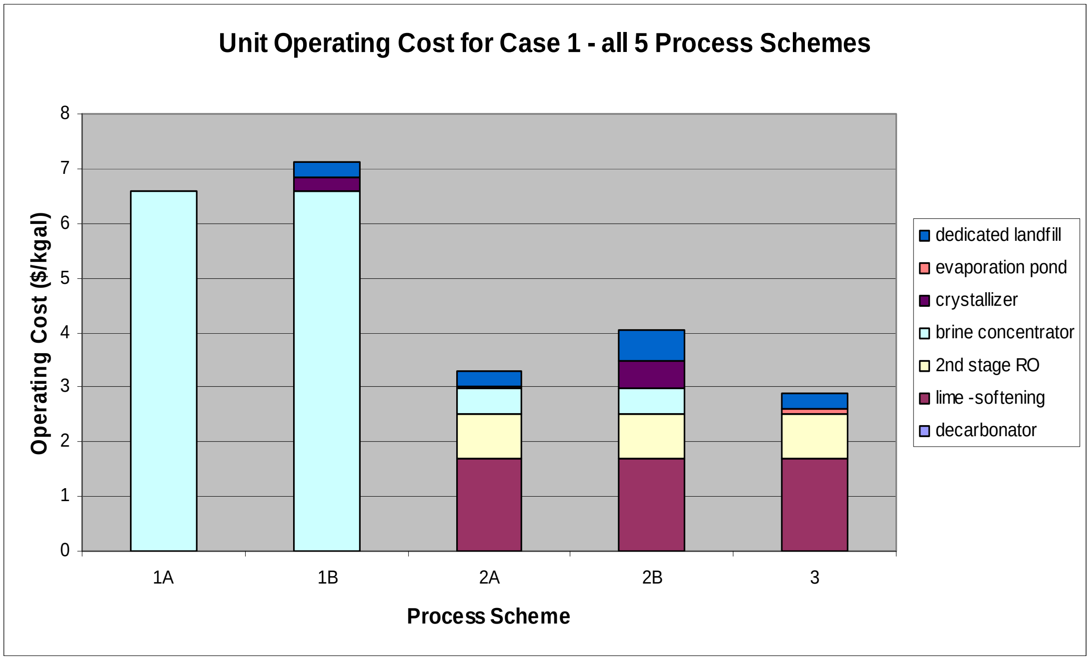


Figure 5.3. Process step contributions to unit capital and operating costs for Case 1.

Also evident from Figure 5.2 is the sensitivities of Schemes 2A, 2B, and 3 (the schemes using RO technology) to salinity. Curves for each of these schemes show significantly greater costs for high salinity (Cases 4 and 5) than for low salinity (Cases 1 and 2). By contrast, the curves for Schemes 1A and 1B are nearly horizontal, particularly for the unit operating and unit annualized costs. The relative insensitivities of Schemes 1A and 1B to salinity suggest that these older, more traditional ZLD schemes are more cost-effective and cost-competitive at higher feed salinities.

The effect of salinity on process costs is reflected in Table 5.4, which lists the high and low values for unit capital costs and unit operating costs for each process scheme. The values were taken from Table 5.2. Administrative and contingency costs were not included in the unit capital costs. Differences between the high and low cost values and the ratio of high to low cost values are provided to indicate the degree of variability in the values for each process scheme. The particular case giving the high or low value is indicated in parentheses. For instance, the high unit capital cost for Scheme 1A was in Case 5, with a value of $23.6/gpd.

Table 5.4. High and Low Values of Unit Capital and Operating Costs for Cases 1–5

	Unit Capital Cost, $/gpd (case no.)				Unit Operating Cost, $/kgal (case no.)			
Scheme	**High**	**Low**	**Difference**	**Ratio**	**High**	**Low**	**Difference**	**Ratio**
1A	23.6 (5)	15.8 (2)	7.8	1.5	6.88 (5)	6.74 (2)	0.14	1.0
1B	21.9 (5)	15.3 (2)	6.6	1.4	8.22 (4)	7.12 (1)	1.1	1.2
2A	21.8 (4)	9.7 (2)	12.1	2.2	7.97 (5)	3.29 (1)	6.06	2.4
2B	28.4 (4)	12 (2)	16.4	2.4	10.14 (4)	4.08 (2)	6.06	2.5
3	28.9 (4)	12.6 (1, 2)	16.3	2.3	6.85 (4)	2.99 (2)	3.86	2.3

Some observations based on Table 5.4 are the following:

- The variability in unit costs (both capital and operating) is less for Schemes 1A and 1B than for Schemes 2A, 2B, and 3.
- The variability in unit operating costs for Schemes 1A and 1B is very low.
- The high unit capital costs are associated with high salinity (Cases 4 and 5).
- The high unit operating costs are also associated with high salinity (Cases 4 and 5).
- The low unit capital costs and low unit operating costs are associated with low salinity (Cases 1 and 2).

High-salinity feedwater results in lower concentration factors for both brine concentrators (Schemes 1A and 1B) and in second-stage RO systems (Schemes 2A, 2B, and 3). This, in turn, leads to greater-sized follow-on process steps in all schemes and thus increased capital costs. The higher-salinity feedwater also results in greater amounts of solids produced in the lime softening steps (Schemes 2A, 2B, and 3) and from the crystallizers (Schemes 1B and 2B). This in turn results in larger landfill area requirements and associated capital costs. The high-salinity feedwater results in greater chemical requirements at the lime softening step (Schemes 2A, 2B, and 3). This leads to greater unit operating costs.

Another observation from Table 5.4 is the range of unit capital and unit operating costs associated with Cases 1–5. This is shown in Table 5.5.

Table 5.5. Range of Costs for Cases 1–5

Cost Type (all cases)	**High Value**	**Low Value**	**Ratio**
Unit Capital Cost, $/gpd	28.9	9.7	3.0
Unit Operating Cost, $/kgal	10.14	2.99	3.4

The ranges for both unit capital and unit operating costs from low to high values vary by a multiple of 3 over the salinities and size values considered. Note that the values are all high relative to that of typical first-stage RO capital and operating costs, which are typically less than $3/gpd and $1.5/kgal, respectively.

The bar charts of Figures 5.3 and of A3.1, A3.2, A3.3, and A3.4 in Appendix 3 provide additional observations. The largest process step contributions to unit capital and unit operating costs for the five process schemes are given in Table 5.6.

Table 5.6. Largest Process Step Cost for Each Processing Scheme, Cases 1–5

Process Scheme	**Step(s) with Largest Cost (no. of cases utilizing indicated process)**	
	Unit Capital Cost	**Unit Operating Cost**
1A	BC (3), EP (2)	BC (5)
1B	BC (5)	BC (5)
2A	EP (3), RO2 (2)	LS (5)
2B	LF (5)	LS (5)
3	EP (5)	LS (5)

Abbreviations: BC, brine concentrator; EP, evaporation pond; RO2, second-stage RO; LS, lime softener; LF, landfill.

Table 5.7 lists the major source of cost variability over the five cases for each of the five process schemes.

Table 5.7. Process Step Contributing Most to Cost Variability, Cases 1–5

Process Scheme	**Step(s) Contributing Greatest Variability**	
	Unit Capital Cost	**Unit Operating Cost**
1A	BC, EP	BC
1B	BC	BC
2A	BC, EP	LS
2B	BC, LF	LS
3	EP	LS

Abbreviations: BC, brine concentrator; EP, evaporation pond; LS, lime softener; LF, landfill.

From Tables 5.6 and 5.7, it can be seen that:

- Evaporation ponds and landfills, the final disposal step for ZLD processes, are frequently the major contributors to the unit capital cost.
- The major operating cost contributors are the brine concentrator energy cost (Schemes 1A and 1B) and lime softening chemical costs (Schemes 2A, 2B, and 3).
- The process steps that are most affected by feedwater salinity, meaning they have the most variability in costs from case to case, are:

  - For unit capital cost, brine concentrator, evaporation pond, and landfill
  - For unit operating cost, brine concentrator (due to the energy requirement) and lime softening (due to the chemical requirement)

5.2.4 Graphical Representation of Costs for Cases 6–12

The feedwater in cases 6–12 is at a TDS of 8000 mg/L for a plant size of 10 MGD. As described in Chapter 4 (Table 4.1), the composition of feedwater varies significantly from case to case. Thus, the only independent variable affecting process scheme costs in Cases 6–12 is the composition of the feedwater.

Figure 5.4 shows the capital, operating, and annualized costs for these seven cases. Figure 5.5 shows the corresponding unit costs for these cases. These curves allow comparison with the unit cost curves of Figure 5.2 for Cases 1–5.

Figure 5.6 shows the cost contribution of each processing step making up Scheme 1A for Cases 6–12. Bar charts for the other four processing schemes (1B, 2A, 2B, and 3) are given in Appendix 3 as Figures A3.5–A3.8. These figures use unit costs and allow comparisons with Figures 5.3 and A3.1–A3.4, which represent Cases 1–5.

5.2.5 Observations and Explanations for Cases 6–12

The effect of feedwater composition on costs is evident from the vertical movement illustrated in Figures 5.4 and 5.5. The most dramatic movement for unit capital cost is for Scheme 2B. The largest movement of the unit operation cost curves is for Schemes 2A, 2B, and 3, whose curves are quite similar. With operating costs dominating the annualized cost, the largest movement in the annualized cost curves is also for Schemes 2A, 2B, and 3, with the most movement associated with Scheme 2B.

Compared to Schemes 2A, 2B, and 3, each of which uses membrane technology, the older, traditional ZLD process schemes (1A and 1B) are less sensitive to composition. Their curves (Figures 5.4 and 5.5) are relatively flat from Case 6 through Case 12.

This same information is evident in more quantitative terms in Table 5.8.

Table 5.8 is similar to Table 5.4. The particular case giving the high or low value is indicated in parentheses. For instance, the high unit capital cost for Scheme 3 was in Case 6, with a value of $24.0/gpd.

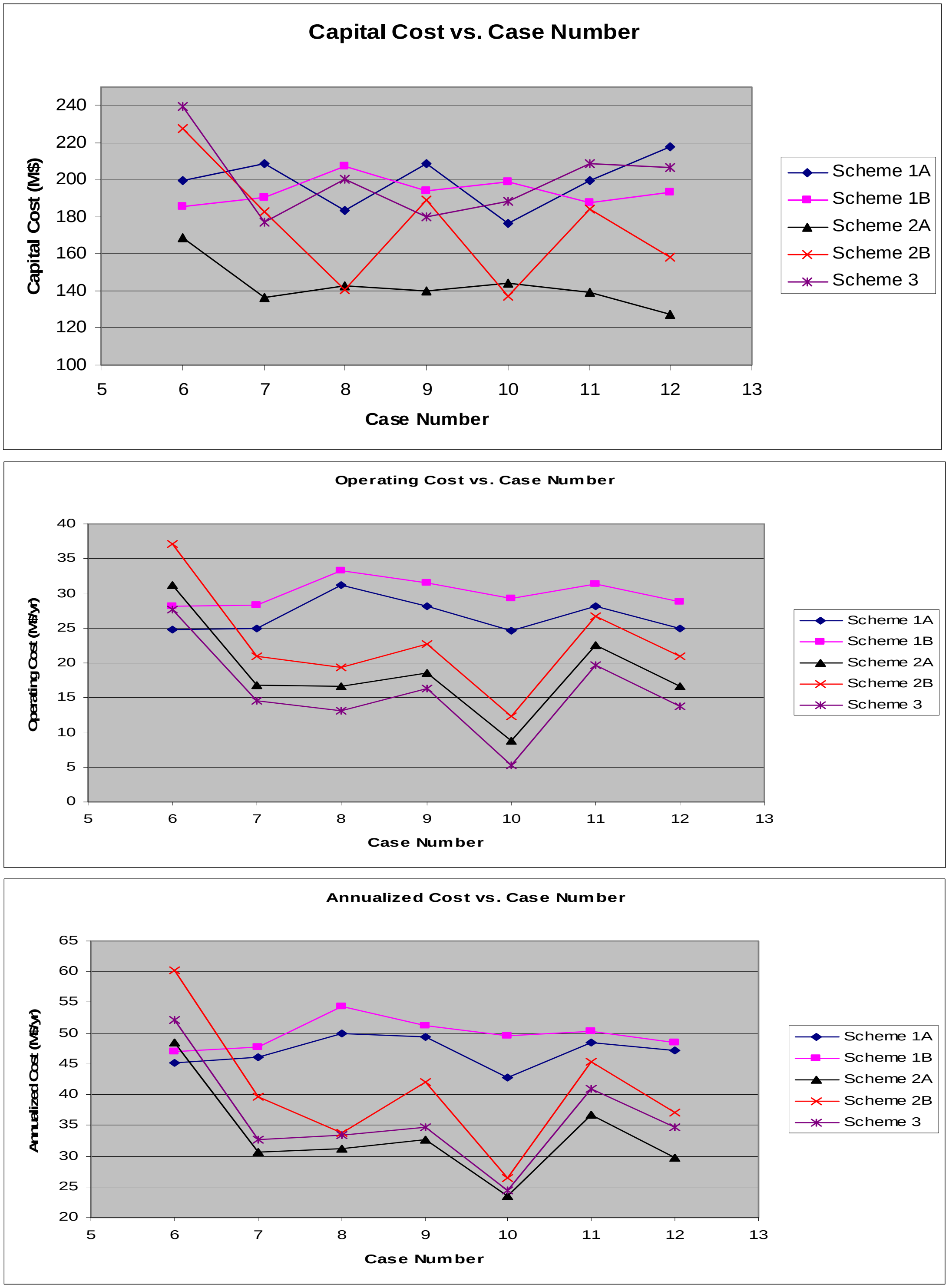


Figure 5.4. Capital, operating, and annualized costs for Cases 6–12.

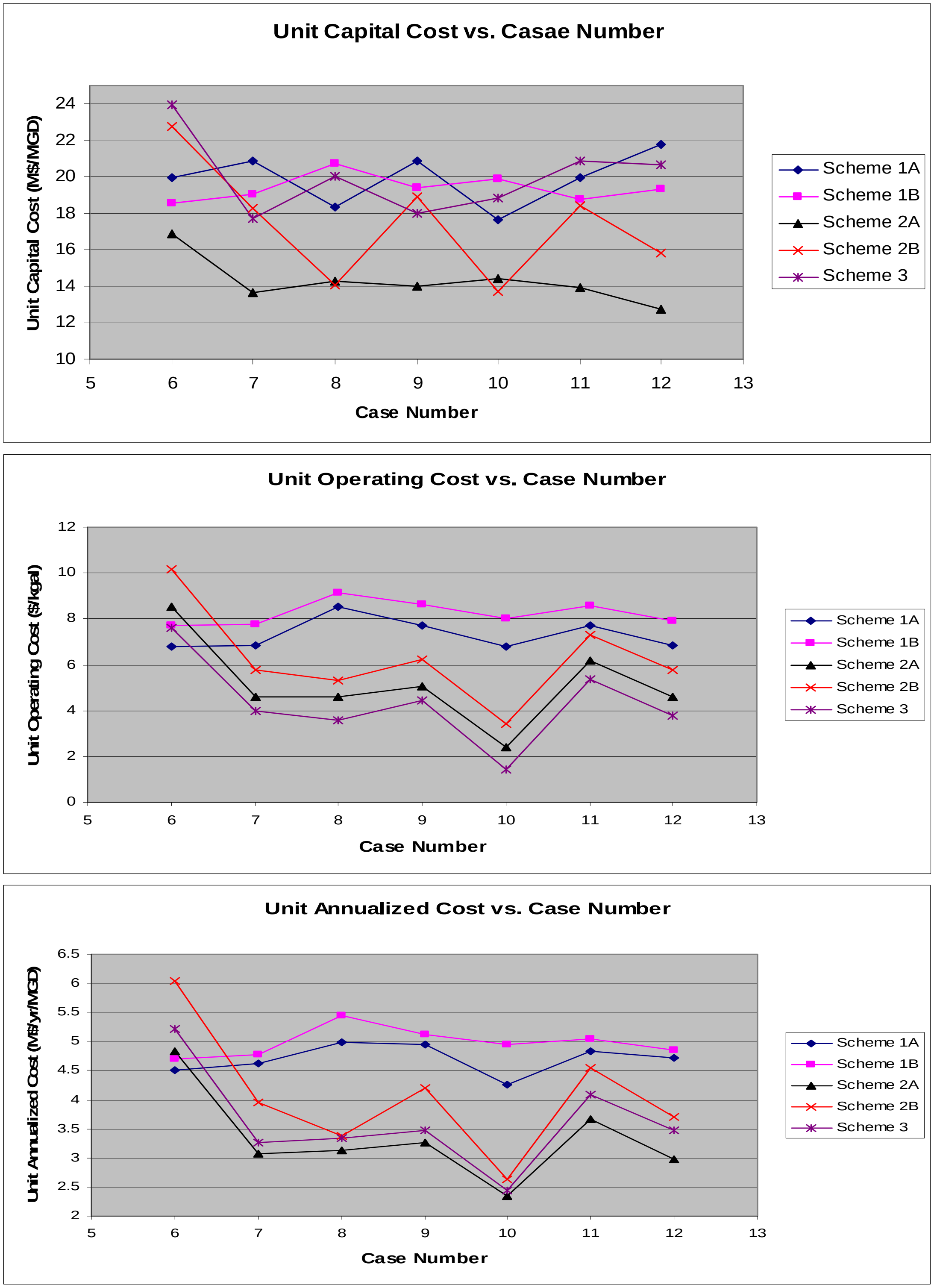


Figure 5.5. Unit capital, operating, and annualized costs for Cases 6–12.

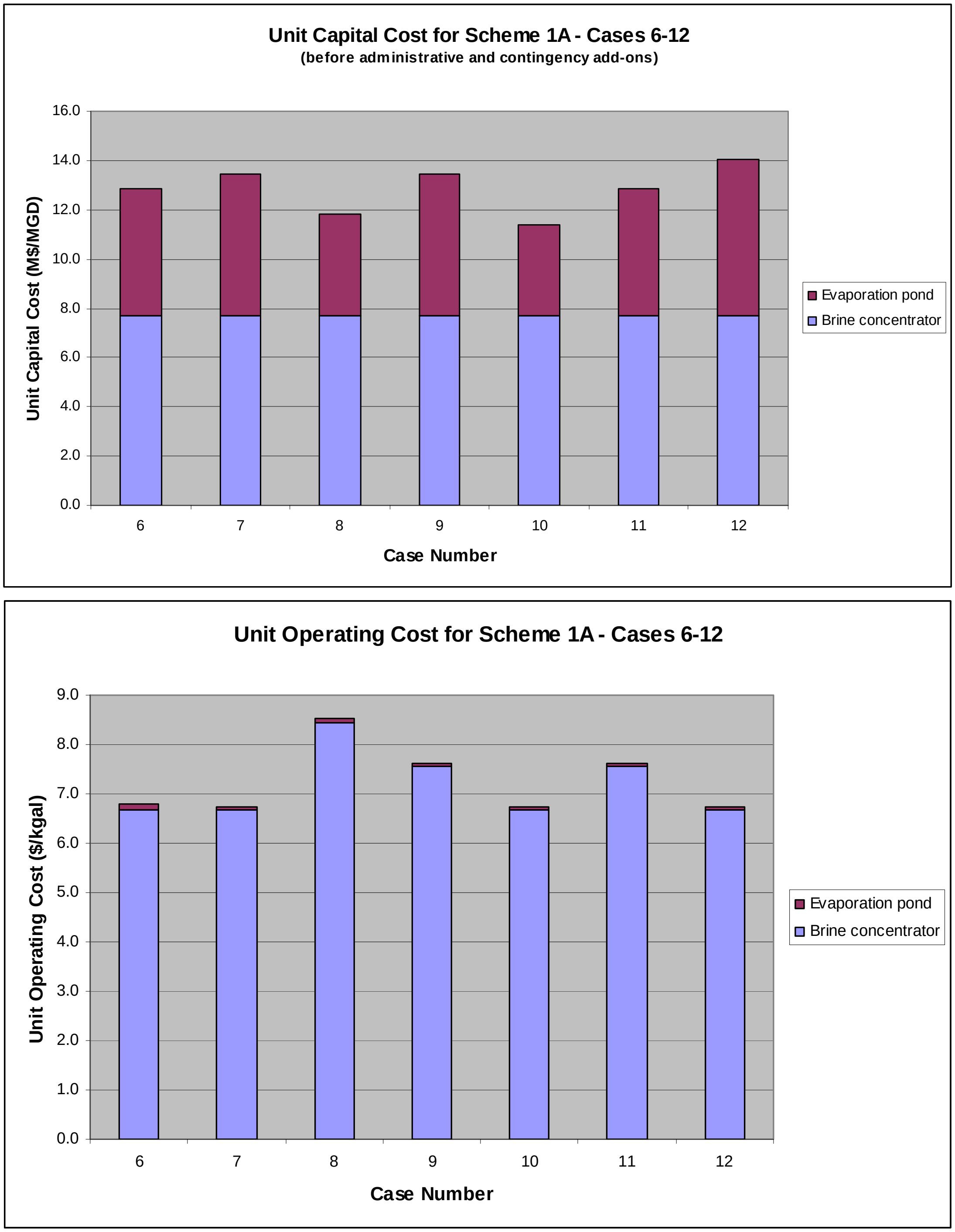


Figure 5.6. Unit capital and operating costs for Scheme 1A, Cases 6–12.

Table 5.8. High and Low Values of Unit Capital and Operating Costs for Cases 6–12

Scheme	Unit Capital Cost, $/gpd (case no.)				Unit Operating Cost, $/kgal (case no.)			
	High	**Low**	**Difference**	**Ratio**	**High**	**Low**	**Difference**	**Ratio**
1A	21.8 (12)	17.7 (10)	4.1	1.2	7.7 (9)	6.8 (10)	0.9	1.1
1B	20.8 (8)	18.5 (6)	2.3	1.1	9.1 (8)	7.7 (6)	1.4	1.2
2A	16.4 (6)	13.6 (7)	2.8	1.2	8.6 (6)	2.4 (10)	6.2	3.6
2B	22.8 (6)	13.7 (10)	9.1	1.7	10.2 (6)	3.4 (10)	6.8	3.0
3	24.0 (6)	17.7 (7)	6.3	1.4	7.6 (6)	1.5 (10)	6.1	5.1

Some observations based on Table 5.8 are the following:

- The variability in the unit costs (both capital and operating) is generally less for Schemes 1A and 1B than for Schemes 2A, 2B, and 3.
- The variability in unit operating cost for Schemes 1A and 1B is very low.
- The high unit capital and operating costs for Schemes 2A, 2B, and 3 are from Case 6.
- The low unit operating costs for Schemes 2A, 2B, and 3 are from Case 10.

Due to the range of feedwater compositions, the cases vary significantly as to the amount of chemicals needed for lime softening treatment in Schemes 2A, 2B, and 3. Case 6 requires the greatest amount of added chemicals and produces the most solids from the lime softening step for landfill. Case 10 has the lowest chemical requirement for the lime softening step and the least amount of solids from the lime softening step for landfill disposal. This effect is evident from Figure 5.5, Cases 6 and 10. While this offers partial explanation for the appearance of Cases 6 and 10 in Table 5.8, other factors come into play. Detailed explanations are provided for each case in Appendix 4.

Another observation from Table 5.8 is the range of unit capital and unit operating costs associated with Cases 6–12. This range is shown in Table 5.9.

Table 5.9. Range of Costs for Cases 6–12

Cost Type (all cases)	High Value	Low Value	Ratio
Unit Capital Cost, $/gpd	24	13.6	1.8
Unit Operating Cost, $/kgal	10.2	1.5	6.8

The range in both unit capital and unit operating costs from low to high values varies greatly. The greater variance in unit operating costs suggests that the operating cost is more sensitive to changes in feedwater composition than capital costs. The cost values are all high (with the one exception of the Case 10 unit operating cost) relative to that of typical first-stage RO capital and operating costs, which are typically less than $3/gpd and $1.5/kgal, respectively.

Examination of the bar charts of Figures 5.6 and Figures A3.5–A3.8 in Appendix 3 allows additional observations. The largest process step contributions to unit capital and unit operating costs for the five process schemes are given in Table 5.10.

Table 5.10. Largest Process Step Cost for Each Processing Scheme, Cases 6–12

Process Scheme	Step with Highest Cost (no. of cases)	
	Unit Capital Cost	Unit Operating Cost
1A	BC (7)	BC (7)
1B	BC (7)	BC (7)
2A	EP (7)	LS (7)
2B	LF (7)	LS (7)
3	EP (7)	LS (7)

Abbreviations: BC, brine concentrator; EP, evaporation pond; LS, lime softener; LF, landfill.

Table 5.11 lists the major source of cost variability over the five cases for each of the five process schemes.

Table 5.11. Process Step Contributing Most to Cost Variability, Cases 6–12

Process Scheme	Step Contributing Most to Cost Variability (no. of cases)	
	Unit Capital Cost	Unit Operating Cost
1A	EP (7)	BC (7)
1B	LF (7)	BC (7)
2A	EP (7)	LS (7)
2B	LF (7)	LS (7)
3	EP (7)	LS (7)

Abbreviations: BC, brine concentrator; EP, evaporation pond; LS, lime softener; LF, landfill.

From Tables 5.10 and 5.11, the following may be seen:

- The tabulated data are very clear for Cases 6–12, more so than for Cases 1–5 (as represented in Figures 5.6 and 5.7).
- While brine concentrators are the major unit capital cost item for Schemes 1A and 1B, in the other three schemes evaporation ponds and landfills (the final disposal step for ZLD processes) are the major contributors to unit capital cost.
- The major operating cost contributors are the brine concentrator energy cost (Schemes 1A and 1B) and lime softening chemical costs (Schemes 2A, 2B, and 3).
- The process steps that are most affected by the feedwater composition (those that have the most variability in costs from case to case) are:
  - For unit capital cost, evaporation pond and landfill
  - For unit operating cost, brine concentrator (due to the energy requirement) and lime softening (due to the chemical requirement)

5.2.6 Process Scheme Recoveries

Table 5.1 for Scheme 1A and Tables A2.1–A2.4 in Appendix 2 for Schemes 1B, 2A, 2B, and 3 list the recoveries attained by the various schemes. The recoveries are based on the concentrate that is feedwater to the ZLD processes. Since water is recovered in producing the concentrates, the overall recovery from groundwater to final treatment is much higher than the values listed. Table 5.12 summarizes the concentrate-based recoveries.

Table 5.12. Process Scheme Recoveries for the 12 Cases

Process Scheme	% Recovery for Case No.:											
	1	2	3	4	5	6	7	8	9	10	11	12
1A	98.4	98.4	96.9	95.4	95.4	96.9	96.5	97.8	96.5	95.2	96.9	96.0
1B	99.9	99.9	99.8	99.6	99.6	100	99.9	98.8	99.6	100	99.8	100
2A	97.5	97.5	94.9	92.3	92.5	94.2	95.9	97.7	95.9	96.8	95.9	95.8
2B	100	100	100	100	100	100	100	100	100	100	100	100
3	93.9	94.0	87.9	81.0	81.3	85.5	90.0	87.0	90.8	87.4	87.6	86.5

Observations from Table 5.12 include the following:

- The very high recoveries of Schemes 1B and 2B are due to the presence of a crystallizer as the final water recovery step, which takes the feed to mixed solids. In Scheme 1B there frequently is a small purge stream from the crystallizer due to high levels of calcium and/or magnesium chloride. In Scheme 2B, lime softening greatly reduces the calcium and magnesium levels such that no purge stream from the crystallizer is necessary.
- Schemes 1A and 1B have similar recoveries; both have brine concentrators as the last water recovery step. Brine from the brine concentrator goes on to evaporation ponds.
- Scheme 3 has the lowest level of recovery with second-stage RO as the final water recovery step. Brine from second-stage RO, which is larger in volume than brine from a brine concentrator (for the same feed), goes on to an evaporation pond.
- The concentrate recoveries range from 81–100%. Assuming that the concentrate used as feedwater to the ZLD processing schemes resulted from a first RO stage achieving 75% recovery, the 81% concentrate recovery represents an overall groundwater-to-final brine recovery of 95.3%.

5.2.7 Energy Requirements for Processing Schemes Utilizing Brine Concentrators

The bottom rows of Table 5.1 and Tables A2.1 and A2.3 (in Appendix 2) list the brine concentrator energy needs for the various processing schemes and cases. For Schemes 1A and 1B, where the entire concentrate is processed by the brine concentrator, and Cases 2 and 5, for which the flow is 20 MGD, the energy requirement is 62.5 MW. This is a tremendous energy requirement and will be a factor in considering the use of evaporative processing for large-scale desalination plants. For these same processing schemes and Cases 6–12, where composition varies for a flow of 10 MGD, the energy requirement ranges from 31.3–39.6.

Processing Scheme 2B also utilizes a brine concentrator, but only after concentrate volume reduction by a second-stage RO step. In this scheme the brine concentrator energy requirement over the entire set of cases is reduced by a factor of 8–15.

5.3 SUMMARY OF OBSERVATIONS AND EXPLANATIONS

The following section includes additional observations and explanations as well as a summary of those discussed earlier in this chapter.

There is a significant effect of salinity on process scheme cost. Costs developed are heavily dependent on salinity, as reflected in Figure 5.2. Increasing the salinity by a factor of 2 will, other factors aside, reduce the amount of volume reduction possible by second-stage

RO or a brine concentrator by a factor of 2. For the same water composition, increasing the salinity by a factor of 2 will result in twice as much chemical use at a lime softening step and will, over the entire process, produce twice as much solids. ZLD processing (and high-recovery processing) costs are generally lowest for low-salinity feedwater.

Process step performance is strongly dependent on salinity and composition. This is reflected in the bar charts of Figures 5.3 and 5.6 and the charts of Appendix 3. The dependence is particularly true of the steps for lime softening (chemical requirements), RO (recovery), and a brine concentrator (recovery, material of construction, and energy requirements), but also for evaporation ponds (net evaporation rate and solids accumulation rate) and a landfill (solids accumulation rate). Performance affects both the operating and capital costs of the given process step.

Process step capital costs are further affected by the process step performance of the previous step. The performance of one process step dictates the equipment size of the following process step.

Costs for Schemes 1A and 1B are relatively insensitive to salinity and composition. Schemes 1A and 1B, the earliest ZLD processing schemes historically used, are less sensitive to salinity (Figure 5.2) and composition (Figures 5.4 and 5.5) than are processing schemes utilizing membranes (Schemes 2A, 2B, and 3). This is primarily due to the strong sensitivity of the lime softening operating costs to salinity and composition, which are parts of Schemes 2A, 2B, and 3.

Process Schemes 1A and 1B are generally of higher cost than Schemes 2A, 2B, and 3. Despite the strong cost dependence of processing Schemes 2A, 2B, and 3 on composition, Schemes 1A and 1B are generally of higher cost (Figures 5.4 and 5.5). Processing Schemes 1A and 1B are also of higher cost, except at the highest salinity (Figure 5.2).

Process Schemes 1A and 1B may become cost-competitive at higher feed salinities. Figure 5.2 (for Cases 1–5) illustrates relatively flat cost curves for these process schemes with an increase in salinity, in contrast to the more significant cost increases with salinity for Schemes 2A, 2B, and 3.

There is a relatively small effect of process size on unit costs. This is reflected in Figure 5.2 and the results from the large size range of processing studied. The large equipment sizes resulted in use of multiple units (brine concentrators, crystallizers) and thus limited the economies of scale, other than what might be available through negotiation with the original equipment manufacturer.

The significant effect of chemical composition is primarily due to calcium and alkalinity. In the processing approach modeled, silica was not a factor. Silica is not a problem in brine concentrators due to it being precipitated or adsorbed onto the recirculating calcium sulfate in the seeded slurry evaporator (Schemes 1A and 1B). In Schemes 2A, 2B, and 3, silica is removed in the lime softening process due to the presence of adequate levels of magnesium.

The primary chemical composition variables that most influence cost are calcium and bicarbonate. The influence is primarily through the lime softening operating costs (chemical cost) and the corresponding solids disposal costs. The amount of lime added increases as alkalinity in excess of calcium increases. The amount of soda ash added increases as the amount of calcium in excess of alkalinity increases. The amount of solids produced increases as the calcium and bicarbonate levels increase and as the chemicals added (lime and soda ash) increase.

Thus, costs generally increase with hardness but are also a function of the relative amount of alkalinity. ZLD processing (and high-recovery processing) costs are generally lowest for low-hardness feedwater.

Simplified performance and cost analyses can have significant error. The sensitivities of performance and cost to salinity and composition suggest that simplified analyses of ZLD processing costs are subject to large errors. Each processing step may be affected by salinity and composition, and in different ways. The effect may increase costs in one step and decrease costs in another. A detailed analysis based on understanding how each processing step is affected by salinity and composition is required for meaningful cost projections on these complicated, complex processing systems.

Volume reduction prior to use of thermal evaporators significantly reduces energy costs and overall operating costs, in general. The use of low-energy volume reduction (second-stage RO) prior to use of high-energy thermal evaporators reduces costs over a wide range of conditions and is, in general, the recommended processing scheme for large flow systems. The one exception was Case 6, where high calcium levels and low bicarbonate levels resulted in very high softening costs, less volume reduction at the second RO step, and larger volumes going to the BC and evaporator steps. This exception again points out the need for a detailed process analysis for making decisions, even at a fairly preliminary level, for such complex systems.

High cost of chemical treatment and solids disposal. The greatest source of unit operating cost variability in processing Schemes 2A, 2B, and 3 is the lime softening step. The chemical costs associated with lime softening (and other possible interstage treatment of concentrate) and the high solids disposal costs associated with such treatment are highly dependent on feedwater composition.

Use of crystallizers was not favored. The processing schemes with crystallizers (Schemes 1B and 2B) were, in all cases, more expensive than the processes without a crystallizer step.

High cost of evaporation ponds and landfills. The study results illustrate the high capital cost impacts of evaporation ponds for brine disposal and landfills for solids disposal. Evaporation ponds and landfills were frequently the largest individual capital cost process step and the largest source of capital cost variability among the many process schemes and cases. Such costs have not been adequately represented in the literature or in previous studies.

The lowest operating cost processing scheme was usually Scheme 3. The processing scheme most consistently of lowest unit operating cost was Scheme 3, in which there was no thermal evaporation equipment used. This was also due to reliance on a greater evaporation pond area and the low operating cost of evaporation ponds.

The lowest capital cost processing scheme was usually Scheme 2A. The processing scheme most consistently of lowest unit capital cost was Scheme 2A, where volume reduction by lime softening and second-stage RO preceded further volume reduction by the brine concentrator.

The lowest unit annualized cost processing scheme was usually Scheme 2A. The processing scheme most consistently of lowest annualized cost was Scheme 2A, the system with lime softening, second-stage RO, and a brine concentrator. Scheme 3, without any thermal evaporative steps, had a high unit capital cost due to the relatively greater evaporation pond acreage. Thus, even with a low operating cost, processing Scheme 3 usually had higher unit annualized costs than Scheme 2B.

High recovery processing produced recoveries (of concentrate) ranging from 81–100%. Recoveries are predictable, depending on the type of processing scheme. The highest recoveries are associated with schemes processing to solids. The next highest recoveries are associated with schemes otherwise using evaporative processing steps. Finally, the lowest recovery is associated with membrane-only processing schemes that do not utilize any evaporative processing steps.

Energy requirements for brine concentrators can be significant. The tremendous energy requirement of over 60 MW for 20 MGD of concentrate, where the concentrate is first sent to a brine concentrator, will be a factor in considering this processing approach (Schemes 1A and 1B) for large-scale desalination plants. Use of volume reduction by second-stage RO processing significantly reduces the size and energy requirements for the brine concentrator used in processing Schemes 2A and 2B.

Operating costs are significant in these conventional ZLD processing schemes. There is a trade-off in operating costs among the processing schemes between high energy (Schemes 1A and 1B) and high chemical and solids disposal costs (Schemes 2A, 2B, and 3). In the high-salinity cases (Cases 4 and 5) of Figure 5.2, operating costs of Schemes 2A, 2B, and 3 can be comparable or higher than those of Schemes 1A and 1B. While operating costs for Schemes 2A, 2B, and 3 are generally less than those of Schemes 1A and 1B for a salinity of 8000 mg/L (Figures 5.4 and 5.5), some compositions (Case 6) can have higher operating costs.

CHAPTER 6

CHARACTERIZATION OF SELECTED HIGH-RECOVERY TECHNOLOGIES

6.1 INTRODUCTION

High-recovery systems were discussed in general terms in Chapter 3. In this chapter, three of the patented commercial systems are reviewed in more detail. These include:

- O'Brien & Gere's ARROW^{TM} technology
- New Logic's VSEP^{TM} technology
- EET's HEEPM^{TM} technology

As with all high-recovery and ZLD systems, these systems are more costly per unit of water produced than low-recovery systems and are not competitive in situations where low-recovery processing is adequate and feasible. None of these technologies has been implemented in a municipal setting and, indeed, the marketing focus in each case has been on other industries.

The characterization presented here is based on a review of the literature, conversations with the technology companies, modeling of the processes, and performance and cost projections provided by the companies.

The three companies were asked to evaluate different water qualities and flows. Some of the situations evaluated overlap. More specifically, the cases studied and the technologies evaluated are shown in Table 6.1.

Table 6.1. Cases Studied

Parameter[1]	Case 1 Concentrate	Case 10 Raw	Case 13 Raw	Case 13 Concentrate
Ca	731	75	65	426
Mg	355	36	43	280
Na	1266	2043	68	414
K	63	63	2.2	13
Ba	0	0	0.16	1.04
Sr	0	0	1	6.6
CO_2	0	0	2.1	2.6
HCO_3	928	5682	348	2012
SO_4	3564	6	74	654
Cl	1111	95	56	368
F	0	0	0.9	5.8
NO_3	0	0	21	113
SiO_2	22	0	25	157
pH	7	7	8	8
TDS	8000	8000	705	4452

[1] All results are in milligrams per liter.

Case 13 is a new case not previously considered in this report. It was developed for a different project, but the information is useful here. Cases 1 and 10 correspond to the previously mentioned Cases 1 and 10.

Because the performance and cost evaluations were requested at different times and for slightly different insights, the flow rates studied for the different technologies differ somewhat. There are cases that overlap and allow some comparisons of performance and general cost levels. The purpose of the effort, however, was not to compare costs but to get a representative indication of performance, cost factors, and general cost levels. No effort was made to standardize cost factors or costs for specific equipment items. No consideration was given to the use of bypass to meet product water goals.

Table 6.2 lists the flow rates used as the basis for the various cost estimates by technology.

Table 6.2. Flows Rates at which Technologies Were Evaluated

Technology	Case 1 Concentrate	Case 10 Raw	Case 13 Raw	Case 13 Concentrate
ARROW™	1.0 MGD	1.0 MGD	2.0 MGD	-----
VSEP™	1.0 MGD	-----	1.0 MGD	-----
HEEPM™	44.2 gpm	1.0 gpm	1.0 gpm	86.1 gpm

6.2 O’BRIEN & GERE’S ARROW^{TM} TECHNOLOGY

6.2.1 Process Description

The ARROW technology (ARROW is an acronym for advanced reject recovery of water) was developed by Dr. Riad Al-Samadi of Advanced Water Solutions, Burlington, ONT, Canada. It is a proprietary, patented technology that has been licensed and commercialized by O’Brien & Gere of Syracuse, NY (O’Brien & Gere, 2007).

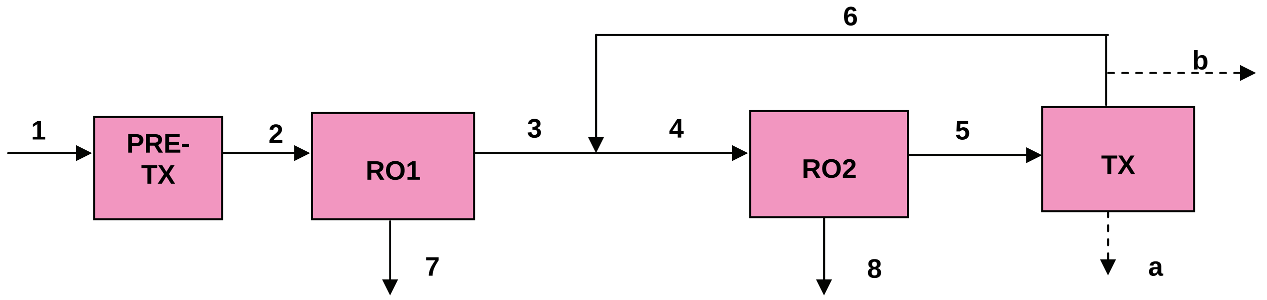
The unique aspect of the ARROW technology is the location of the treatment step that allows high-recovery processing. Instead of the typical front-end or interstage treatment, the ARROW technology places the treatment step at the back end after the second membrane unit. The back-end treated water is then recycled, most typically, to the interstage site. Although the feasibility of this processing scheme is not obvious, modeling of the process reveals that this design can allow very high recovery operation. The primary benefits of this processing configuration are that the size of the stream to be treated is smaller in volume and the process has a smaller footprint. The result is a savings in capital cost.

A general process flow diagram is provided in Figure 6.1. Pretreatment prior to the first RO stage is similar to that routinely done in other brackish RO processes. Concentrate from RO1 is blended with the recycled treated water from the back-end treatment step. The blend makes up the feed to the second RO stage (RO2). RO2 is a high-pressure system that uses a seawater RO membrane for which water recovery is limited only by osmotic pressure. Concentrate from RO2 is treated to remove sparingly soluble salts and silica, which limit recovery. Treatment may be chemical treatment and/or IX. If silica is not a problem, the treatment is usually by sodium zeolite IX, except for larger systems, where the cost of the zeolite softener is too expensive. When silica is a concern, and in general when IX is not used, chemical treatment with hydroxide and possibly carbonate is typically used. Treated water is recycled back to a point between the two RO stages.

While the treatment process removes sparingly soluble salts and silica, there is no sink for highly soluble ions and salts, predominantly sodium salts. To avoid sodium salt buildup in the system, sufficient water is removed either via decanted solids from a chemical precipitation process (represented by stream a) or from a purge stream (represented by stream b) when the treatment is by IX. In this latter case, the size of the purge is typically 2–3% of the recycle stream. The waste stream from chemical precipitation is typically 2–3 wt% slurry.

The back-end treatment can also be applied in a single-membrane stage configuration where treated water is recycled to mix with the feed to the RO step. This configuration may be preferred for small applications.

Recoveries are frequently greater than 95% and have been greater than 97% in some applications. Various piloting operations are underway.



Stream #/Process Step	Description
1	Raw water
PRE-TX	Pretreatment as needed that may include MF, UF, micromedia filtration, aeration, GAC, acid, antiscalant
2	Feed to RO1
RO1	First RO stage; low-pressure RO using a brackish RO membrane
3	Concentrate from RO1
4	Feed to RO2; includes concentrate from RO1 plus recycle stream from treatment step
RO2	Second RO stage; high-pressure RO using a seawater RO membrane
5	Concentrate from RO2
TX	Treatment step: may include chemical precipitation and/or IX
6	Treated water from treatment step; recycled to mix with concentrate from RO1
7	Product from RO1
8	Product from RO2
a	Decant off of tank bottoms if treatment is chemical precipitation
b	Purge stream of 2–3% if treatment is IX

Figure 6.1. General process flow diagram of ARROW^{TM} technology.

6.2.2 Processing Costs

O’Brien & Gere were asked to consider how they would process three different water quality cases and to assign capital and operational costs to these processes. These three cases are presented above in Tables 6.1 and 6.2.

Figures 6.2 and 6.3 represent the process schematics supplied by O’Brien & Gere for these three cases. Table 6.3 presents parameters and costs associated with these cases.

Table 6.3 includes the water quality and flow rates, previously shown in Table 6.1, along with the operational parameters and costs developed by O’Brien & Gere for these cases. The capital cost was provided for equipment alone, equipment installed in an existing building, and a greenfield situation. Operating cost is given in terms of dollars per kilogallon and annual dollars. Energy requirements were estimated based on typical brackish and seawater RO system energy requirements. The costs provided reflect the costing practices and approach taken by O’Brien & Gere, and no effort was made on the study author’s part to standardize the costs and approach to costing taken by them or by the other system providers discussed in the following report sections. From discussions with O’Brien & Gere, costs not included have been estimated as follows:

- Contingency: 5% of equipment capital cost
- Engineering: 15% of equipment capital cost
- Administration: 3% of equipment capital cost

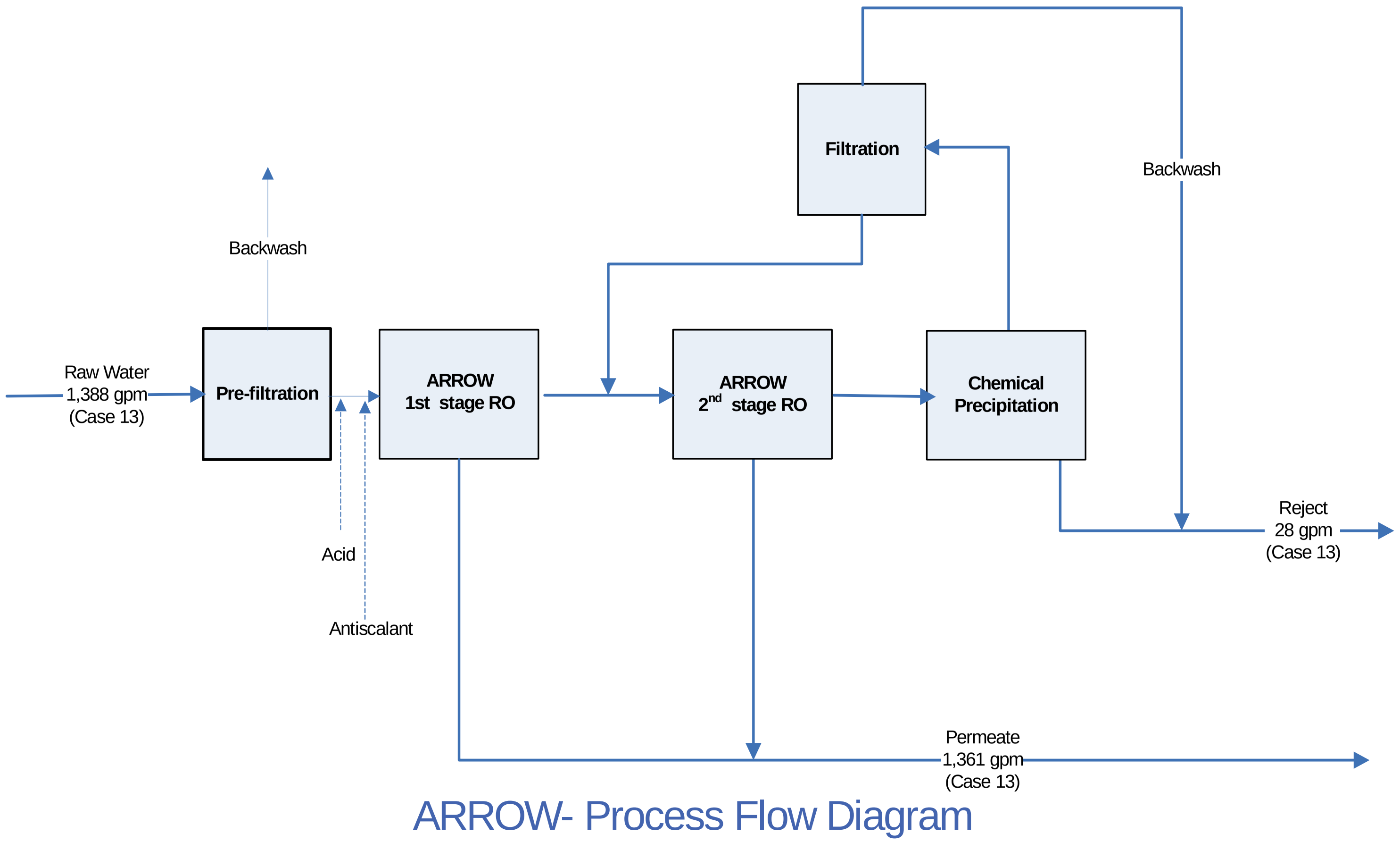


Figure 6.2. Process flow diagram for ARROW^{TM}, Cases 1 and 13.

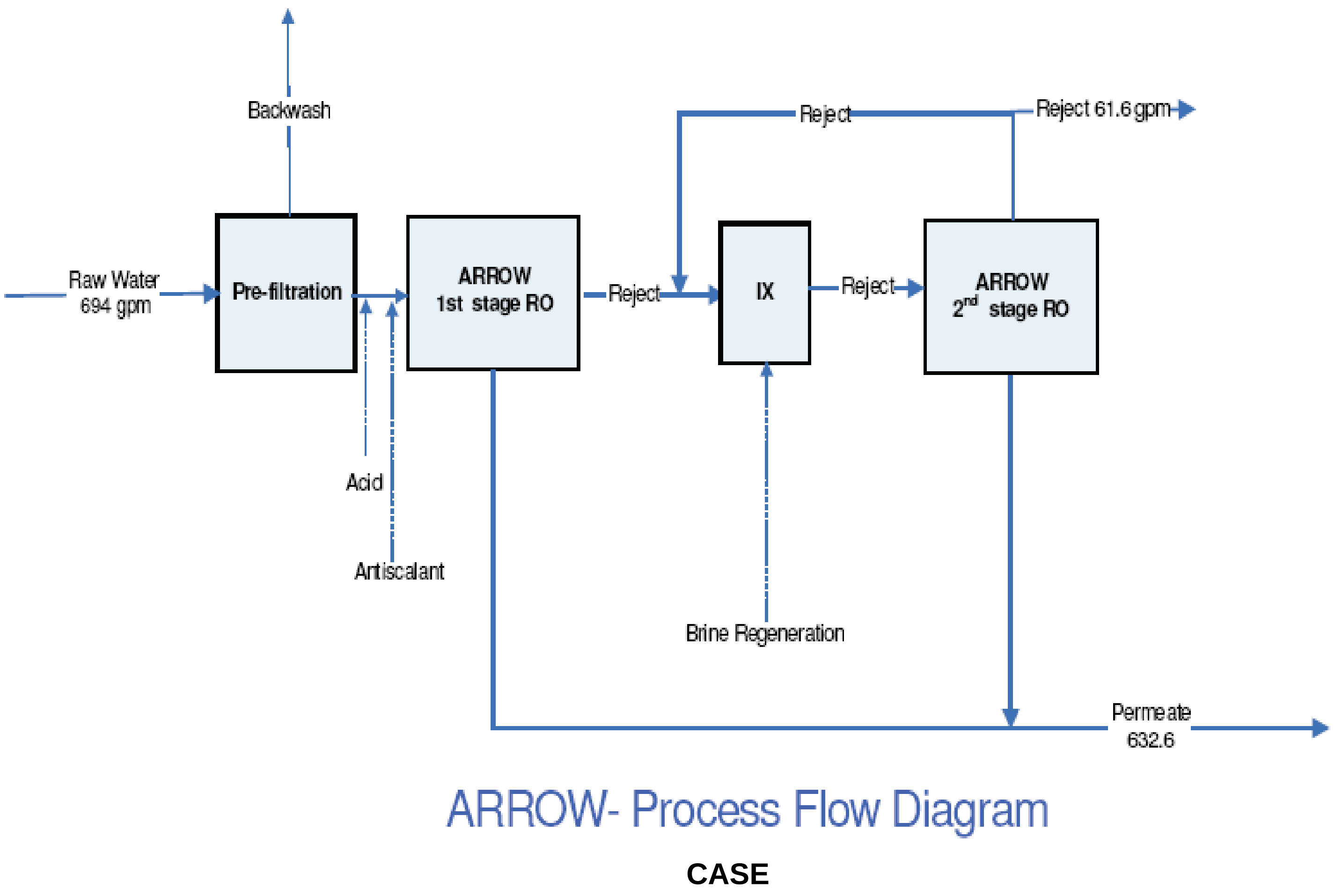


Figure 6.3. Process flow diagram for ARROW^{TM}, Case 10.

Table 6.3. Parameters and Costs for ARROW^{TM}

Constituent or Parameter	Case 1 Concentrate	Case 10 Raw Water	Case 13 Raw Water
Ca, mg/L	731	75	65
Mg, mg/L	355	36	43
Na, mg/L	1226	2043	68
K, mg/L	63	63	2.2
Ba, mg/L	0	0	0.16
Sr, mg/L	0	0	1
CO_2, mg/L	0	0	2.1
HCO_3, mg/L	928	5682	348
SO_4, mg/L	3564	6	74
Cl, mg/L	1111	95	56
F, mg/L	0	0	0.9
NO_3, mg/L	0	0	21
SiO_2, mg/L	22	0	25
pH	7	7	8
TDS, mg/L	8000	8000	705
Other Parameters			
Nominal feed flow, MGD	1	1	2
Feed flow, gpm	694	694	1388
Recovery, %	90	91	98
Feed TDS, mg/L	8000	8000	705
Permeate (product) flow, gpm	626	633	1361
Permeate (product) TDS, mg/L	372	376	54
Concentrate (waste) flow, gpm	68	62	28
Concentrate (waste) TDS, mg/L			
Energy requirement, kWh/kgal	6.21*	6.4*	5.1*
Time operated (h/day)	24	24	24
Capital cost, M$	1.55	2.1	2.95
Unit capital cost, M$/MGD	1.55	2.1	1.48
Capital cost (installed, existing bldg.)	2.65	3.5	5
Capital cost (installed, new bldg.)	4.65	6.3	8.85
Operating cost, $/kgal	1.34	1.54	1.55
Operating cost, M$/year	0.44	0.51	1.11
Cost of energy, $/kWh	0.07	0.07	0.07

*Estimate based on 4 kWh/kgal for low-pressure RO stage or 9 kWh/kgal for high-pressure RO stage

6.2.3 Miscellaneous Information for ARROW^{TM}

The recoveries for Cases 1 and 10 are based on an initial TDS feed level of 8000 mg/L, the level of a hypothetical concentrate from a brackish RO system. If the ARROW^{TM} technology had been used on an original feed of much lower TDS, the recoveries would be much higher, as reflected in the 98% recovery estimated for Case 13.

An interesting aspect of high-recovery systems with recycling is the attainment of high recovery levels where the recovery levels from the individual membrane steps are lower than would be possible without recycling.

As an illustration of this, Table 6.4 lists the recoveries of RO1 and RO2 and the total recovery of the ARROW^{TM} systems and contrasts them with the overall recovery possible with the same RO steps when recycling is not employed.

Table 6.4. Overall Recovery for the ARROW^{TM} System Compared to Recovery without Recycling

	% Recovery			
Case	**RO1**	**RO2**	**Without Recycle***	**ARROW^{TM}**
1	50	51.6	75.7	90
10	50	51.6	75.7	91
13	75	47.9	89	98

*Recovery without recycle was calculated as RO1 + (1 – RO1) × RO2.

ARROW^{TM} systems are either skid mounted (for smaller systems) or provided in packaged form. A footprint for a 600-gpm system is estimated to be approximately 50 ft by 20 ft.

The unique aspect of the ARROW^{TM} technology is the back-end treatment of the concentrate and recycle of the treated concentrate back into the system, allowing high recovery to be achieved using smaller equipment. The process also has inherent flexibility in balancing the first-stage and second-stage recoveries to mitigate scale formation.

6.3 NEW LOGIC'S VSEP^{TM} TECHNOLOGY

6.3.1 Process Description

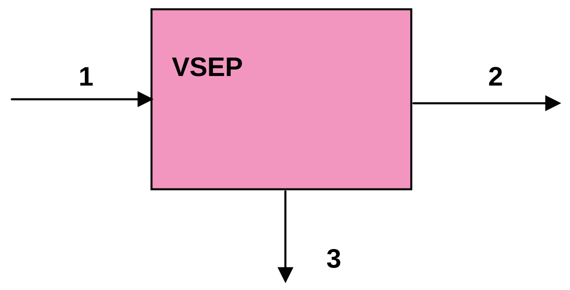
The name VSEP^{TM} stands for vibratory shear enhanced process (New Logic, 2007). The process membrane module, shown in Figure 6.4, is exceedingly simple and involves only a flat sheet membrane module mounted in a mechanism that vibrates the module. The high sheer at the membrane–solution interface minimizes the effects of sparingly soluble salts, silica, and foulants on the membrane and thus on membrane performance. Sparingly soluble salts and silica are allowed to precipitate, and thus high recoveries are attained without pretreatment or treatment. The process flow diagram of Figure 6.5 is simply for the membrane module. No chemicals are required for the process. Due to mechanical considerations, the individual module size is limited to flows of up to 60 gpm; however, many individual modules are easily incorporated into a multi-MGD processing scheme.

New Logic developed the technology in the 1980s, and their marketing focus and applications have been for difficult-to-separate solutions, such as food waste streams and oil–water mixtures.

Figure 6.4. A standard VSEP$^{\text{TM}}$ module.

More recently, a successful potable water pilot test was completed at Big Bear Ski Resort in California (Lozier, 2007). The VSEP$^{\text{TM}}$ technology is not cost-competitive for a first-stage (low-recovery) desalination step and thus is used to treat concentrate or waste from a first-stage RO or EDR system, without the need for chemical treatment or processing of the concentrate or waste prior to processing.

As solutions and TDS become more concentrated, the osmotic forces increase until precipitation of a salt occurs. Precipitation reduces the TDS level and thus the osmotic force. Processing using conventional high-pressure RO pumps can proceed to recoveries beyond what spiral-wound RO units would achieve with the same pressures. Recovery in an individual VSEP$^{\text{TM}}$ unit may become hydrodynamically limited at very high recoveries, as so much water is removed that the velocity of concentrate in the flow path is reduced to a very low level. In this case, the concentrate from the VSEP$^{\text{TM}}$ unit may be fed to a smaller, second VSEP$^{\text{TM}}$ unit for additional recovery.



Stream #/Process Step	Description
1	raw water
VSEP	VSEP unit: allows precipitation to occur; high shear minimizes effects of solids on performance
2	product
3	concentrate comprised of dissolved and suspended solids

Figure 6.5. General process flow diagram of VSEP^{TM} technology.

The flat sheet membrane or module packaging arrangement involving stainless steel pressure vessels allows for higher pressure and higher osmotic force operation. Current commercial equipment can process waters up to the limit of sodium sulfate precipitation (typically around 150,000 mg/L of TDS). Higher-pressure units under development will be able to operate beyond this limit.

The VSEP^{TM} system has a small footprint relative to other membrane systems, with an individual module skid measuring 4 ft by 5 ft.

6.3.2 Processing Costs

New Logic was asked to consider how they would process two different water quality cases and to assign capital and operational costs to these processes. These two cases are presented in Tables 6.1 and 6.2 above. Case 13 is the same as that studied for the ARROW^{TM} technology, except that the flow rate here is 1 MGD instead of 2 MGD. Case 1 is identical to that studied for the ARROW^{TM} technology.

Table 6.5 presents the costs associated with these cases. Table 6.5 includes the water quality and flow rates previously shown in Table 6.1, along with operation parameters and costs developed by New Logic for these cases. The capital cost was provided for equipment alone. VSEP^{TM} systems come as a complete system; the only additional items would be an equalization tank on the front end and collection tanks on the back end.

Units are skid mounted or packaged depending on system size. The major operating cost is membrane replacement, with individual membrane modules lasting 4–5 years and costing $65,000. As with the other system providers, the costs provided reflect the costing practices and approach taken by New Logic, and no effort was made on the study author's part to standardize the costs and approach to costing taken by the company. Installation is simplified, as each membrane module is supplied on a skid along with cleaning, automation, and control elements. From discussions with New Logic, installation costs can range from 10–40% of the equipment costs, depending on the application, industry, and site. The costs provided are considered to be complete.

6.3.3 Miscellaneous Information for VSEP^{TM}

In contrast to the two-stage (two desalination steps) ARROW^{TM} and HEEPM^{TM} technologies presented in this chapter, the VSEP^{TM} system is presented here as a single-stage system. The recoveries given are thus single-stage recoveries. Higher recoveries are possible by employing a second VSEP^{TM} stage. The recoveries estimated in Table 6.5 are thus considered conservative. As previously mentioned, the high recoveries achieved in a single-membrane unit can be limited hydrodynamically prior to reaching osmotic force limits. In this case, the brine from the initial VSEP^{TM} unit may be further processed in a small, second-stage VSEP^{TM} unit to obtain very high net recoveries. In this arrangement the system is still without pretreatment or interstage treatment.

Table 6.5. Parameters and Costs for VSEP^{TM}

Constituent or Parameter	Case 1 Concentrate	Case 13 Raw water
Ca, mg/L	731	65
Mg, mg/L	355	43
Na, mg/L	1226	68
K, mg/L	63	2.2
Ba, mg/L	0	0.16
Sr, mg/L	0	1
CO_2, mg/L	0	2.1
HCO_3, mg/L	928	348
SO_4, mg/L	3564	74
Cl, mg/L	1111	56
F, mg/L	0	0.9
NO_3, mg/L	0	21
SiO_2, mg/L	22	25
pH	7	8
TDS, mg/L	8000	705
Other Parameters		
Nominal feed flow, MGD	1	1
Feed flow, gpm	700	700
Recovery, %	80[1]	85[1]
Feed TDS, mg/L	8000	705
Permeate (product) flow, gpm	560	595
Permeate (product) TDS, mg/L	300[2]	300[2]
Concentrate (waste) flow, gpm	140	105
Concentrate (waste) TDS, mg/L		
Energy requirement, kWh/kgal	13.4	11.6
Time of operation (h/day)	22	22
Capital cost, M$	4.91	4.35
Unit capital cost, M$/MGD	4.91	4.35
Capital cost (installed, existing bldg.)	2.65	5
Capital cost (installed, new bldg.)	4.65	8.85
Operating cost, $/kgal	5.34	4.43
Operating cost, M$/year	1.44	1.26
Cost of energy, $/kWh	0.04	0.04

[1] Recovery is for a single VSEP^{TM} stage.

[2] Assumed; value is dependent on membrane used; no membrane was specified in the performance estimate.

Most VSEP^{TM} applications in the past have been for industries with difficult-to-treat wastewaters that frequently contain substantial organic fractions. The application to “cleaner” waters with low organic levels is a more recent one. A potable water pilot system utilizing a VSEP^{TM} unit after a standard brackish RO unit has recently achieved recoveries of up to 92% operating directly on the first-stage concentrate. This type of processing arrangement is more typical for a relatively low TDS feed. As an illustration of the recoveries possible in this situation, Table 6.6 calculates the total recovery possible as a function of an initial brackish RO stage and a second VSEP^{TM} stage.

Table 6.6. Overall Recoveries Possible in a VSEP^{TM}-Based Processing System

VSEP^{TM} Recovery in Stage 2	% Recovery from Brackish RO with Stage 1 Recovery of:		
	65%	**75%**	**85%**
80%	93	95	97
85%	94.75	96.25	97.75
90%	96.5	97.5	98.5
95%	98.25	98.75	99.25

The VSEP^{TM} can be used with any membrane: RO, NF, UF, or MF. Applications in addition to the high recovery discussed include use of VSEP^{TM} with an NF membrane to remove calcium and to avoid the osmotic pressure limit on an RO system. The high recovery of the NF system is possible because it allows calcium salts to precipitate.

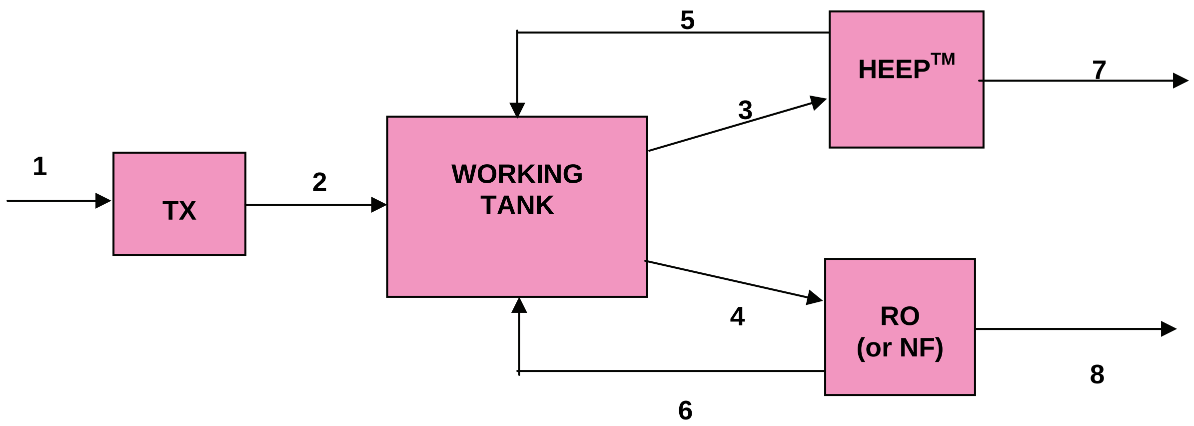
The unique aspect of the VSEP^{TM} technology is the capability for directly processing concentrate with minimal or no pretreatment to achieve high recoveries beyond that of an additional stage of RO. This is made possible by allowing sparingly soluble salts and silica to precipitate without significantly fouling the membrane. The simple nature and small footprint of the system are additional benefits.

6.4 EET'S HEEPM^{TM} TECHNOLOGY

6.4.1 Process Description

HEEPM^{TM} stands for high-efficiency electro-pressure membrane (EET, 2007). There are two keys to the HEEPM^{TM} technology. The first, and primary, key is a proprietary and patented ED stack design that significantly reduces the energy requirement and allows processing to high salinities. Salinities in excess of 200,000 mg/L have been achieved. EET Corporation developed the ED design and has been awarded multiple patents for both the HEEPM^{TM} system and the HEEP^{TM} (ED) system. The second key is a unique processing arrangement where ED and RO (or NF) technologies are both used, taking feed from the same working tank. This arrangement is depicted in the process flow diagram of Figure 6.6. ED product is returned to the tank, and the RO (or NF) concentrate is returned to the tank. Thus, the system waste is the ED waste, and the system product is the RO (or NF) product. This processing arrangement minimizes ED membrane area relative to ED-only systems while maximizing recovery relative to RO-only (or NF-only) systems. The HEEPM^{TM} system is applicable to batch, semibatch, or continuous flow arrangements. The advantages are most obvious for batch processing, where the arrangement allows for maintaining a lower feed concentration to the RO system while the batch volume is being reduced due to the treatment. Here, the purpose of the ED is to keep the RO (or NF) feed TDS at a relatively constant level over the high-recovery processing time. The arrangement allows for both the ED and RO (or NF) subsystems to work synergistically, each operating in their respective optimal ranges.

Figure 6.7 is a picture of a small HEEPM^{TM} skid-mounted system.



Stream #/Process Step	Description
1	Raw water
TX	Treatment step; includes typical pretreatment steps to remove sparingly soluble salts and silica
2	Treated water from pretreatment step
WORKING TANK	Holding tank capable of holding several minutes of feed flow; reaches a steady-state concentration
3	Feed to ED unit
HEEP™	ED unit based on EET proprietary design and patents
4	Feed to RO (or NF) unit
RO (or NF) Unit	Standard RO (or NF) system
5	Product stream from HEEP™ unit
6	Concentrate from RO (or NF) unit
7	Waste stream from HEEP™ unit; can reach 150,000 to more than 200,000 mg/L
8	Product stream from RO (or NF) unit

Figure 6.6. General process flow diagram of HEEPM™ technology.

While processing to high recovery could be done by the ED system alone, the combination of ED and RO is frequently more cost effective. Product quality is also improved with the dual system. Treatment to reduce the potential scaling of sparingly soluble salts and silica is the same as that used with other RO systems. Treatment is done on the front end of the process, prior to the working tank.

6.4.2 Processing Costs

EET Corporation was asked to consider how they would process four different water quality cases and to assign capital and operational costs to these processes. The four cases are presented in Tables 6.1 and 6.2 above. Case 13 is the same as the one studied for the ARROW^{TM} technology, except that the flow rate here is 1 MGD instead of 2 MGD. Case 1 is identical to that studied with the ARROW^{TM} and VSEP^{TM} technologies.

Table 6.7 presents the costs associated with these cases and includes the water quality and flow rates previously shown in Table 6.1, along with operational parameters and costs developed by EET Corporation for these cases. The capital cost was provided for equipment alone. Treatment costs are not included. Clean-in-place and PLC controls are standard features. Installation and contingency costs may vary from 10–40% of equipment costs.

The energy requirement can be seen to increase with feed salinity and is reflected in the operating costs per kilogallon.

As with the other systems’ manufacturers, the costs provided reflect the costing practices and approach taken by EET Corporation, and no effort was made on the study author’s part to standardize the costs and approach to costing taken by them.

Table 6.7. Parameters and Costs for HEEPM™

Constituent or Parameter	Case 1 Concentrate	Case 10 Raw Water	Case 13 Raw Water	Case 13 Concentrate
Ca, mg/L	731	75	65	426
Mg, mg/L	355	36	43	280
Na, mg/L	1226	2043	68	414
K, mg/L	63	63	2.2	13
Ba, mg/L	0	0	0.16	1.04
Sr, mg/L	0	0	1	6.6
CO_2, mg/L	0	0	2.1	2.6
HCO_3, mg/L	928	5682	348	2012
SO_4, mg/L	3564	6	74	654
Cl, mg/L	1111	95	56	368
F, mg/L	0	0	0.9	5.8
NO_3, mg/L	0	0	21	113
SiO_2, mg/L	22	0	25	157
pH	7	7	8	8
TDS, mg/L	8000	8000	705	
Other Parameters				
Nominal feed flow, MGD	0.064	1	1	0.124
Feed flow, gpm	44.2	709.2	709.2	86.1
Recovery, %	95.1	96.1	99.8	97.5
Feed TDS, mg/L	8000	8000	705	4500
Permeate (product) flow, gpm	42.1	682	708	2.2
Permeate (product) TDS, mg/L	705	275	275	705
Concentrate (waste) flow, gpm	2.11	26.8	1.5	
Concentrate (waste) TDS, mg/L	150,000	200,000	200,000	150,000
Energy requirement, kWh/kgal	36	41.1	10	18
Time in operation (h/day)	23.5	23.5	23.5	23.5
Capital cost, M$	0.52	4.59	0.98	0.69
Unit capital cost, M$/MGD	8.16*	4.59	0.98	5.56*
Capital cost (installed, existing bldg.)				
Capital cost (installed, new bldg.)				
Operating cost, $/kgal	3.95**	3.95	1.46	2.66**
Operating cost, M$/year	0.09	1.33	0.42	0.12
Cost of energy, $/kWh	0.08	0.08	0.08	0.08

*Higher cost due to small size; no benefit of economy of scale.

**Estimated from operating cost values in other cases.

6.4.3 Miscellaneous Information for HEEPM™

As with the other high-recovery technologies, most applications have been outside the United States and in industries other than the municipal sector. Some applications have been held off until patents are in place.

The RO system is typically low pressure, and the feed tank concentration is typically between 5000–20,000 mg/L of TDS; if it is below this level, batch-wise processing is used. The size of the ED system increases relative to the RO system as feed salinity increases and as RO recovery increases.

The systems are typically skid mounted or prepackaged, depending on size. The system can also be used with NF membranes.

Benefits of the HEEPM™ technology appear to be that high recoveries are possible due to the unique arrangement of the ED and RO processing steps. The two can be sized independently to achieve a given processing result, and the independent control of each offers a robust system capable of responding to changes in feed quality and quantity. For instance, changes in feed can be compensated for by adjusting the flow to the ED step. The system is capable of producing a highly concentrated brine stream of from 150,000–200,000 mg/L.

Figure 6.7. Small HEEPM™ skid-mounted system.

6.5 DISCUSSION AND SUMMARY

Table 6.8 is a composite table of the parameters and costs presented in Tables 6.3, 6.5, and 6.7. Direct cost comparisons between the technologies are of questionable validity for several reasons:

- Costing approaches and practices among the different groups providing costs could have differed significantly.
- Some capital costs are for equipment only, and some are for installed equipment. Some do not include all process steps.
- Capital costs vary as to the inclusion of contingency, engineering, administrative, and other costs.
- The operating costs were determined based on different membrane life assumptions, different cost-of-energy assumptions, and possibly other differing assumptions.

The table reflects some interesting trends.

- The energy requirements range from a low for the ARROW^{TM} system (5.5 kWh/kgal) to a high for the brine concentrator process (75 kWh/kgal).
- The general recovery levels include very high values for the ARROW^{TM} Case 13 (98%) and the HEEPM^{TM} Case 13 (99.8%). Although the single-step VSEP^{TM} recoveries are conservative estimates, treatment of such waters would likely first have a spiral-wound RO step followed by a VSEP^{TM} step. From this perspective, Table 6.7 reveals that similar very high recoveries could be attained.
- All costs appear less than those associated with brine concentrator-alone processing, consistent with the findings from the cost analysis of the different commercial ZLD processing schemes discussed in Chapter 5. The brine concentrator processes have enormous installation costs and high energy costs.

Table 6.8. Summary of Parameters and Costs

	Case 1				Case 10			Case 13			Case 13
Constituent or Parameter	Concentrate				Raw Water			Raw Water			Concentrate
Ca, mg/L	731				75			65			426
Mg, mg/L	355				36			43			280
Na, mg/L	1226				2043			68			414
K, mg/L	63				63			2.2			13
Ba, mg/L	0				0			0.16			1.04
Sr, mg/L	0				0			1			6.6
CO_2, mg/L	0				0			2.1			2.6
HCO_3, mg/L	928				5682			348			2012
SO_4, mg/L	3564				6			73.5			654
Cl, mg/L	1111				95			56			368
F, mg/L	0				0			0.9			5.8
NO_3, mg/L	0				0			20.5			113
SiO_2, mg/L	22				0			25			157
pH	7				7			8			8
TDS, mg/L	8000				8000			705			4452
Technology	**ARROW$^{\text{TM}}$**	**VSEP$^{\text{TM2}}$**	**BC**	**HEEPM$^{\text{TM1}}$**	**ARROW$^{\text{TM}}$**	**HEEPM$^{\text{TM1}}$**	**BC**	**ARROW$^{\text{TM}}$**	**VSEP$^{\text{TM2}}$**	**HEEPM$^{\text{TM1}}$**	**HEEPM$^{\text{TM1}}$**
Nominal feed flow, MGD	1 MGD	1 MGD	1 MGD	0.064	1 MGD	1 MGD	1 MGD	2 MGD	1 MGD	1 MGD	0.124
Feed gpm	694	700	694	44.2	694	709.2	694	1,388	700	709.2	86.1
recovery, %	90	80	97	95.1	91	96.1	95.2	98	85	99.8	97.5
feed TDS, mg/L	8000	8000	8000	8000	8000	8000	8000	705	705	705	4,500
permeate (product) gpm	626	560	673	42.1	633	682	661	1,254	595	707.7	2.2
permeate (product) TDS, mg/L	372	300*****	10	705	376	275	10	54	300*****	275	705
concentrate (waste) gpm	68	140	21	2.11	61.6	26.8	30.2	134	105	1.5	
concentrate (waste) TDS, mg/L			257,000	150,000		200,000	167,000			200,000	150,000
energy requirement, kWh/kgal	6.21***	13.4	75	36	6.14***	41.1	75	5.1***	11.6	10	18
hr/day	24	22	24	23.5	24	23.5	24	24	22	23.5	23.5
Capital cost	1.55	4.91	5	0.52	2.1	4.59	5	2.95	4.35	0.98	0.69
Capital cost (M$/MGD)	1.55	4.91	5	8.16*	2.1	4.59	5	1.48	4.35	0.98	5.56*
Capital cost (installed - existing bldg.)	2.65				3.5			5			
Capital cost (installed - new bldg.)	4.65		7.72**		6.3		7.72**	8.85			
Operating cost, $/kgal	1.34	5.34	6.86	3.95****	1.54	3.95	6.99	1.55	4.43	1.46	2.66****
Operating cost, M$/yr	0.44	1.44	2.43	0.09	0.51	1.33	2.43	1.11	1.26	0.42	0.12
cost of energy , $/kWh	0.07	0.04	0.08	0.08	0.07	0.08	0.08	0.07	0.04	0.08	0.08

COMMENTS

* Higher cost due to small size; no benefit of economy of scale

**BC system would be installed outdoors

***Estimated based on 4 kWh/kgal for low-pressure RO stage and 9 kWh/kgal for high-pressure RO stage

**** = estimated from operating cost values from other cases

***** = assumed; value is dependent on membrane used; no membrane specified in performance estimate

More generally, characterization of the three technologies reveals the following:

- Only a few years ago, when high-recovery systems were being discussed with respect to the municipal market, a fundamental question was whether high recovery systems could attain recoveries in the 90%+ range; the technologies presented here demonstrate that high recoveries are possible up to 95+% (dependent on the feed water quality).
- High-recovery systems, beyond the typical RO1–chemical treatment–RO2 linear processing scheme, are available, patented, and commercial.
- These systems have innovative designs and characteristics that allow higher recoveries beyond those of a linear system.
- VSEP^{TM} offers a system with minimal, if any, chemical requirements and is suitable for use with any membrane; it is typically used after a conventional brackish RO membrane system, operating directly on the concentrate.
- The ARROW^{TM} system has a recycle characteristic which offers the benefit of total system recoveries greater than the recoveries of the individual membrane steps if the individual membrane steps were in series.
- Direct comparison of the projected costs of the systems is problematical in that (1) the costs were developed on different bases, (2) the costs were calculated by different groups; and (3) the costs presented for the different technologies (as presented in Table 6.8) were sometimes on a different flow basis.
- In several of the cases, the starting water TDS was high to begin with, and higher than would typically be considered for potable water processing. This was because the waters represented concentrate from an initial RO processing step. In nearly all cases the technologies discussed would be applied to original, lower-TDS feedwaters, with the result that the costs would likely be less and the recoveries likely higher.
- The costs do not reflect the possible benefit of bypass in cases where feedwater TDS is low and the potable water TDS requirement is low.
- It appears that each technology has achieved success in other industries or is expected to achieve success based on pilot and demonstration studies.
- A common theme is that these companies have focused their marketing in other industries and largely outside of the United States. Industries other than municipal systems have more money to spend, can spend it quicker, and do not have to procure bids.
- Only the VSEP^{TM} system has been piloted in a potable water setting.
- Other benefits beyond cost come into play in determining the overall suitability of a technology. VSEP^{TM} has a small footprint and utilizes little or no chemicals. ARROW^{TM} has a unique recycle configuration that should provide benefits over the conventional linear RO1–chemical treatment–RO2 processing scheme. HEEPM^{TM} benefits increase when applied to treating batch systems, i.e., smaller applications.
- The characterization presented here is by no means complete, but it illustrates the potential of each of these technologies to play a role in high-recovery municipal systems.

CHAPTER 7

REGULATORY ISSUES

7.1 INTRODUCTION

Membrane desalination concentrate from conventional processing has rarely been found toxic, hazardous, or highly radioactive. When concentrate is further processed in high-recovery and ZLD situations to produce concentrated brine or a solid, the increased levels of constituents may render the waste toxic, hazardous, or otherwise of concern. It is this situation that is addressed in this chapter.

The disposal of concentrate, brine, and solids resulting from volume reduction processing is regulated based on the makeup of the concentrate, brine, and/or solids and the particular disposal method being considered.

Due to the higher salinity of the concentrate or brine from a volume reduction process, disposal options are typically limited to evaporation ponds, deep-well injection, and (if in proximity to the ocean) ocean discharge. Disposal is dependent on the level of toxicity, level of hazardous constituents, and levels of radionuclides.

Disposal of solids to a landfill is dependent on the levels of hazardous and radionuclide constituents.

This chapter discusses the regulatory framework for characterizing and categorizing brine and solid wastes. It does not address specific numerical standards, as they vary from state to state, with many states presently in the process of addressing these issues with regard to radionuclides.

7.2 TOXICITY

Toxicity is a concern primarily associated with the disposal of concentrate or brine to surface waters. Toxicity is determined through whole-effluent toxicity tests (bioassays).

Membrane concentrate usually has very low levels of process-added chemicals. The small amounts of chemical additives and the nontoxic choices available are such that if toxicity is found, it is most likely due to raw water constituents or salinity.

The fact that membrane concentrate can be toxic to specific indicator organisms is well established. A handful of historical occurrences of toxicity, as determined in whole-effluent toxicity tests, have been noted (Mickley et al., 1993). These include concentrates low in dissolved oxygen (typical of groundwater sources), concentrates high in dissolved H_2S or NH_3 from groundwater sources, concentrates with high heavy metal levels due to processing material incompatibilities, and concentrates (such as in southwest Florida) with high levels of naturally occurring radionuclide materials (NORMs). Since these historical occurrences, concentrates from groundwater sources intended for surface discharge have been routinely aerated to increase dissolved oxygen, degassed to remove H_2S and NH_3 (when present), and processed using materials carefully chosen to avoid metal contamination. In the case of

NORM processing in Florida, the concentrates have been disposed of by deep-well injection rather than by surface water discharge. In more recent times, some concentrates from groundwater sources have had major ion toxicity, a toxicity due to high or low levels of various common ions, such as calcium, fluoride, and potassium (Mickley, 2001b). This type of toxicity has occurred most often where mysid shrimp, a highly sensitive test organism, is required in the bioassay, such as in Florida. This type of toxicity is considered differently from other types, as it is due to common ions and is quickly diluted away. Toxicity is also possible when test organisms are exposed to salinity outside their normal range of adaptability.

Toxicity due to high levels of contaminants such as arsenic, perchlorate, methyl *tert*-butyl ether, and pesticides have not, to date, been a major problem in the United States, although situations involving high levels of such contaminants will likely increasingly occur due to the treatment of lower-quality, impaired source waters.

7.3 HAZARDOUS WASTES

The Resource Conservation and Recovery Act (RCRA) defines hazardous wastes. The definitions include both specific lists of hazardous wastes and clear and specific characteristics of such wastes. Membrane concentrate, brine, or solids from concentrate are not listed hazardous wastes. If, however, concentrate, brine, and solids from a concentrate contained sufficient amounts of listed hazardous wastes, concentrate, brine, or solids from the concentrate would be hazardous wastes. Membrane concentrate, brine, and solids from concentrate are generally not ignitable, toxic, reactive, or corrosive by RCRA definitions. Thus, the concern is associated with constituents in the original raw water or constituents added during processing that become concentrated enough to render the concentrate, brine, or solids from the concentrate hazardous. A helpful resource is the August 2006 U.S. EPA document entitled *A System’s Guide to the Identification and Disposal of Hazardous and Non-Hazardous Water Treatment Plant Residuals* (USEPA 816-F-06-001).

7.4 RADIONUCLIDES

7.4.1 Technologically Enhanced NORMs

The concern for radionuclides is mostly from waters containing NORMs and their subsequent treatment and concentration to produce technologically enhanced NORMs (TENORMs). Some locations may also have radionuclides from the processing or detonation of nuclear materials, but this would be the general exception. Treatment processes such as dewatering, IX, RO, and other volume reduction and ZLD processes may concentrate radionuclides to a level of concern.

TENORMs may or may not be hazardous. If a waste has radionuclides present *and* is also hazardous by RCRA standards, the waste is considered a “mixed waste.” In the absence of hazardous materials, a waste containing radionuclides is typically regulated by the same state agency that regulates effluents not containing radionuclides. These wastes may be considered separately from other wastes and regulated as such.

Whereas the guidelines and regulations for wastes other than those containing radionuclides are well defined and relatively straightforward to understand and document, the same is not true with radionuclide wastes. The Radionuclide Rule (pertaining to drinking water facilities) went into effect on Dec. 7, 2000, with a schedule through Dec. 31, 2007, for drinking water systems and states to meet requirements. It provided maximum contaminant levels for radionuclides in drinking water and mandated compliance for both drinking water and residuals produced in drinking water treatment facilities. It did not, however, provide guidelines on how to dispose of the residuals. A July 2005 U.S. EPA document titled *A Regulator's Guide to the Management of Radioactive Residuals from Drinking Water Treatment Technologies* provides general guidelines but no numerical standards. In the previous draft version of the document, numbers were provided. The U.S. EPA, however, subsequently decided to leave the numbers and details up to the states. States are at various levels of progress in developing policies and standards for disposal of these residuals. The general situation is not easily documented and represented. Many states are dealing with situations of radioactive residuals on a case-by-case basis. Other states have no residual issues at all due to low levels of NORMs.

Wastes containing uranium or thorium may be subject to the Nuclear Regulatory Commission's licensing requirements under the Atomic Energy Act. If the amount of uranium or thorium makes up less than 0.05 wt% of the residuals, it is exempt from Nuclear Regulatory Commission regulations because it is considered an "unimportant quantity." If Nuclear Regulatory Commission licensing is not required, the state regulations apply for uranium and thorium. The need for licensing due to a desalination plant residual is unlikely.

The regulation of radionuclides is an increasingly important concern for municipalities. The focus of this study is on desalination concentrate, brine, and solids residuals. However, a more significant concern exists with exhausted resins from IX processing, sludges (such as alum sludge) from coagulation processes, and used RO membrane elements from desalination processing. It is fundamentally important that radionuclide levels be determined in potential feedwaters at an early stage of planning for a new WTP. Many utilities may potentially accrue significant unforeseen disposal costs associated with disposal of resins, sludges, and RO membrane elements containing high levels of radionuclides.

7.4.2 State Regulation of Radionuclides

Radioactive waste is categorized according to its origin and not necessarily according to the level of radioactivity. The literature is not explicit or clear with regard to the numerical levels of radionuclides that dictate different categories of waste. Much of the detail of dealing with TENORMs is left to the states, many of which are in the process of defining their programs and policies. This includes defining what levels of radioactivity need be present for a waste to be radioactive and in need of special consideration. The level and complexity of state regulation of radionuclides vary widely. On one extreme, situations are handled on a case-by-case basis. On the other extreme, the regulations can be quite complex. Nevertheless, various general regulatory categories of liquid and solid wastes may be seen to apply to municipal membrane and ZLD wastes. The levels have more to do with types of wastes than levels or concentrations. The following list is the study author's interpretation of the different categories:

- Simple waste (nontoxic, nonhazardous, and does not contain radioactive material)
- Toxic waste (for waters considered for surface discharge; toxicity determined by whole-effluent toxicity tests), waste not containing hazardous or radionuclide material

- TENORM waste (containing levels of naturally occurring radionuclide material but no hazardous materials)
- Hazardous waste (contains some hazardous material)
- Mixed waste (contains radionuclides and hazardous material)
- Low-level radionuclide waste (contains radioactive source material; this category is separate from TENORMs)
- High-level radionuclide waste

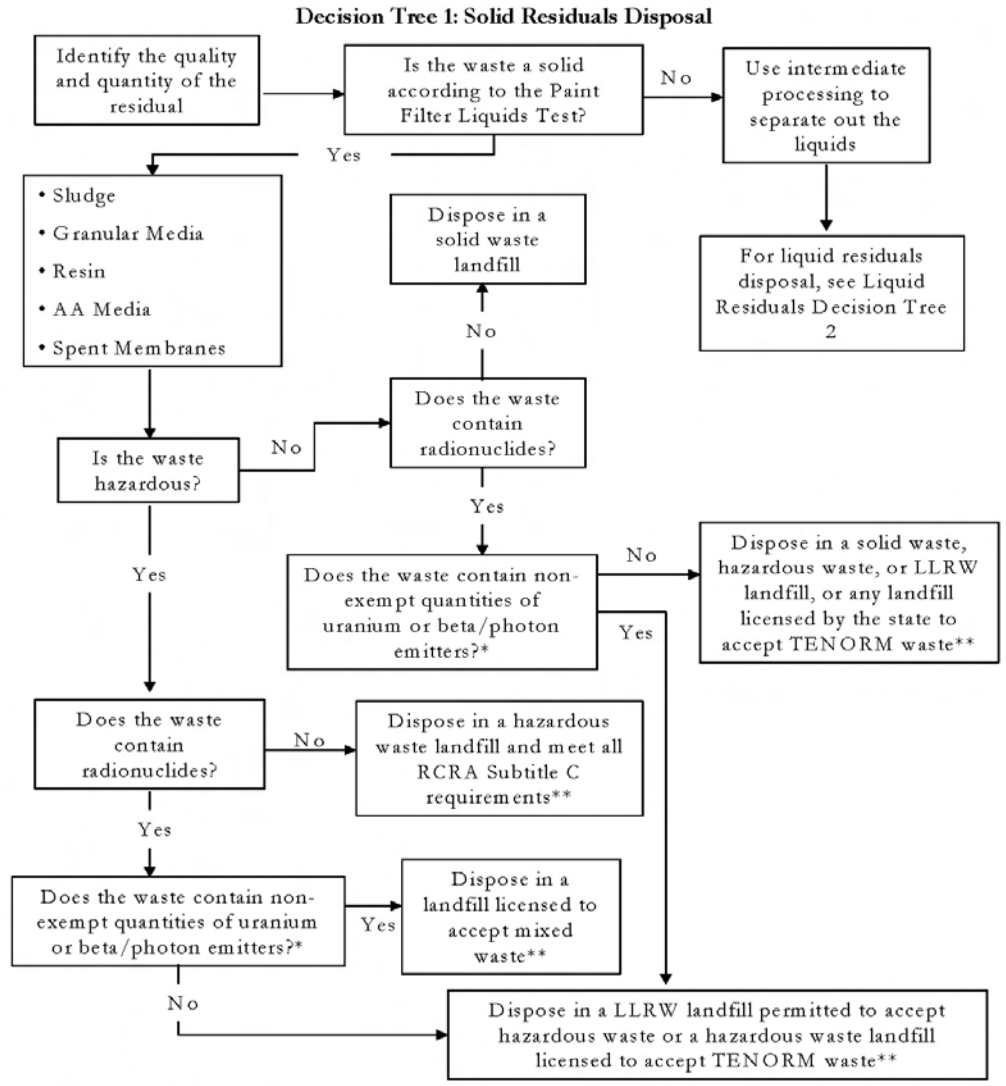
These categories do not reflect the potential difficulty of finding a disposal solution, as this depends on the levels of contaminants present. Disposal costs can increase significantly as the classification goes from standard to mixed waste. In most cases the presence of hazardous constituents or radionuclides will result in such high disposal costs that disposal will not be feasible for municipal settings unless volumes are small.

7.5 U.S. EPA DECISION TREES FOR LIQUID AND SOLID WASTE DISPOSAL

Figures 7.1, 7.2, and 7.3 are from the July 2005 U.S. EPA document entitled *A Regulators' Guide to the Management of Radioactive Residuals from Drinking Water Treatment Technologies* (USEPA 816-R-05-004). The decision trees are for liquid residuals disposal, solid residuals disposal, and liquid residuals disposal from intermediate processing. They are useful in that they include the full breadth of the various types of wastes.

In the solid residuals disposal decision tree (Figure 7.2), a box in the middle of the page asks, "Does that waste contain radionuclides?" The details as to what level of radionuclides renders a waste to be considered as containing radionuclides is left to the states. The box below this, however, asks if exempt quantities of certain radionuclides are present. The definition of exempt for uranium is (as previously mentioned) dictated by the federal Nuclear Regulatory Commission at 0.05 wt%. If the levels are not exempt, the waste must go to a licensed facility. If the levels are exempt, the waste may go to one of various types of landfills, depending on the level of the radionuclides present. The situation for beta and photon emitters is different, and the U.S. EPA directives suggest obtaining information on those radionuclides from the primary state agency.

The decision trees provided are comprehensive in scope and illustrate the dependence of the disposal option on whether or not the waste is hazardous and whether or not the waste contains radionuclides.



* Check with the state Radiation Program to see if beta/photon emitters are considered byproduct material and advise system to contact the NRC Regional office or relevant Agreement State agency to discuss potential licensing requirements.

** LDR treatment standards also apply. Check with the state Radiation Program to determine the proper disposal methods for waste containing radionuclides and hazardous waste.

Figure 7.1. Decision tree for solids residual disposal.

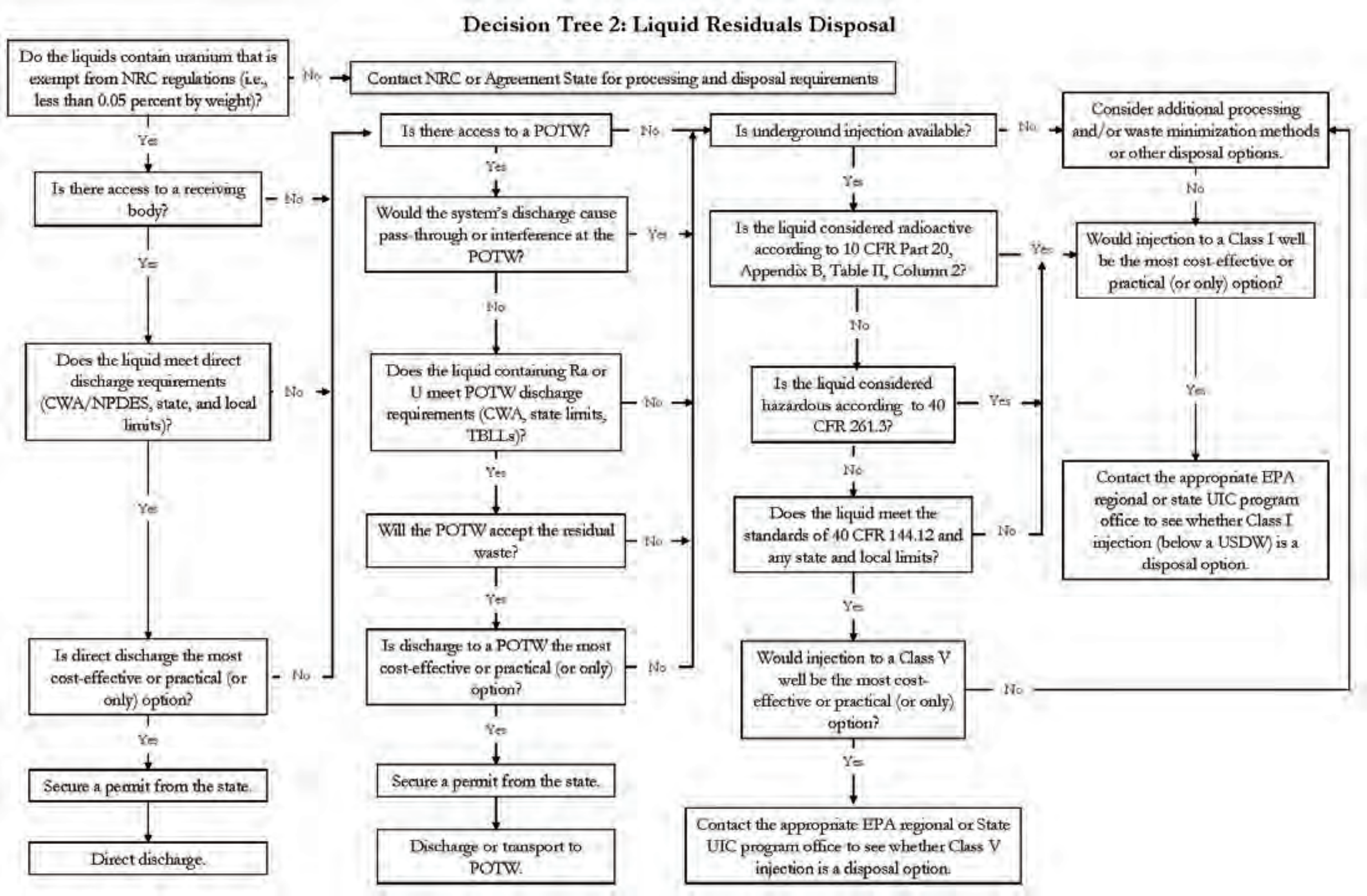


Figure 7.2. Decision tree for liquid residuals disposal.

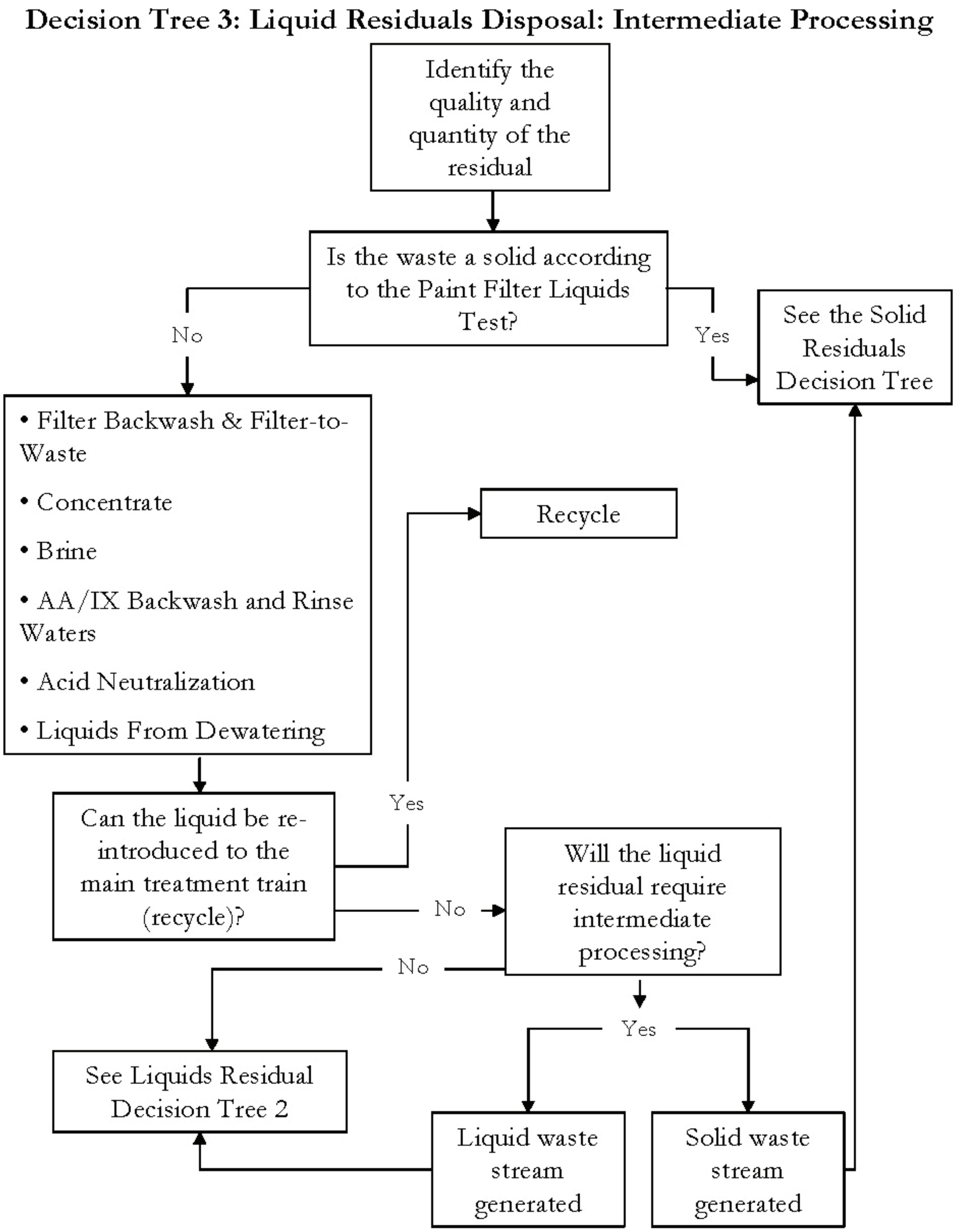


Figure 7.3. Decision tree for liquids residual from intermediate processing.

7.6 BRINE DISPOSAL

Disposal options for brine resulting from high-recovery processing are generally limited to those represented in Figure 3.1, above.

- Disposal of lower-salinity brine to a surface water
- Crystallization to solids with subsequent landfill of solids
- Evaporation ponds
- Deep-well injection
- Solidification of high-solids brine with subsequent landfill of solids
- Selective salt recovery

In cases where low-salinity liquid waste does occur (such as where the original water being treated is of low salinity), levels of contaminants in the concentrate need to be compared to listed hazardous wastes levels. Provided that the brine is not hazardous, the brine constituents then would be compared to receiving water standards to determine the feasibility of surface discharge.

Disposal of brine to evaporation ponds (not covered in the U.S. EPA decision trees) is governed by state regulations. Issues include pond construction and would reflect concerns for groundwater protection and concerns for wildlife. Higher-salinity brines are of more concern than lower-salinity brines. Individual state regulations need to be consulted for the particulars. Solids accumulating in evaporation ponds may need to be removed and disposed of in a landfill. In this case, the issues of solid disposal, discussed in the following section, apply.

Disposal via deep-well injection is governed by the Underground Injection Control Program. Issues peculiar to high-salinity brines include greater corrosion potential and greater concern for reservoir plugging. Where brines contain suspended solids, filtration prior to injection may be advisable. Solids produced from this filtration are subject to the issues of solids disposal, discussed in the following section. When solids are not present, the brine may contain salts at or near their solubility limits, and the potential for solids coming out of solution should be considered.

Treatments that produce solids transfer the regulatory issues to those of solids disposal. These are discussed in following sections.

7.7 SOLIDS ISSUES

Solids handling and disposal are not typical areas of consideration in conventional municipal membrane concentrate disposal. High-recovery systems may produce solids in pretreatment processes, such as lime softening. ZLD systems produce solids as a by-product of the processing. These solids may come from disposal of brine into evaporation ponds (with resultant solids buildup over time) or from final processing of brine into solids by crystallizers or spray dryers.

Issues concerning solids produced include:

- Solids disposal practices
- Solids disposal regulations
- Solids disposal costs

Solids not containing hazardous materials and not containing NORMs and TENORMs may be disposed of in an industrial waste landfill. Disposal costs include hauling (trucking) and tipping fees. The hauling cost depends on the distance between the solids generation site and the landfill. Hauling fees of $10–40/ton are not uncommon. Tipping fees are typically in the range of $15–30/ton, resulting in total solids disposal costs frequently in the range of $25–70/ton. Disposal of solids typically requires passing the paint filter liquids test, a method used to determine the presence of free liquids in a representative sample of waste. Wastes not passing the test may be “solidified” by addition of other salts or chemicals such that the waste will pass the test.

7.7.1 General Regulation of Solids Content (Not Containing Radionuclides)

The regulatory levels for nonradionuclide solid constituents are based on leachable levels. The Toxic Characteristic Leaching Procedure (TCLP) is a U.S. EPA SW-846 analytical method (Method 1311) that simulates sanitary landfill contaminant leaching from waste samples. Based upon concentrations of the TCLP constituents and guidelines set forth in 40 CFR 261.4, the solid waste samples can be deemed hazardous or nonhazardous. Different types of leaching tests have been developed in recent years to address a variety of leaching situations. The concentration of constituents leached from the solid are compared to established limits for various constituents. If the leached concentration is greater than the limit, the solid is considered a “toxicity characteristic” hazardous waste. The TCLP is the most commonly used test in the United States. It is intended to simulate worst-case conditions in a municipal solid waste landfill. The leaching test time is 18 h, the acid used is acetic acid, and the ratio of leachant (liquid) to solid (waste) is 20:1. California has developed the waste extraction test. The waste extraction test is a 48-h test that uses citric acid. The ratio of leachant to solids is 10:1. Both acids maintain a lower pH during the leaching procedure (compared to water leaching tests), and both the acetate and citrate ions (from acetic acid and citric acid, respectively) have the ability to complex or chelate with metals and thus achieve greater amounts of leaching than water leaching. In general, the TCLP test is more representative of a landfill, while the California waste extraction test leaches greater concentrations. Both the TCLP and the waste extraction tests are used as part of hazardous waste determinations.

Other tests have been developed to simulate waste leaching in environments outside of landfills (such as land application) or in landfills for which large amounts of acids are not expected to form. Such tests usually utilize a simulated rainfall or use deionized water.

There is no one leaching test that is better than others for all scenarios. The leaching test that is used really depends on the desired objective. If the objective is to determine whether a solid waste is a hazardous waste, a TCLP must be performed (or, in California, the waste extraction test).

The TCLP test involves a dilution ratio of 20:1 for a minimum 100-g sample. For a 100-g solids sample, the leaching fluid is 2000 g (2 L). The concentration of a leached constituent is then divided by 20 and compared to the TCLP limit for the metal (constituent) in question. Alternatively, the concentration can be compared directly to 20 times the TCLP limit. If the leached concentration is greater than 20 times the TCLP limit for the metal (constituent) in question, then the waste is considered to be a “toxicity characteristic” hazardous waste.

Table 7.1 lists the TCLP limit and 20 times the TCLP limit for several metals:

Table 7.1. TCLP Limits for Eight TCLP Metals and Benzene

Metal	TCLP Limit, mg/L	20× TCLP Limit, mg/L
Arsenic	5.0	100
Barium	100	2000
Cadmium	1.0	20
Chromium	5.0	100
Lead	5.0	100
Mercury	0.2	4
Selenium	1.0	20
Silver	5.0	100
Benzene	0.5	10

The TCLP test does not differentiate between the 3+ and 6+ chromium species but reports the combined concentration. The waste extraction test reports the 3+ and 6+ species separately. A more extensive list of TCLP limits is available through the Internet.

Prior to conducting TCLP tests, it is helpful to make a screening calculation to determine if the solid waste has enough of a particular metal (constituent) to fail the TCLP test for that metal (constituent). In this approach it is assumed that all of a given metal (constituent) in the solid will leach from the solid and be in the extraction fluid (leachant).

As an example, if 100 g of solid contained 200 ppm of lead (200 mg of lead/kg of total solid, and thus 20 mg of lead in the 100-g sample) and if all the lead were leached from the solid, the concentration of lead in the extraction fluid would be 20 mg in 2 L, or 10 mg/L, or 10 ppm. More simply, this can be calculated as 1/20 of 200 ppm, or 10 ppm (due to the 20:1 dilution). This is twice the TCLP limit for lead (5 mg/L) (see Table 7.1), and it is thus possible for the solid sample to fail the TCLP test. It might not fail the TCLP test if less than 5 ppm lead were extracted, and thus a TCLP test is warranted. (This same determination may be made by comparing the 200 ppm solid concentration with the 20× TCLP limit value, which is 100.) If, on the other hand, the 100 g of solid contained 40 ppm of lead, the maximum concentration in the extraction fluid would be 1/20 of 40, or 2 ppm. This is less than the TCLP limit for lead of 5 ppm. Thus, the solid would not fail the TCLP test for lead because there is not enough lead in the solid.

This screening calculation requires a solids analysis of metals to be reported in units of milligrams per kilogram of dry solids.

If the California leaching test (waste extraction test) were conducted instead of the TCLP test, the same approach could be taken to determine if the solid might fail the waste extraction test procedure. In this case, a factor of 10 would be used instead of 20, and comparison would then be made with the California waste extraction test standards.

The above screening calculation is based on having a metal (constituent) analysis of the solids. In lieu of this information, the solids analysis itself may be estimated from the concentration of a concentrate or brine.

7.7.2 Estimation of Solids Content from Concentrate (or Brine) Concentrations

Water quality analyses of concentrate (or brine) can be used to estimate solids content by assuming all solids present in the concentrate, and only solids present in the concentrate, end up as dry solids. While this may not be entirely accurate when treatment steps add or remove solids prior to the production of dry solids, usually it gives a worst-case estimate that is useful for an initial evaluation of the waste classification (unless solids with the same contaminants are added during treatment).

The concentration of individual constituents in dry solids may be estimated by eq 7.1:

$$[\text{mg/kg}] = 10^6 \times [\text{mg/L}] / \text{TS} \qquad \text{eq 7.1}$$

where: [mg/kg] is the constituent concentration in the dry solids, [mg/L] is the constituent concentration in the concentrate, and TS is the total solids in the concentrate, in milligrams per liter.

For instance, if a brine of 20,000 mg/L contained an arsenic concentration of 1 mg/L, the predicted dry solids content would be:

$$[\text{mg/kg}] = 10^6 \times 1/20{,}000 = 50 \text{ mg/kg}$$

If all of the 50 ppm of arsenic were able to be leached from the solid in a TCLP test, the extraction fluid (leachant) would contain 2.5 ppm of arsenic (1/20 of 50), which is less than the TCLP limit of 5. Alternatively, the same result is obtained by comparing the solids concentration of 50 ppm to the 20× TCLP limit of 100 ppm. In this case the solid is not a toxicity characteristic hazardous waste due to the arsenic content.

A similar analysis would be conducted for each metal (constituent).

7.7.3 Mixing Solids with Cement (for Disposal)

Mixing of the solids with cement can prevent contaminants from moving into the environment from treated wastes. This approach may be used to treat solids failing the TCLP tests to render them nonhazardous and likely to be disposed in industrial landfills. The cement used is the same as used in concrete. This solidification–stabilization process immobilizes contaminants within the cement–waste material. The immobilization occurs through physical and chemical bonding with the contaminants and in some cases via chemical changes within the contaminants themselves (PCA, 2007). The cement-based solidification–stabilization has been used to treat a variety of contaminants, including both inorganics and organics. The U.S. EPA considers solidification–stabilization an established treatment technology and has identified solidification–stabilization as a Best Demonstrated Available Technology for over 57 commonly produced industrial wastes (RCRA-listed hazardous wastes).

The amount of cement added typically varies between 5–30%. A granular final product is desired and can typically be obtained depending on the particular solids makeup and the mixing recipe and method used. The appropriate recipe is determined through a treatability study and depends on the individual waste. The residual water within the dewatered solids will be of high salinity, and this may accelerate the rate of setup of the cement. The high chloride content is not as much of a concern as it is in concrete work, because there is no need for steel reinforcement within the resulting waste.

The costs involved include:

- Cement
- Possible binders
- Mixing and processing
- Transportation
- Landfill (tipping)

The April 10, 2006, Engineering News Record reported a general cost of cement at just over $92/ton. If the cement content of the waste is 20 wt%, this represents a cost of about $14/ton final waste. Transportation and tipping costs for standard landfill are in the range of $20–$70/ton. When adding the cost of mixing and processing to the cost of cement and considering the disposal of an additional 20% in weight, it can be seen that the cost of disposal of solids may double when mixed with cement.

7.7.4 General Regulation of Solids Content (Containing Radionuclides)

The radionuclide content of solids may be estimated using the same general approach just discussed to estimate the solids content of nonradionuclide constituents. There are no federal guidelines for the standards, and particular standards thus can vary from state to state. The following example assumes a state screening level of 3 pCi/g for the radionuclide in question.

For radionuclides, the formula is shown in eq 7.2:

$$[\text{pCi/g}] = 1000 \times [\text{pCi/L}]/\text{TS} \qquad \text{eq 7.2}$$

where: [pCi/g] is the radionuclide concentration in dry solids and [pCi/L] is the radionuclide concentration in the concentrate; TS is the total solids in the concentrate, in milligrams per liter.

If the solids screening level is 3 pCi/g for a given radionuclide, then eq 7.2 can be used to construct a table relating the allowable radionuclide level in the concentrate to the TDS of the concentrate:

TDS	pCi/L
1000	3
5000	15
10,000	30
20,000	60

If a concentrate with TDS of 10,000 mg/L has a concentration of an individual radionuclide over 30 pCi/L, then a flag would be raised for this radionuclide. The suggestion is that if this concentrate were taken to dry solids, it would be singled out for further consideration.

7.8 SUMMARY

The primary concern is that further concentration of a membrane concentrate to a brine or solid may result in the brine or solid being hazardous or radioactive. The situation may be predicted to a large extent by an analysis of the starting concentrate (or even the starting raw water). In the case of raw water, the concentrate may be simulated using various software programs. It is important to obtain a broad characterization of the concentrate (or raw water) in terms of the constituents present to allow consideration of minor constituents. The level of a constituent present in the final brine or solid may be estimated by assuming that all of that constituent would still be present in the brine or solid. This calculated level of a constituent can then be compared with available standards to determine if, in this worst case, the resulting brine or solid would be considered hazardous or radioactive.

This task is complicated somewhat for the case of radionuclides, as standards are left to the individual states and many states are still in the process of defining policies and standards.

CHAPTER 8

REVIEW OF DESIGN, COST, AND OPERATING CONSIDERATIONS FOR BRINE CONCENTRATOR ZLD SYSTEMS

8.1 DESIGN CONSIDERATIONS

- One of the most essential factors in the design and operation of a successful ZLD facility must be the realization by all parties involved that a zero discharge facility is exactly that, zero discharge. If the actual operating conditions deviate somewhat from the assumed design criteria in a discharge facility, typically there is time and opportunity for corrective action to be taken while the plant continues to operate. However, if there is an equipment malfunction or the wastewater flow exceeds the design in a ZLD facility, shutdown of the facility can be imminent. In a ZLD facility the equipment responsible for disposal of the wastewater is just as important as the equipment producing drinking water. If either one fails, the facility has to be shut down.
- The wastewater disposal system must include an adequate design contingency. It must be recognized that there will always be some deviation from the theoretical design once the facility enters real-world operation. From time to time, operating variances will also occur which will result in flows that deviate from the design basis.
- When the first ZLD power-generating stations were developed in the 1970s, a survey of operating ZLD stations was conducted to develop a contingency basis that could be applied across the board to new ZLD generating stations. It was found that the liquid discharges were highest during the initial years, when operation of the plant was being optimized. The investigation resulted in the following contingencies:
    - 50% during the first year
    - 30% during the second year
    - 20% thereafter

  Although the survey was conducted on a limited number of ZLD operating stations several decades ago, these contingencies are still used on facilities designed today. The long-term 20% contingency is applied to ZLD equipment sizing. Typically, evaporation ponds are utilized at these stations for disposal of the final salts produced, and the first- and second-year contingencies are applied to the sizing of these ponds. An adequate design contingency is essential.
- Before detailed design begins, it is important to assemble the best design information available; in other words, learn from the experiences of others. Read articles and talk to those with direct operating experience with zero discharge facilities. There are numerous zero discharge electricity-generating stations, some with decades of operating experience, that can provide a wealth of practical design and operating experience. Most of the same design experiences at a ZLD generating station would apply equally to a ZLD municipal drinking water facility. Talking to an experienced operator can be invaluable. Learn from their successes and try to avoid their mistakes.

8.2 COST CONSIDERATIONS

A very important factor in any ZLD system is cost. In general, zero discharge systems are relatively costly compared to discharge-based systems, and significant attention is focused on optimizing costs. Selection of the right system at an optimum cost involves several considerations:

- **Economical versus Cheap:** Equipment suppliers and developers are keenly aware of the emphasis on controlling costs and are constantly seeking ways to be more competitive. Frequently, this effort results in cost savings. However, there is always the potential for a lower cost to reflect a lower-quality system or misleading economics. It is important to ensure that a lower-cost system truly reflects a more efficient system and not a reduction in quality or a bias in presenting the true cost of the system.
- **Design Contingency:** A design contingency is applied to most components in a treatment system to allow for variations in the operating conditions and allowances for wear and degradation of equipment over the projected life of the system. A design contingency is also implicit in sizing and selecting equipment to operate at the mid-range or nominal value of the potential range. One way to reduce costs is to select equipment and components that can meet the design conditions but nothing more. This approach results in a seemingly lower first cost but does little to support long-term integrity, reliability, and ease of operation.
- **Reliability:** Reliability is a factor which has a major impact on cost. Increased reliability is typically achieved by providing spare or redundant components. Increased reliability is also achieved by selecting higher-grade or more robust components. Increased reliability comes at a cost; deciding how much reliability is adequate can be a challenging exercise. One way to put these considerations into perspective is to look at the consequences of a system outage. What are the consequences or costs of a day of system outage? Can storage or another system supply the system demands while equipment is repaired? If so, high reliability and equipment redundancy have nominal value. At the other extreme are facilities such as power-generating stations and pharmaceutical facilities, where a day of outage can result in economic losses in the hundreds of thousands of dollars. When comparing the cost of two systems or alternatives, it is important to ensure they offer the same level of reliability and integrity.

  Long-term operation will be the best judge of the level of integrity and redundancy required. However, economics dictate that those decisions be made up front. For instance, the incremental cost to upgrade a pump to a higher-grade alloy can be nominal. However, the cost to replace that same pump later with one of a higher-grade alloy may be significant.

  ZLD systems with a proven operating history at other locations may warrant a nominal redundancy or contingency. On the other hand, developmental systems or those with a limited record of proven reliable operation will require a high level of contingency. For developmental systems, allowance should also be provided for an extended start-up period.

- **Simple versus Complex:** One of the most common approaches to achieving zero discharge is a brine concentrator. The heart of such systems is a large evaporator vessel fabricated of higher-cost alloys. This one piece of equipment can be costly, and reducing the volume of water it must process (and its corresponding size) can result in cost savings.

One approach to reducing the cost of these systems is to install additional equipment ahead of the brine concentrator to concentrate the water, reduce the volume, and thereby reduce the required capacity of the brine concentrator itself. Processes to reduce the volume going to the brine concentrator have included precipitation processes, ion exchange systems, and membrane systems. Frequently, by looking at only installed capital cost, some of these schemes appear to produce tangible savings which may not be realized when the overall operating costs of the system are taken into account.

More pieces of equipment require more operation and maintenance, which equates to more manpower. Also, as the complexity goes up, the reliability frequently goes down, since each piece of equipment must be operational for the overall system to function.

Pretreatment systems that require significant volumes of chemicals will produce corresponding volumes of waste that require disposal. Every truckload of chemicals delivered to the site will create a corresponding truckload of waste that must be hauled off-site or disposed of on-site.

- **Total Cost:** The total cost of each option must be evaluated completely when comparing options. The total cost includes not only the installed capital cost but also the annual operating costs, which include energy (power), labor, chemicals, maintenance, waste disposal, etc. There are a couple of accounting approaches to put all these costs on a common basis for comparison. One approach is the levelized cost basis, which translates the up-front capital cost to a series of equal annual payments. Adding the annual operating cost to the levelized capital cost provides an annual expense for the total cost of the system. This cost is referred to as an annualized cost. Another approach is the first-cost approach, which equates the annual operating costs to a lump sum current cost. Adding this cost to the up-front capital cost results in the actual total cost of owning and operating the system.

8.3 OPERATING CONSIDERATIONS

- While many people may be involved in the day-to-day operation of a zero discharge facility, it is recommended that one person have overall responsibility for the water management program. Automation can play a major role in coordinating equipment, but an overall lead person is still needed to coordinate operation and personnel. This person should have prime responsibility for the integrity of the overall water management program. At least once per month, this person should conduct an in-depth review of all critical operating parameters of the zero discharge operation to ensure that all goals and operating parameters are being met. This lead person should also coordinate the operations of the various site disciplines to resolve any differences in opinions and ensure an efficient team operating environment. The following factors are examples that should be included in the review:

    - Is the system operating according to design?
    - Is water usage correct and optimized for the system?
    - If chemical additives are used, are they the correct type and quantity?

- After each monthly critical review, any items of equipment contributing to abnormal or substandard operation should be identified. The equipment should then be inspected and corrective action recommended. Follow-up procedures should be established to ensure that corrective action has been taken and that it has produced the desired results.
- Operating data should be summarized the first of every month, normalized to account for the different lengths of months, and then compared with the corresponding data for the previous month to identify any evolving abnormalities in operation so that corrective action can be taken early. Long-term data analysis may also be useful for detecting more subtle operating trends.
- All plant data should be maintained in a complete, concise, and up-to-date condition. These data should be accessible to all personnel in a readily available format.
- Water analyses should be performed regularly, not only to validate operation of the equipment but also to ensure that none of the raw water parameters has varied significantly from the original design basis. Significant changes in the feedwater quality could impair the system’s ability to meet zero discharge.
- The zero discharge facility will have been designed on a theoretical plant water balance based on design data. Once plant operating data and on-site environmental measurements become available, the water balances should be refined to reflect actual operating conditions. The actual operating water balance should then be used as the reference basis for day-to-day operation.

CHAPTER 9

FINDINGS, CONCLUSIONS, AND RECOMMENDATIONS

Volume minimization and ZLD processes will play an increasingly important role in membrane concentrate management in the municipal setting. The situation is driven by the growing need to tap alternative water resources to meet potable water needs. While membrane processing is the technology of choice for treating such waters, the general feasibility of doing this depends on the ability to manage membrane concentrate in a cost-effective and environmentally sound manner. Traditional concentrate management options will continue to provide satisfactory solutions in many parts of the United States; however, there is a growing need for alternative solutions, particularly in the arid southwestern United States. Foremost in terms of alternative solutions are volume minimization and ZLD processing of concentrate. At present there are no municipal facilities utilizing high-recovery and ZLD technologies in the United States.

Against this background, the report has sought to achieve the following:

- Characterize the major commercially available volume minimization and ZLD technologies
- Evaluate the performance and economics of several of these technologies
- Identify and characterize promising commercial technologies
- Discuss associated regulatory issues

A summary of the points made in this report is provided in the next section, followed by sections that provide the resulting conclusions and recommendations.

9.1 FINDINGS

9.1.1 General findings

- **Distinction between high-recovery and ZLD processing systems:** It is recommended that ZLD be defined as a system in which no effluent leaves the plant boundary. Industrial wastewater ZLD processing systems are high-recovery systems that typically produce either brine sent to evaporation ponds within the plant boundary or solids transported to a suitable landfill. Not all high-recovery systems are ZLD systems, as disposal of brine may take place outside of the plant boundary. While use of the term ZLD is sometimes taken to mean evaporative high-recovery systems, with the increasing use of membrane processes in industrial wastewater ZLD systems this use of the term is now outdated.
- **Feasibility of high-recovery and ZLD processing:** The technical feasibility of high-recovery and ZLD processing is not an issue. Several commercial approaches have been developed and successfully applied in nonmunicipal industries. There are approximately 120 industrial wastewater ZLD facilities in the United States. Application of the ZLD processing schemes in these industries is obviously economically feasible. However, due to high costs, there are no high-recovery or ZLD operations at municipal sites. The question then has to do with the economic feasibility of high-recovery and ZLD systems for application in the municipal industry. For general application within the municipal industry, capital and operating cost reductions are necessary.

- **Characterization of commercial ZLD processing schemes:** There are five general commercial processing schemes used in the approximately 120 wastewater ZLD facilities (nonmunicipal) in the United States.

  Individual processing steps for treating wastewater in these schemes include:
  - Brackish RO
  - Lime softening
  - Thermal brine concentrator
  - Thermal crystallizer
  - Spray dryer (used only for small flows)
  - Evaporation pond
  - Landfill

  Product water is produced by the RO, brine concentrator, and crystallizer process steps. Processing steps for the wastewater treatment, beginning with concentrate (conc), are as follows:

Scheme 1A:	conc →	BC →		EP
Scheme 1B:	conc →	BC→	CRYST →	LF
Scheme 2A:	conc → LS → RO →	BC →		EP
Scheme 2B:	conc → LS → RO →	BC →	CRYST →	LF
Scheme 3:	conc → LS → RO →			EP

  Solids produced from lime softening also go to a landfill. Possible small-volume purge from the crystallizer would go to an evaporation pond.

- **Availability of disposal options for final wastes:** This report focused on the economic evaluation of commercial high-recovery and ZLD processes. This assumes that disposal options for final wastes will be available, which is not necessarily the case for a given site. As shown in the previous process characterization, the final disposal steps for the commercial ZLD processing schemes involve either evaporation ponds for brine disposal or landfills for solids disposal. In locations where evaporation ponds are not suitable, other processing schemes that produce solids must be used, even though they may be more expensive. Furthermore, for large desalination projects, the magnitude of evaporation pond area or landfill area required may be so large as to make the project infeasible.

- **High-recovery brine is generally a sodium-dominated brine:** Most salts of lower solubility than sodium salts are precipitated in processing steps, such as the softening and brine concentrator steps. In the latter case, lower-solubility salts are allowed to precipitate and become suspended solids in the brine. Variability in final brine TDS composition after high-recovery or ZLD processing is much less than the variability in TDS composition of the starting concentrate (feed to the high-recovery and ZLD processes), as a result of removal of lower-solubility salts. This final brine is most typically dominated by sodium salts.

9.1.2 ZLD Processing Performance

Concentrate salinity and composition significantly affect the performance of individual process steps and the five processing schemes. Examples include the following:

- **Brine Concentrator: concentrate going directly to the brine concentrator (Schemes 1A and 1B).** The degree to which the brine concentrator can concentrate feedwater is strongly dependent on the feedwater composition. This in turn determines the exiting brine total solids concentration. Over the range of feedwater (concentrate) compositions studied, and for the processing schemes in which concentrate went directly to the brine concentrator unit, the exiting brine concentration varied from a low of 167,000 mg/L to a high of 358,000 mg/L TDS. Factors limiting the degree of concentration included the following: (1) the formation of the double salt glauberite, $Na_2Ca(SO_4)_2$; (2) high levels of TSS produced; (3) the onset of NaCl precipitation.

- **Lime Softening–RO: concentrate going to softening and then second-stage RO units (Schemes 2A, 2B, and 3).** Regardless of the concentrate salinity and composition, conditions exiting the softening and second-stage RO treatment steps were similar in terms of residual hardness level and salinity. The salinity exiting the second-stage RO unit was assumed to be 60,000 mg/L in every case. Thus, variability in feed salinity to subsequent steps is eliminated and the variability in feed composition is reduced considerably by the softening–second-stage RO steps.

- **Brine Concentrator: concentrate undergoing softening and second-stage RO treatment prior to the brine concentrator (Schemes 2A and 2B).** In the processing schemes where concentrate was softened and then treated by a second-stage RO unit, variability in salinity and composition of feedwater to the brine concentrator was reduced significantly. The changes in composition significantly affected the degree to which the feedwater could be concentrated in the brine concentrator. In most cases, the exiting brine concentration was less than in the case without softening and second-stage RO processing. In two cases (Cases 8 and 10) the softening–second-stage RO processing resulted in NaCl-dominated water that could be concentrated from 247,000–263,000 mg/L TDS. In the other cases the exiting concentration from the brine concentrator was limited to 145,000 mg/L due to the formation of glauberite.

- **Brine Concentrator: the feed composition influence on energy requirements and materials of construction.** Unit energy requirements (kilowatt-hours per kilogallon) and materials of construction were also dependent on the feedwater composition. The situation with the highest energy requirement and the most costly material of construction corresponded to the feedwater most dominated by NaCl. This same feedwater was included in the case that achieved the highest exiting brine concentration (358,000 mg/L) from the brine concentrator.

- **Lime Softener: amount of chemicals required and solids produced (Schemes 2A, 2B, and 3).** Doubling the salinity for a given composition doubles the amounts of chemicals required and solids produced in the softening process. Changing the composition, such as by doubling the hardness level (while keeping the salinity constant), also doubles the amount of chemicals required and solids produced.

- **Crystallization: effects of prior processing on crystallization performance.** The primary effect of composition on crystallizer performance is the presence of highly soluble calcium and magnesium chlorides. Their presence may require the need for a purge stream from the crystallizer. Purge streams are assumed to go to evaporation ponds.

- **Evaporation Ponds: effects of salinity and composition of brine going to evaporation ponds (all schemes).** The salinity and composition of brine going to evaporation ponds varied considerably due to the changes taking place in the prior processing steps as well as the salinity and composition of the original concentrates. The specific conditions of salinity and composition are not easily predictable without a detailed analysis of the effects of each processing step. Salinity affects the net evaporation rate and in particular plays a key role in determining how quickly an evaporation pond will fill with solids. Higher salinity levels may result in ponds filling with solids within the lifetime of the desalination plant. In this case, ponds may need to be cleaned out and solids taken to a landfill or the ponds may need to be covered over and new ponds constructed. Consequently, the ultimate number of ponds and resultant costs are not easily predictable without a detailed analysis of how prior processing steps may affect the salinity and composition coming to the evaporation ponds.

- **Interaction between Processing Steps.** As an example, in the processing schemes that include lime softening and second-stage RO processing prior to the brine concentrator (Schemes 2A and 2B), while the second-stage RO processing considerably reduces the volume going to the brine concentrator, it also reduces the brine concentrator concentration limit due to changes in composition. This, in turn, has an effect on the flow reduction due to brine concentrator processing and increases the flow going to the evaporation ponds.

- **Specific Chemical Effects.** Silica levels did not affect the performance or cost of the processing schemes. Silica is removed with calcium in lime softening, given that there are adequate magnesium levels, which was the situation in all but one case. Silica is not a problem in brine concentrator processing, as it is adsorbed onto the circulating calcium sulfate slurry.

  With calcium (and magnesium) being substantially removed in lime softening, a minimal amount of calcium must be added back to ensure adequate calcium sulfate slurry formation in the brine concentrator. The level of calcium in the concentrate affects the chemical requirements and amount of solids produced at the lime softening step.

  Alkalinity levels and more specifically carbonate species levels in the concentrate affect the amount of acid addition prior to the brine concentrator. The carbonate species level also affects the amount and type of chemicals added and the amount of solids produced at the lime softening step.

  Sulfate levels may affect the amount of antiscalant added prior to the second RO unit. If sulfate levels are low, sulfate may need to be added prior to the brine concentrator to ensure adequate calcium sulfate for the seeded slurry operation.

9.1.3 ZLD System Costs

- **System size has a relatively small effect on unit system costs (dollars per MGD).** Large system sizes require multiple equipment modules for RO, brine concentrator, and crystallization components, thus minimizing the economy-of-scale factors more typically associated with increasing the size for individual modules. Evaporation ponds and landfills have a minimum economy of scale. Thus, overall there was little economy of scale influencing the total costs. Negotiated cost reductions on large orders of equipment may change this picture somewhat, but this possibility was not taken into consideration.

- **Concentrate salinity and composition had significant effects on unit capital costs (dollars per MGD), operating costs (dollars per kilogallon or dollars per year per MGD), and annualized costs (dollars per MGD) for the five processing schemes.** The significant effects of salinity and composition on individual process step performance translate into wide variations in equipment size and capital costs as well as operating costs. The result of adding up capital and operating costs for the five processing schemes and normalizing these costs based on process size (feed MGD) revealed the following:
    - The processing schemes with crystallizers (Schemes 1B and 2B) were in all cases more expensive than the processes without the crystallizer step.
    - The processing schemes with lime softening and second-stage RO (Schemes 2A, 2B, and 3) were in nearly all cases less expensive than the cases without a second-stage RO. Volume reduction prior to the application of thermal brine concentrators is nearly always cost-effective.
    - The processing scheme most consistently of lowest unit operating cost was Scheme 3, where there was no thermal evaporation equipment used. This was due to the fewest processing steps and the low operating cost of evaporation ponds.
    - The processing scheme most consistently of lowest unit capital cost was Scheme 2A, in which volume reduction by lime softening and second-stage RO preceded further volume reduction by the brine concentrator.
    - Scheme 3, without any thermal evaporative steps, had a relatively high unit capital cost due to the relatively larger evaporation pond acreage.
    - The processing scheme most consistently of lowest annualized cost was Scheme 2A, the system with lime softening, second-stage RO, and brine concentration.
    - Costs for processing Schemes 1A and 1B were relatively insensitive to salinity over the range considered.
    - Costs for processing Schemes 2A, 2B, and 3 increased significantly with salinity.
    - Costs for the original (historically) ZLD processing Schemes 1A and 1B may be cost-competitive for higher-salinity feedwaters.
    - Capital costs for evaporation ponds and landfills can be significant. They were frequently the largest individual capital cost process step and the largest source of capital cost variability among the processing schemes and cases studied.

9.1.4 Geo-Processors' Selective Salt Recovery Technology

- This technology for selective salt recovery has been patented, developed, and licensed outside of the United States (Geo-Processors, 2007).
- The processing approach, steps, performance, and costs were defined by analysis of concentrate treatment by this technology.
- The costs analysis demonstrates how salt recovery and marketing can beneficially impact total plant costs.
- In general, for the concentrates evaluated the most cost-effective salt to recover is precipitated calcium carbonate, a high-value salt used in specialty paper manufacturing.
- This potentially attractive economic situation perhaps obscures one important benefit: that environmental impacts are lessened by reducing the amount of solids disposed and

reducing the potential CO_2 footprint of the desalination plants through removal of dissolved CO_2 via product formation.

- Thus, the potential benefits are both economic and environmental, and selective salt recovery represents a movement toward the goal of sustainability.
- Selective salt recovery needs to be piloted and tested in the United States to determine its applicability and feasibility for treating concentrates and other U.S. waters.

9.1.5 Commercial High-Recovery Technologies

- There are several patented, commercial, high-recovery systems that may offer advantages over the conventional high-recovery approach of interstage treatment between two RO stages.
- These technologies may each have a marketing window of application in municipal settings.
- The suppliers of these technologies are all currently focused on industries other than municipal industries and generally on applications outside of the United States.
- A preliminary cost evaluation suggests that these technologies are all more cost-effective than direct use of brine concentrators, consistent with the results of the ZLD process scheme analysis.

9.1.6 Regulatory Issues

- Disposal options for high-salinity brine or mixed solids that result from high-recovery processing include evaporation ponds, deep-well injection, and solidification to solids for brines and landfill for solids.
- The primary issue is whether further concentration of concentrate will render the resulting brine or solid hazardous or radioactive.
- The hazardous material question can be addressed early in the planning process by (1) obtaining a broad characterization of concentrate (or raw water) constituents, including contaminants of high concern, (2) estimating the levels of these contaminants in brine or solid wastes generated from further concentration, and (3) comparing the estimated levels with hazardous waste standards.
- The radionuclide question can be addressed in the same manner but is complicated by the fact that the standards are to be developed by individual states and in many cases do not yet exist. In this situation, the state regulatory agencies handle such questions on a case-by-case basis.
- If high-salinity brine or mixed solids resulting from high-recovery processing are hazardous or contain radionuclides, disposal of these wastes may be considerably more expensive, and perhaps prohibitively expensive.
- Because the levels of contaminants are greater in the concentrate, a detailed water quality analysis of a concentrate may be better suited than a detailed analysis of raw water to determine the levels of concern that contaminants may play. It is possible, for example, that the level of a contaminant is nondetectable at feedwater concentrations but is problematic in the concentrate or subsequent brine or solids.

9.2 CONCLUSIONS

- **Feasibility of high-recovery and ZLD processing in the municipal desalination industry is not a technical issue but an economic one.** Commercial processing schemes readily achieve high recovery of wastewater in many nonmunicipal industries. These systems include evaporative-based, membrane-based and evaporative-based, and membrane-based processing schemes. The capital and operating costs of these processing schemes are presently cost-prohibitive for municipal use.

- **Salinity and composition of feedwater significantly affect the performance and costs of high-recovery and ZLD processing schemes.** Salinity and composition of feedwater affect the performance of individual processing steps in different ways. The general result is that total processing costs increase with increasing salinity and with increasing need to treat feedwater to allow high-recovery membrane processing. Thus, processing schemes employing membrane technology are more strongly affected by salinity and composition than processing schemes employing only evaporative technologies. Consequently, the costs of high-recovery and ZLD processing systems that utilize membrane technology can vary significantly with salinity and composition of feedwater.

  Despite the strong influence of feedwater salinity and composition on high-recovery and ZLD processing schemes employing membrane technology, these technologies are generally more cost-effective than evaporative-based processing schemes. As the feedwater salinity increases, evaporative-based processing schemes become more cost-competitive with schemes utilizing membrane processing.

- **Membrane use in high-recovery and ZLD systems results in lower capital cost and dramatic energy savings, but the energy savings may be offset in large part by increased chemical costs and increased solids disposal costs.**

- **Factors that most significantly contribute to high costs have been identified.** For the high-recovery and ZLD capital costs they are:
  - Cost of the main volume reduction processing step
  - Cost of the final disposal step (evaporation ponds for brine or landfill for solids)

  For the high-recovery and ZLD operating costs they are:
  - Energy
  - Chemicals

- **Potential avenues for needed cost reductions to support application of high-recovery and ZLD processing in the municipal industry have been identified.**

  For capital cost reduction:
  - Lower capital costs through improved or new volume reduction technologies

  For operating cost reduction:
  - Reduced energy consumption through use of membranes prior to or in place of evaporative processes
  - Reduced chemical use through use of technologies, such as VSEP^{TM}, which minimize chemical use
  - Reduced solids disposal through selective salt recovery

- **Detailed understanding of the effects of feedwater quality on high-recovery and ZLD processing steps is required for accurate performance and cost estimations.** The sensitivities of performance and cost to salinity and composition suggest that simplified analyses of high-recovery and ZLD processing costs are subject to large error. A detailed analysis based on an understanding of how each processing step is affected by salinity and composition is required for meaningful cost projections on these complicated, complex processing systems.

- **Concerns for producing brine or solids of a hazardous nature or containing radionuclides are sufficient to warrant extensive water quality analysis early in desalination projects where high-recovery processing is considered.**

9.3 RECOMMENDATIONS

- Cost reductions are necessary for application of most high-recovery and ZLD processing schemes. Research should be conducted with this goal in mind. In particular, the goal should be reductions in capital cost and reduction of operating costs due to energy and chemical requirements.

- Due to the significant effects of salinity and composition on system performance and cost, simplified analyses of high-recovery and ZLD system costs may be subject to considerable error. Future analyses need to consider the effects of water quality on each processing step to ensure good cost projections.

- It is important that the newer commercial technologies (Geo-Processors' SAL-PROC, EET Corporation's HEEPM, New Logic's VSEP, and O'Brien & Gere's ARROW) be piloted for municipal applications and benchmarked against the more traditional high-recovery and ZLD approaches.

- Detailed water quality analyses need to be done at the concentrate level to ensure that contaminants present at low levels in the feedwater (perhaps at undetectable levels) do not result in brine or solids being hazardous or containing problematic levels of radionuclides.

- There is a need to develop a knowledge base of high-recovery and ZLD processing and to acknowledge the real cost and environmental consequences of large-scale concentrate disposal for inland desalination plants.

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APPENDIX 1

ZLD PROCESS SCHEMATICS

A1.1 INTRODUCTION

The schematic for processing Scheme 2B was presented in Figure 4.1. Schematics for the other four general commercial ZLD processes are given here.

- Processing Scheme 1A: Fig. A1.1
- Processing Scheme 1B: Fig. A1.2
- Processing Scheme 2A: Fig. A1.3
- Processing Scheme 3: Fig. A1.4

SCHEME 1A

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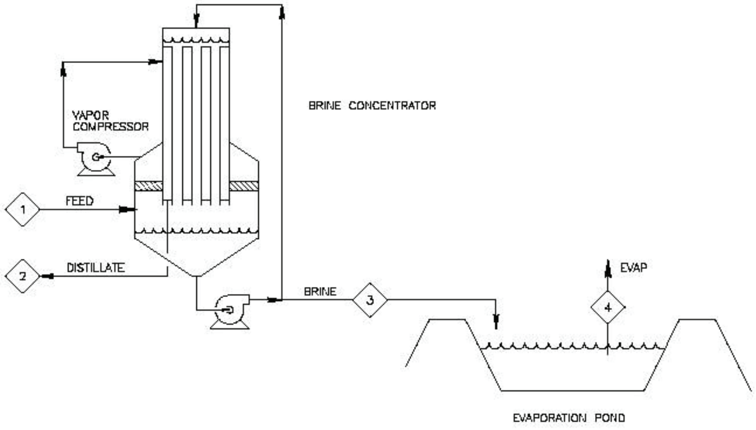

	CASE 1				CASE 2				CASE 3				CASE 4				CASE 5				CASE 6			
	1	2	3	4	1	2	3	4	1	2	3	4	1	2	3	4	1	2	3	4	1	2	3	4
Flow- MGD	10	9.7	0.3	0.3	1	0.97	0.03	0.03	20	19.4	0.6	0.6	1	0.96	0.04	0.04	20	19.1	0.9	0.9	10	9.7	0.3	0.3
Na	1226	1.5	39,385	0	613	1.5	39,416	0	613	1.5	39,416	0	1840	1.5	40,296	0	1840	1.5	40,296	0	1226	0	39,968	0
Ca	731	0.9	23,483	0	365	0.9	23,470	0	365	0.9	23,470	0	1096	0.9	24,002	0	1096	0.9	24,002	0	731	0	23,831	0
Mg	355	0.4	11,404	0	178	0.4	11,445	0	178	0.4	11,445	0	533	0.4	11,673	0	533	0.4	11,673	0	355	0	11,573	0
K	63	0.1	2,024	0	32	0.1	2,058	0	32	0.1	2,058	0	95	0.1	2,081	0	95	0.1	2,081	0	63	0	2,054	0
SO_4	3564	4.5	114,494	0	1782	4.5	114,583	0	1782	4.5	114,583	0	5346	4.5	117,077	0	5346	4.5	117,077	0	3564	0	116,186	0
Cl	1111	1.4	35,691	0	555	1.4	35,687	0	555	1.4	35,687	0	1666	1.4	36,485	0	1666	1.4	36,485	0	1111	0	36,219	0
HCO_3	926	1.2	29,748	0	464	1.2	29,835	0	464	1.2	29,835	0	1393	1.2	30,507	0	1393	1.2	30,507	0	926	0	30,188	0
Si (As SiO_2)	22	0.0	707	0	11	0.0	707	0	11	0.0	707	0	32	0.0	701	0	32	0.0	701	0	22	0	717	0
TDS-Approx	8,000	10	257,000	0	4,000	10	257,000	0	4,000	10	257,000	0	12,000	10	263,000	0	12,000	10	263,000	0	8,000	0	261,000	0
Pond Acres	81				8				162				12				232				81			

	CASE 7				CASE 8				CASE 9				CASE 10				CASE 11				CASE 12			
	1	2	3	4	1	2	3	4	1	2	3	4	1	2	3	4	1	2	3	4	1	2	3	4
Flow- MGD	10	9.7	0.3	0.3	10	9.8	0.2	0.2	10	9.7	0.3	0.3	10	9.6	0.4	0.4	10	9.7	0.3	0.30	10	9.6	0.4	0.40
Na	1226	1.5	34,696	0	1226	1.5	54,925	0	1226	1.5	34,696	0	1226	1.5	25,623	0	1226	1.5	39,968	0	1226	1.5	30,527	0
Ca	731	0.9	20,687	0	731	0.9	32,749	0	731	0.9	20,687	0	731	0.9	15,278	0	731	0.9	23,831	0	731	0.9	18,202	0
Mg	355	0.4	10,047	0	355	0.4	15,904	0	355	0.4	10,047	0	355	0.4	7,420	0	355	0.4	11,573	0	355	0.4	8,840	0
K	63	0.1	1,783	0	63	0.1	2,822	0	63	0.1	1,783	0	63	0.1	1,317	0	63	0.1	2,054	0	63	0.1	1,569	0
SO_4	3564	4.5	100,861	0	3564	4.5	159,667	0	3564	4.5	100,861	0	3564	4.5	74,488	0	3564	4.5	116,186	0	3564	4.5	88,744	0
Cl	1111	1.4	31,441	0	1111	1.4	49,773	0	1111	1.4	31,441	0	1111	1.4	23,220	0	1111	1.4	36,219	0	1111	1.4	27,664	0
HCO_3	926	1.2	26,206	0	926	1.2	41,485	0	926	1.2	26,206	0	926	1.2	19,353	0	926	1.2	30,188	0	926	1.2	23,057	0
Si (As SiO_2)	22	0.0	623	0	22	0.0	986	0	22	0.0	623	0	22	0.0	460	0	22	0.0	717	0	22	0.0	548	0
TDS	8,000	10	226,000	0	8,000	10	358,000	0	8,000	10	226,000	0	8,000	10	167,000	0	8,000	10	258,000	0	8,000	10	199,000	0
Pond Acres	89				59				89				116				81				107			

Figure A1.1. Process schematic for Scheme 1A.

SCHEME 1B

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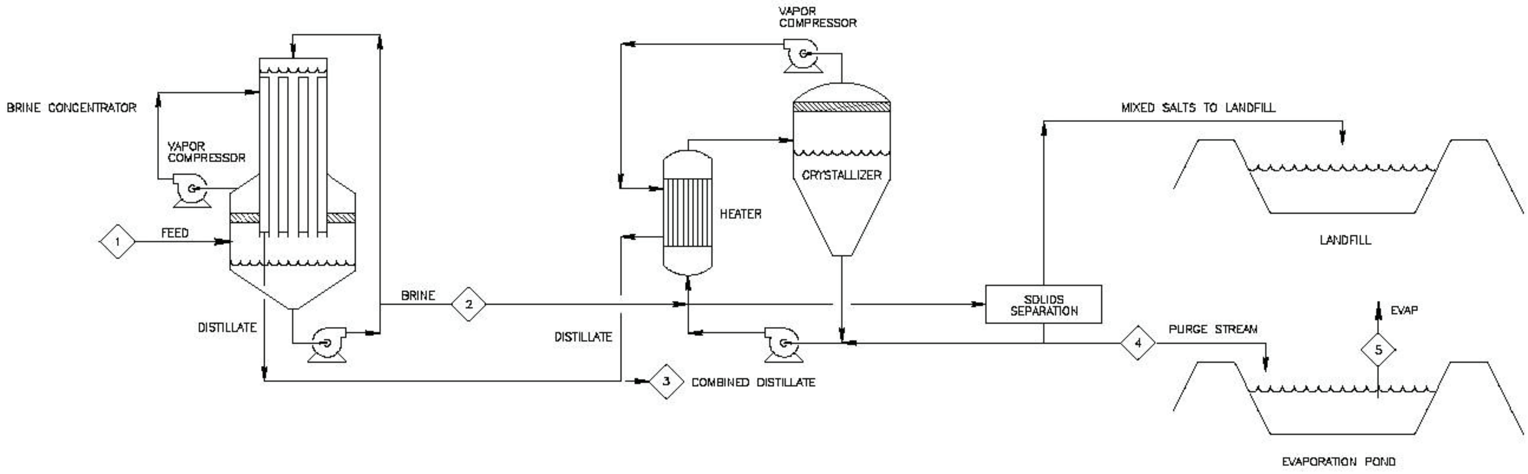

	CASE 1					CASE 2					CASE 3					CASE 4					CASE 5					CASE 6				
	1	2	3	4	5	1	2	3	4	5	1	2	3	4	5	1	2	3	4	5	1	2	3	4	5	1	2	3	4	5
Flow- MGD	10	0.302	9.984	0.016	0.016	1	0.030	0.999	0.001	0.001	20	0.606	19.984	0.016	0.016	1	0.043	0.996	0.004	0.004	20	0.870	19.911	0.089	0.089	10	0.302	10.000	0.000	0.000
Na	1226	39,385	1.5	68,963	0	613	39,416	1.5	68,963	0	613	39,385	1.5	68,963	0	1840	40,296	1.5	58,880	0	1840	40,296	1.5	58,880	0	1226	39,968	1.5	0	0
Ca	731	23,483	0.9	41,119	0	365	23,470	0.9	41,063	0	365	23,451	0.9	41,063	0	1096	24,002	0.9	35,072	0	1096	24,002	0.9	35,072	0	731	23,831	0.9	0	0
Mg	355	11,404	0.4	19,969	0	178	11,445	0.4	20,025	0	178	11,437	0.4	20,025	0	533	11,673	0.4	17,056	0	533	11,673	0.4	17,056	0	355	11,573	0.4	0	0
K	63	2,024	0.1	3,544	0	32	2,058	0.1	3,600	0	32	2,056	0.1	3,600	0	95	2,081	0.1	3,040	0	95	2,081	0.1	3,040	0	63	2,054	0.1	0	0
SO_4	3564	114,494	4.5	200,475	0	1782	114,583	4.5	200,475	0	1782	114,494	4.5	200,475	0	5346	117,077	4.5	171,072	0	5346	117,077	4.5	171,072	0	3564	116,186	4.5	0	0
Cl	1111	35,691	1.4	62,494	0	555	35,687	1.4	62,438	0	555	35,659	1.4	62,438	0	1666	36,485	1.4	53,312	0	1666	36,485	1.4	53,312	0	1111	36,219	1.4	0	0
HCO_3	926	29,748	1.2	52,088	0	464	29,835	1.2	52,200	0	464	29,812	1.2	52,200	0	1393	30,507	1.2	44,576	0	1393	30,507	1.2	44,576	0	926	30,188	1.2	0	0
Si (As SiO_2)	22	707	0.0	1,238	0	11	707	0.0	1,238	0	11	707	0.0	1,238	0	32	701	0.0	1,024	0	32	701	0.0	1,024	0	22	717	0.0	0	0
TDS-Approx	8,000	257,000	10	450,000	0	4,000	257,000	10	450,000	0	4,000	257,000	10	450,000	0	12,000	263,000	10	384,000	0	12,000	263,000	10	384,000	0	8,000	261,000	10	0	0
Pond Acres	10.7					0.6					10.7					2.7					56.1					0				

	CASE 7					CASE 8					CASE 9					CASE 10					CASE 11					CASE 12				
	1	2	3	4	5	1	2	3	4	5	1	2	3	4	5	1	2	3	4	5	1	2	3	4	5	1	2	3	4	5
Flow- MGD	10	0.333	9.990	0.010	0.010	10	0.222	10.000	0.000	0.000	10	0.333	9.958	0.042	0.042	10	0.435	10.000	0.000	0.000	10	0.302	9.980	0.020	0.020	10	0.400	10.000	0.000	0.000
Na	1226	34,696	1.5	62,986	0	1226	54,925	1.5	0	0	1226	34,696	1.5	57,316	0	1226	25,623	1.5	0	0	1226	39,600	1.5	62,833	0	1226	30,527	1.5	0	0
Ca	731	20,687	0.9	37,555	0	731	32,749	0.9	0	0	731	20,687	0.9	34,174	0	731	15,278	0.9	0	0	731	23,611	0.9	37,464	0	731	18,202	0.9	0	0
Mg	355	10,047	0.4	18,238	0	355	15,904	0.4	0	0	355	10,047	0.4	16,596	0	355	7,420	0.4	0	0	355	11,467	0.4	18,194	0	355	8,840	0.4	0	0
K	63	1,783	0.1	3,237	0	63	2,822	0.1	0	0	63	1,783	0.1	2,945	0	63	1,317	0.1	0	0	63	2,035	0.1	3,229	0	63	1,569	0.1	0	0
SO_4	3564	100,861	4.5	183,101	0	3564	159,667	4.5	0	0	3564	100,861	4.5	166,617	0	3564	74,488	4.5	0	0	3564	115,117	4.5	182,655	0	3564	88,744	4.5	0	0
Cl	1111	31,441	1.4	57,078	0	1111	49,773	1.4	0	0	1111	31,441	1.4	51,939	0	1111	23,220	1.4	0	0	1111	35,885	1.4	56,939	0	1111	27,664	1.4	0	0
HCO_3	926	26,206	1.2	47,573	0	926	41,485	1.2	0	0	926	26,206	1.2	43,291	0	926	19,353	1.2	0	0	926	29,910	1.2	47,458	0	926	23,057	1.2	0	0
Si (As SiO_2)	22	623	0.0	1,130	0	22	986	0.0	0	0	22	623	0.0	1,029	0	22	460	0.0	0	0	22	711	0.0	1,128	0	22	548	0.0	0	0
TDS	8,000	226,000	10	411,000	0	8,000	358,000	10	0	0	8,000	226,000	10	374,000	0	8,000	167,000	10	0	0	8,000	258,000	10	410,000	0	8,000	199,000	10	0	0
Pond Acres	6.5					78					26					0					13					0				

Figure A1.2. Process schematic for Scheme 1B.

SCHEME 2A

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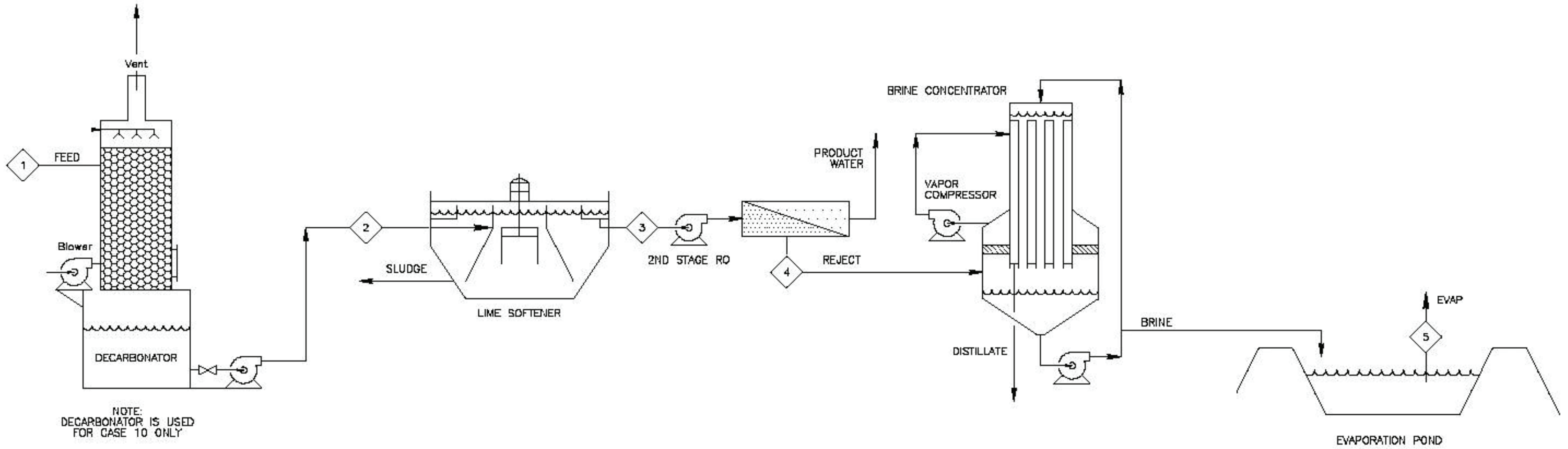


NOTE:
DECARBONATOR IS USED
FOR CASE 10 ONLY

	CASE 1					CASE 2					CASE 3				
	1	2	3	4	5	1	2	3	4	5	1	2	3	4	5
Flow- MGD	10	10.000	10	1.210	0.264	1	1	1	0.060	0.001	20	20	1	0.060	0.263
Na	1,226	1,226	2,450	19,695	0	613	613	1,244	19,364	0	613	613	1,244	19,364	0
Ca	731	731	0	0.0	0	365	365	0	0.0	0	365	365	0	0.0	0
Mg	355	355	0	0.0	0	178	178	0	0.0	0	178	178	0	0.0	0
K	63	63	63	524.0	0	32	32	32	507.0	0	32	32	32	507.0	0
SO_4	3,564	3,564	3,564	30,661	0	1,782	1,782	1,782	29,702	0	1,782	1,782	1,782	29,702	0
Cl	1,111	1,111	1,111	9,166	0	555	555	555	8,788	0	555	555	555	8,788	0
HCO_3	926	926	165	602.0	0	464	464	133	2302.0	0	464	464	133	2302.0	0
Si (As SiO_2)	22	22	0	0.0	0	11	11	0	0.0	0	11	11	0	0.0	0
TDS-Approx	8,000	8,000	7,350	60,803	282,000	4,000	4,000	3,743	61,010	282,000	4,000	4,000	3,743	61,010	282,000
Pond Acres					147					7					146

	CASE 4					CASE 5					CASE 6				
	1	2	3	4	5	1	2	3	4	5	1	2	3	4	5
Flow- MGD	1	1	1	0.190	0.004	20	20	1	0.190	0.819	10	10	10	1.450	0.317
Na	1,840	1,840	3,655	19,179	0	1,840	1,840	3,655	19,179	0	456	456	2,720	18,315	0
Ca	1096	1096	0	0.0	0	1096	1096	0	0.0	0	912	912	0	0.0	0
Mg	533	533	0	0.0	0	533	533	0	0.0	0	677	677	0	0.0	0
K	95	95	95	501.0	0	95	95	95	501.0	0	114	114	114	774.0	0
SO_4	5,346	5,346	5,346	28,943	0	5,346	5,346	5,346	28,943	0	4,577	4,577	4,577	24,634	0
Cl	1,666	1,666	1,666	8,793	0	1,666	1,666	1,666	8,793	0	798	798	798	5,415	0
HCO_3	1393	1393	189	989.0	0	1393	1393	189	989.0	0	336	336	163	0.4	0
Si (As SiO_2)	32	32	0	0.0	0	32	32	0	0.0	0	130	130	0	0.0	0
TDS-Approx	12,000	12,000	10,949	58,506	282,000	12,000	12,000	10,949	58,506	282,000	8,000	8,000	8,370	10	282,000
Pond Acres					23					457					96

	CASE 7					CASE 8					CASE 9				
	1	2	3	4	5	1	2	3	4	5	1	2	3	4	5
Flow- MGD	10	10	10	0.920	0.215	10	10	10	1.300	0.231	10	10	10	0.920	0.191
Na	1,074	1,074	1,831	19106	0	2,062	2,062	2959	22,488	0	1,009	1,009	1,907	19,908	0
Ca	967	967	0	0	0	574	574	16	122	0	941	941	0	0	0
Mg	212	212	0	0	0	147	147	2	0	0	310	310	0	0	0
K	73	73	73	771	0	97	97	0	737	0	24	24	24	253	0
SO_4	2,206	2,206	2,206	24,634	0	638	638	0	5,431	0	1,669	1,669	1,669	18,933	0
Cl	1,158	1,158	1,158	12,223	0	4,141	4,141	0	31,472	0	1,640	1,640	1,640	17,311	0
HCO_3	2176	2176	176	2,000	0	161	161	1075	882	0	2210	2210	172	1,953	0
Si (As SiO_2)	134	134	0	0.0	0	180	180	0	0.0	0	194	194	0	0	0
TDS	8,000	8,000	5,441	58,953	282,000	8,000	8,000	7,964	61,156	282,000	8,000	8,000	5,409	58,678	282,000
Pond Acres					116					142					104

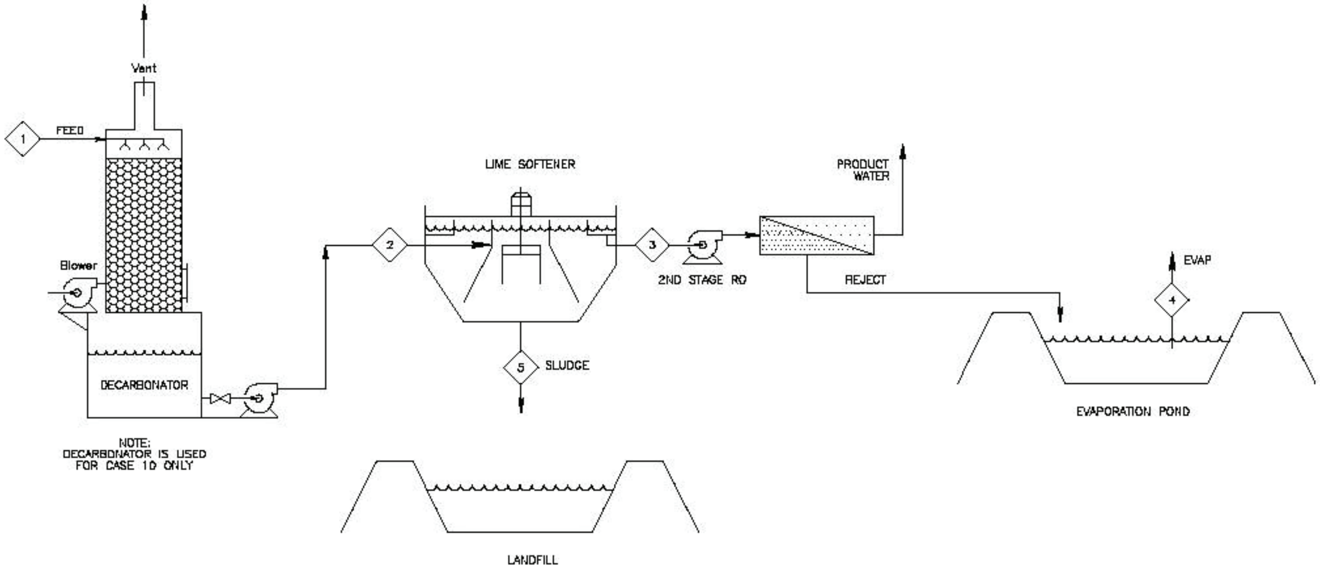
	CASE 10					CASE 11					CASE 12				
	1	2	3	4	5	1	2	3	4	5	1	2	3	4	5
Flow- MGD	10	10	10	1.260	0.315	10	10	10	1.240	0.272	10	10	10	1.350	0.273
Na	2,043	2,043	2,043	23,320	0	1,015	1,015	2,454	19,261	0	1,857	1,857	2,795	20,272	0
Ca	75	75	16	127.0	0	944	944	0	0.0	0	488	488	0	0.0	0
Mg	36	36	2	16.0	0	338	338	0	0.0	0	206	206	0	0.0	0
K	63	63	63	499.0	0	81	81	81	641.0	0	7	7	7	51.2	0
SO_4	6	6	6	1,550	0	3,313	3,313	3,313	27,152	0	3,852	3,852	3,852	28,956	0
Cl	95	3,013	3,013	34,224	0	1,323	1,323	1,323	10,474	0	1,376	1,376	1,376	10,055	0
HCO_3	5682	650	255	2,115.0	0	920	920	162	1347.0	0	184	184	154	1,171.0	0
Si (As SiO_2)	0	0	0	0.0	0	64	64		0.0	0	29	29		0.0	0
TDS	8,000	5,886	5,398	61,851	282,000	8,000	8,000	7,331	59,153	282,000	8,000	8,000	8,181	60,766	0
Pond Acres					162					151					144

Figure A1.3. Process schematic for Scheme 2A.

SCHEME 3

MICKLEY & ASSOCIATES
BOULDER, COLORADO

DATE:	SCALE:	SHEET:
07/07/06	NONE	1 of 1



	CASE 1					CASE 2					CASE 3				
	1	2	3	4	5	1	2	3	4	5	1	2	3	4	5
Flow- MGD	10	10	10	1.210	421 TPD	1	1	1	0.060	21 TPD	20	20	1	0.060	423 TPD
Na	1,226	1,226	2,429	19,695		613	613	1,223	19,364		613	613	1,223	19,364	
Ca	731	731	16	133.0		365	365	16	253.0		365	365	16	253.0	
Mg	355	355	2	17.0		178	178	2	32.0		178	178	2	32.0	
K	63	63	63	524.0		32	32	32	507.0		32	32	32	507.0	
SO_4	3,564	3,564	3,564	30,661		1,782	1,782	1,782	29,702		1,782	1,782	1,782	29,702	
Cl	1,111	1,111	1,111	9,166		555	555	555	8,788		555	555	555	8,788	
HCO_3	926	926	165	602.0		464	464	133	2302.0		464	464	133	2302.0	
Si (As SiO_2)	22	22	0	0.0		11	11	0	0.0		11	11	0	0.0	
TDS-Approx	8,000	8,000	7,350	60,803		4,000	4,000	3,743	61,010		4,000	4,000	3,743	61,010	
Acreage				284	68				14	3				281	69

	CASE 4					CASE 5					CASE 6				
	1	2	3	4	5	1	2	3	4	5	1	2	3	4	5
Flow- MGD	1	1	1	0.190	63 TPD	20	20	1	0.190	1,261 TPD	10	10	10	1.450	1,043 TPD
Na	1,840	1,840	3,634	19,179		1,840	1,840	3,634	19,179		456	456	2,699	18,315	
Ca	1096	1096	16	84.0		1096	1096	16	84.0		912	912	16	109.0	
Mg	533	533	2	11.0		533	533	2	11.0		677	677	2	14.0	
K	95	95	95	501.0		95	95	95	501.0		114	114	114	774.0	
SO_4	5,346	5,346	5,346	28,943		5,346	5,346	5,346	28,943		4,577	4,577	4,577	31,855	
Cl	1,666	1,666	1,666	8,793		1,666	1,666	1,666	8,793		798	798	798	5,415	
HCO_3	1393	1393	189	989.0		1393	1393	189	989.0		336	336	163	1,141.0	
Si (As SiO_2)	32	32	0	0.0		32	32	0	0.0		130	130	0	0.0	
TDS-Approx	12,000	12,000	10,949	58,506		12,000	12,000	10,949	58,506		8,000	8,000	8,370	57,750	
Acreage				45	10				877	205				340	96

	CASE 7					CASE 8					CASE 9				
	1	2	3	4	5	1	2	3	4	5	1	2	3	4	5
Flow- MGD	10	10	10	0.920	486 TPD	10	10	10	1.300	214 TPD	10	10	10	0.920	536 TPD
Na	1,074	1,074	1,810	19106		2,062	2,062	2959	22,488		1,009	1,009	1,908	19,908	
Ca	967	967	16	169		574	574	16	122		941	941	16	169	
Mg	212	212	2	21		147	147	2	15		310	310	2	21	
K	73	73	73	771		97	97	97	737		24	24	24	253	
SO_4	2,206	2,206	2,206	24,634		638	638	638	5,431		1,669	1,669	1,669	18,933	
Cl	1,158	1,158	1,158	12,223		4,141	4,141	4141	31,472		1,640	1,640	1,640	17,311	
HCO_3	2176	2176	176	2,000		161	161	111	882		2210	2210	172	1,953	
Si (As SiO_2)	134	134	0	0.0		180	180	0	0.0		194	194	0	0	
TDS	8,000	8,000	58,953			8,000	8,000	7,964	61,156		8,000	8,000	5,409	58,678	
Acreage				216	79				305	35				216	87

	CASE 10					CASE 11					CASE 12				
	1	2	3	4	5	1	2	3	4	5	1	2	3	4	5
Flow- MGD	10	10	10	2.000	472 TPD	10	10	10	1.240	455 TPD	10	10	10	1.350	233 TPD
Na	2,043	2,043	2,043	9,704		1,015	1,015	2,433	19,261		1,857	1,857	2,774	20,272	
Ca	75	75	16	76.0		944	944	16	127.0		438	488	16	117.0	
Mg	36	36	2	10.0		338	338	2	16.0		206	206	2	15.0	
K	63	63	63	299.0		81	81	81	641.0		7	7	7	51.2	
SO_4	6	6	6	19,724		3,313	3,313	3,313	27,152		3,852	3,852	3,852	28,956	
Cl	95	3,013	3,013	451		1,323	1,323	1,323	10,474		1,376	1,376	1,376	10,055	
HCO_3	5682	650	255	24,520.0		920	920	162	1347.0		134	184	154	1,171.0	
Si (As SiO_2)	0	0	0	0.0		64	64	0	0.0		29	29		0.0	
TDS	8,000	5,886	5,398	59,417		8,000	8,000	7,331	59,153		8,000	8,000	8,181	60,766	
Acreage				295	77				291	74				317	38

Figure A1.4. Process schematic for Scheme 3.

APPENDIX 2

MASTER COST TABLES

A2.1 INTRODUCTION

Table 5.1 presented the master cost table for processing Scheme 1A. Tables for the other four processing schemes are provided here. All other cost-related tables and figures were developed from these master cost tables.

The correspondence between the processing schemes and their respective tables is:

- Processing Scheme 1B: Table A2.1
- Processing Scheme 2A: Table A2.2
- Processing Scheme 2B: Table A2.3
- Processing Scheme 3: Table A2.4

Table A2.1. Performance, Design, and Cost Parameters for Process Scheme 1B

CONCENTRATE -> BRINE CONCENTRATOR-> CRYSTALLIZER->EVAPORATION POND-> LANDFILL														
Item	**Note**	**Cost Factor**	**Case No.**											
			1	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**
RO Concentrate	Concentrate flow, MGD		1	20	10	1	20	10	10	10	10	10	10	10
	Concentrate TDS, mg/L		4000	4000	8000	12 000	12,000	8000	8000	8000	8000	8000	8000	8000
Brine Concentrator	Feed flow, MGD		1	20	10	1	20	10	10	10	10	10	10	10
	Feed TDS, mg/L		4000	4000	8000	12,000	12,000	8000	8000	8000	8000	8000	8000	8000
	Brine TDS, mg/L	Depends on feedwater quality	257,000	257,000	257,000	263,000	263,000	281,000	226,000	356,000	226,000	167,000	258,000	199,000
	Brine flow, gpm	Depends on feedwater quality	11	216	216	32	633	213	246	155	246	332	215	279
	Concentration factor	Depends on feedwater quality	84.3	84.3	32.1	21.9	21.9	32.6	28.3	44.8	28.3	20.9	32.3	24.9
	Capital cost, installed, $M	Depends on feedwater quality	8.1	152.4	77.2	6.5	152.4	77.2	77.2	77.2	77.2	77.2	77.2	77.2
	Energy , kWh/1000 gal	Depends on feedwater quality	75.0	75.0	75.0	75.0	75.0	75.0	75.0	95.0	85.0	75.0	85.0	75.0
	Annual energy cost, M$/year	@ $0.08/kWh	2.2	43.8	21.9	2.2	43.8	21.9	21.9	27.7	24.8	21.9	24.8	21.9
	Annual O&M cost, $M	Assume 90% of cost is energy	2.4	48.7	24.3	2.4	48.7	24.3	24.3	30.8	27.6	24.3	27.6	24.3
Crystallizer	Feed flow, gpm		11.0	216.0	216.0	32.0	633.0	213.0	246.0	155.0	246.0	332.0	215.0	279.0
	Purge volume, gpm		0.6	11.0	11.0	3.0	62.0	0.0	7.0	80.0	29.0	0.0	14.0	0.0
	Purge salinity, mg/L		450,000	450,000	450,000	384,000	384,000	0	411,000	450,000	374,000	0	410,000	0
	Capital cost installed, $M	Depends on capacity	1.8	17.4	17.4	3.4	49.1	17.2	19.7	12.8	19.7	26.2	17.3	22.2
	Annual energy cost, M$/year	At 225 kWh/kgal, $0.08/kWh	0.1	2.0	2.0	0.3	6.0	2.0	2.3	1.5	2.3	3.1	2.0	2.6
	Annual O&M cost, $M	Assume 95% of cost is energy	0.1	2.2	2.2	0.3	6.3	2.1	2.4	1.5	2.4	3.3	2.1	2.8
Evaporation Pond (for purge stream)	Feed flow, gpm		0.6	11.0	11.0	3.0	62.0	0.0	7.0	80.0	29.0	0.0	14.0	0.0
	Evaporation Rate, gpm/acre	Depends on salinity	2.05	2.05	2.05	2.21	2.21	3.10	2.14	2.05	2.23	3.10	2.15	3.10
	Area, acres		0.3	5.4	5.4	1.4	28.1	0.0	3.3	39.0	13.0	0.0	6.5	0.0
	Purge salts for disposal, tons/day		2	30	30	7	143	0	17	216	65	0	34	0
	Final depth of solids, ft	74.9 lb/ft^3 for 20 years	24.8	24.8	24.8	22.8	22.8	0.0	23.6	24.8	22.4	0.0	23.7	0.0
	No. of ponds required	Assume 12-ft pond depth	3.0	3.0	3.0	2.0	2.0	0.0	2.0	3.0	2.0	1.0	2.0	0.0
	Total pond acreage required		0.9	16.1	16.1	2.7	56.1	0.0	6.5	117.1	26.0	0.0	13.0	0.0
	Capital cost, installed, $M	$0.30M per acre	0.3	4.8	4.8	0.8	16.8	0.0	2.0	35.1	7.8	0.0	3.9	0.0
	Annual O&M cost, $M	1% times capital	0.0	0.0	0.0	0.0	0.2	0.0	0.0	0.4	0.1	0.0	0.0	0.0
Dedicated Landfill	Total solids in feed, tons/day		17	333	333	51	999	334	334	333	334	333	333	333
	Solids to be landfilled, tons/day		15	304	304	44	856	334	316	117	269	333	299	333
	Wet solids, tons/day	85% for crystallizer	18	357	357	51	1,008	393	372	138	316	392	351	392
	Area, acres	10-ft pond depth, 143.6 lb/ft3, 20 years	4.6	91.2	91	13.1	257.5	100.3	95.1	35.2	80.8	100.0	89.7	100.2
	Capital cost, installed, $M	$0.25M per acre	1.2	22.8	22.8	3.3	64.4	25.1	23.8	8.8	20.2	25.0	22.4	25.0
	Annual O&M cost, $M	1% times capital+ $10/ton haul	0.1	1.5	1.5	0.2	4.3	1.7	1.6	0.6	1.4	1.7	1.5	1.7
Cost Summary														
Capital Cost, $M			11.3	197.5	122.3	14.0	282.7	119.5	122.6	133.9	124.9	128.4	120.9	124.5
Administrative, $M		25%	2.8	49.4	30.6	3.5	70.7	29.9	30.7	33.5	31.2	32.1	30.2	31.1
Contingency $M														
	Project contingency	25%	2.8	49.4	30.6	3.5	70.7	29.9	30.7	33.5	31.2	32.1	30.2	31.1
	Process contingency	5%	0.6	9.9	6.1	0.7	14.1	6.0	6.1	6.7	6.2	6.4	6.0	6.2
Total Capital Cost for Scheme, installed, $M			17.5	306.1	189.5	21.7	438.2	185.2	190.1	207.5	193.6	199.1	187.4	192.9
Total Annual O&M Cost for Scheme, $M/year			2.6	52.4	28.1	3.0	59.5	28.1	28.4	33.3	31.5	29.3	31.2	28.8
Total Annual O&M Cost for Scheme, $/kgal			7.2	7.2	7.7	8.2	8.1	7.7	7.8	9.1	8.6	8.0	8.6	7.9
Annualized Capital, O&M Cost, $M/year		20 years @ 8% interest	4.4	83.6	47.4	5.2	104.1	47.0	47.8	54.4	51.2	49.6	50.3	48.4
Recovery of Water from Concentrate, %			99.9	99.9	99.8	99.8	99.6	100.0	99.9	98.8	99.6	100.0	99.6	100.0
MWs Needed for Brine Concentrator			3.1	62.5	32.3	3.1	62.5	31.3	31.3	39.8	35.4	31.3	35.4	31.3

Table A2.2. Performance, Design, and Cost Parameters for Process Scheme 2A

CONCENTRATE ~> LIME SOFTENER ~> 2ND STAGE RO ~> BRINE CONCENTRATOR~> EVAPORATION POND														
Item	**Nots**	**Cost Factor**	**Case No.**											
			1	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**
RO Concentrate	Concentrate flow, MGD		1	20	10	1	20	10	10	10	10	10	10	10
	Concentrate TDS, mg/L		4000	4000	8000	12,000	12,000	8000	8000	8000	8000	8000	8000	8000
Decarbonator	Capital cost, installed, $M											0.15		
	Annual O&M cost, $M											0.02		
Lime Softener	Feed flow, MGD		1	20	10	1	20	10	10	10	10	10	10	10
	Lime added, tons/day	Hydrated lime @ $75/ton	3.8	76.6	74.4	11.1	221.3	99.0	91.0	25.4	104.4	24.9	72.0	34.6
	Soda ash added, tons/day	58% @ $135 /ton	5.9	117.3	115.8	17.3	345.2	215.9	70.9	86.3	84.4	0.0	136.4	88.3
	Solids produced (dry), tons/day		10.57	211.4	210.4	31.5	630.3	295	243	107	268.2	44.76	227.5	116.4
	Capital cost, installed, $M	@ $0.7M/MGD	0.7	14	7	0.7	14	7	7	7	7	7	7	7
	Annual chemical cost, $M		0.6	12.1	11.9	1.8	35.4	21.1	8.5	8.0	10.0	0.7	13.6	8.4
	Other annual O&M cost, $M	@ 2% of capital	0.014	0.28	0.14	0.014	0.28	0.14	0.14	0.14	0.14	0.14	0.14	0.14
	Total annual O&M cost, $M		0.62	12.34	12.01	1.78	35.67	21.19	8.65	8.17	10.17	0.82	13.70	8.59
2nd Stage RO	Feed flow, MGD	Mimimal water loss with lime sludge	1.0	20.0	10.0	1.0	20.0	10.0	10.0	10.0	10.0	10.0	10.0	10.0
	Feed TDS, mg/L		3743	3743	7350	10,949	10,949	8370	5441	7964	5409	7511	7331	8181
	% Recovery		94.0	94.0	88.0	82.0	82.0	86.0	91.0	87.5	91.0	88.0	88.0	87.0
	Brine flow, MGD		0.1	1.2	1.2	0.2	3.7	1.5	0.9	1.3	0.9	1.3	1.2	1.4
	Capital cost, installed, $M	@ $2.5M/MGD product	2.35	47	22	2.05	41	21.5	22.75	21.875	22.75	22	22	21.75
	Annual O&M cost, $M	@ $1.00/kgal	0.3	6.9	3.2	0.3	6.0	3.1	3.3	3.2	3.3	3.2	3.2	3.2
Brine Concentrator	Feed flow, MGD		0.06	1.2	1.21	0.19	3.74	1.45	0.92	1.3	0.92	1.26	1.24	1.35
	Feed TDS, mg/L		61,010	61,010	60,803	58,506	58,506	57,623	58,953	61,156	58,578	61,851	59,153	60,766
	Brine TDS, mg/L	Depends on feedwater quality	145,000	145,000	145,000	145,000	145,000	145,000	145,000	263,000	143,000	247,000	145,000	145,000
	Brine flow, gpm	Depends on feedwater quality	17.5	350	352	53.2	1047	400	282	161	284	219	283	291
	Concentration factor	Depends on feedwater quality	2.38	2.38	2.38	2.48	2.48	2.52	2.46	4.30	2.44	3.99	2.45	2.39
	Capital cost, installed, $M	Depends on feedwater quality	2	9	9	3	23.7	13	7.7	12	7.7	11.7	11.6	12.4
	Energy , kWh/1000 gal	Depends on feedwater quality	85	85	85	85	85	85	85	95	85	95	85	85
	Annual energy cost, M$/year	@ $0.08/kWh	0.15	2.98	3	0.47	9.28	3.6	2.28	3.61	2.28	3.5	3.08	3.35
	Annual O&M cost, $M	Assume 90% of cost is energy	0.17	3.31	3.33	0.52	10.31	4.00	2.53	4.01	2.53	3.89	3.42	3.72
Evaporation Pond	Feed flow, gpm		17.5	350	352	53.2	1047	400	282	161	284	219	283	291
	Evaporationrate, gpm/acre	Depends on salinity	2.8	2.8	2.8	2.8	2.8	2.8	2.8	2.3	2.8	2.7	2.8	2.8
	Area, acres		6.3	126.8	127.5	19.3	379.3	144.9	102.2	70.9	102.9	80.8	102.5	105.4
	Total mixed salts for disposal, tons/day		15	305	306	46	911	348	245	254	244	325	246	253
	Final depth of solids, ft	74.9 lb/ft3 for 20 years	10.7	10.7	10.7	10.7	10.7	10.7	10.7	16.0	10.6	18.0	10.7	10.7
	No. of ponds required	Assume 12-ft pond depth	1.0	1.0	1.0	1.0	1.0	1.0	1.0	2.0	1.0	2.0	1.0	1.0
	Total pond acreage required		6	127	128	19	379	145	102	142	103	162	103	105
	Capital cost, installed, $M	$0.30M/acre	1.9	38.0	38.3	5.8	113.8	43.5	30.7	42.6	30.9	48.5	30.8	31.6
	Annual O&M cost, $M	1% times capital	0.02	0.38	0.38	0.06	1.14	0.43	0.31	0.43	0.31	0.48	0.31	0.32
Dedicated Landfill	Wet solids, tons/day	50% solids for lime softener	21	423	421	63	1,261	590	486	214	536	90	455	233
	Area, acres	10-ft pond depth, 103 lb/ft3, 20 years	3	69	68	10	205	96	79	35	87	15	74	38
	Capital cost, installed, $M	$0.25M/acre	0.9	17.2	17.1	2.6	51.3	24.0	19.8	8.7	21.8	3.6	18.5	9.5
	Annual O&M cost, $M	1% times capital+ $10/ton haul	0.1	1.7	1.7	0.3	5.1	2.4	2.0	0.9	2.2	0.4	1.8	0.9
Cost Summary														
Capital Cost, $M			7.8	125.2	93.4	14.1	243.8	109.0	87.9	92.1	90.1	93.0	89.9	82.2
Administrative, $M		25%	2.0	31.3	23.3	3.5	60.9	27.2	22.0	23.0	22.5	23.2	22.5	20.6
Contingency, $M														
	Project contingency	25%	2.0	31.3	23.3	3.5	60.9	27.2	22.0	23.0	22.5	23.2	22.5	20.6
	Process contingency	5%	0.4	6.3	4.7	0.7	12.2	5.4	4.4	4.6	4.5	4.6	4.5	4.1
Total Capital Cost for Scheme, installed, $M			12.1	194.1	144.7	21.8	377.9	168.9	136.2	142.8	139.7	144.1	139.3	127.5
Total Annual O&M Cost for Scheme, $M/year			1.2	24.6	20.6	2.9	58.2	31.2	16.8	16.7	18.5	8.8	22.5	16.7
Total Annual O&M Cost for Scheme, $/kgal			3.4	3.4	5.7	8.0	8.0	8.5	4.6	4.6	5.1	2.4	6.2	4.6
Annualized Capital, O&M Cost, $M/year		20 years @ 8% interest	2.5	44.4	35.4	5.1	96.7	48.4	30.7	31.2	32.7	23.5	36.7	29.7
Recovery of Water from Concentrate, %			97.5	97.5	94.9	92.3	92.5	94.2	95.9	97.7	95.9	96.8	95.9	95.8
MWs needed for Brine Concentrator			0.2	4.3	4.3	0.7	13.2	5.1	3.3	5.1	3.3	5.0	4.4	4.8

Table A2.3. Performance, Design, and Cost Parameters for Process Scheme 2B

CONCENTRATE ~> LIME SOFTENER ~> 2ND STAGE RO ~> BRINE CONCENTRATOR~> CRYSTALLIZER~>LANDFILL														
Item	**Note**	**Cost Factor**	**Case No.**											
			1	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**
RO Concentrate	Concentrate flow, MGD		1	20	10	1	20	10	10	10	10	10	10	10
	Concentrate TDS, mg/L		4000	4000	8000	12,000	12,000	8000	8000	8000	8000	8000	8000	8000
Decarbonator	Capital cost, installed, $M											0.15		
	Annual O&M cost, $M											0.02		
Lime Softener	Feed flow, MGD		1	20	10	1	20	10	10	10	10	10	10	10
	Lime added, tons/day	Hydrated lime @ $75/ton	3.8	76.6	74.4	11.1	221.3	99.0	91.0	25.4	104.4	24.9	72.0	34.6
	Soda ash added, tons/day	58% @ $135 /ton	5.9	117.3	115.8	17.3	345.2	215.9	70.9	86.3	84.4	0.0	136.4	88.3
	Solids produced (dry), tons/day		10.57	211.4	210.4	31.5	630.3	295	243	107	268.2	44.76	227.5	116.4
	Capital cost, installed, $M	@ $0.70M/MGD	0.7	14	7	0.7	14	7	7	7	7	7	7	7
	Annual chemical cost, $M		0.6	12.1	11.9	1.8	35.4	21.1	8.5	8.0	10.0	0.7	13.6	8.4
	Other annual O&M cost, $M	@ 2% of capital	0.014	0.28	0.14	0.014	0.28	0.14	0.14	0.14	0.14	0.14	0.14	0.14
	Total annual O&M cost, $M		0.62	12.34	12.01	1.78	35.67	21.19	8.65	8.17	10.17	0.82	13.70	8.59
2nd Stage RO	Feed flow, MGD	Mimimal water loss with lime sludge	1.0	20.0	10.0	1.0	20.0	10.0	10.0	10.0	10.0	10.0	10.0	10.0
	Feed TDS, mg/L		3743	3743	7350	10,949	10,949	8370	5441	7964	5409	7511	7331	8181
	% Recovery		94.0	94.0	88.0	82.0	82.0	86.0	91.0	87.5	91.0	88.0	88.0	87.0
	Brine flow, MGD		0.1	1.2	1.2	0.2	3.7	1.5	0.9	1.3	0.9	1.3	1.2	1.4
	Capital cost, installed, $M	@ $2.5M/MGD product	2.35	47	22	2.05	41	21.5	22.75	21.875	22.75	22	22	21.75
	Annual O&M cost, $M	@ $1.00/kgal	0.3	6.9	3.2	0.3	6.0	3.1	3.3	3.2	3.3	3.2	3.2	3.2
Brine Concentrator	Feed flow, MGD		0.06	1.2	1.21	0.19	3.74	1.45	0.92	1.3	0.92	1.26	1.24	1.35
	Feed TDS, mg/L		61,010	61,010	60,803	58,506	58,506	57,623	58,953	61,156	58,578	61,851	59,153	60,766
	Brine TDS, mg/L	Depends on feedwater quality	145,000	145,000	145,000	145,000	145,000	145,000	145,000	263,000	143,000	247,000	145,000	145,000
	Brine flow, gpm	Depends on feedwater quality	17.5	350	352	53.2	1047	400	282	161	284	219	283	291
	Concentration factor	Depends on feedwater quality	2.4	2.4	2.4	2.5	2.5	2.5	2.5	4.3	2.4	4.0	2.5	2.4
	Capital cost, installed, $M	Depends on feedwater quality	2	9	9	3	23.7	13	7.7	12	7.7	11.7	11.6	12.4
	Energy, kWh/1000 gal	Depends on feedwater quality	85	85	85	85	85	85	85	95	85	95	85	85
	Annual energy cost, M$/year	@$0.08/kWh	0.15	2.98	3	0.47	9.28	3.6	2.28	3.61	2.28	3.5	3.08	3.35
	Annual O&M cost, $M	Assume 90% of cost is energy	0.17	3.31	3.33	0.52	10.31	4.00	2.53	4.01	2.53	3.89	3.42	3.72
Crystallizer	Feed flow, gpm		17.5	350.0	352.0	53.2	1,047.0	400.0	282.0	161.0	284.0	219.0	283.0	291.0
	Purge volume, gpm		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
	Solids produced (dry), tons/day		15.2	304.6	306.4	46.3	911.3	348.2	245.4	254.2	243.8	324.7	246.3	253.3
	Capital cost, installed, $M	Depends on feed water quality	2.3	27.6	27.8	4.0	80.6	31.4	22.4	13.5	22.6	16.0	22.5	23.1
	Energy consumption, kWh/kgal	Depends on feed water quality	225	225	225	225	225	225	225	250	225	250	225	225
	Annual energy cost, M$/year	at 250 kWh/kgal; $0.08/kWh	0.18	3.68	3.70	0.56	11.01	4.20	2.96	1.69	2.99	2.30	2.97	3.06
	Annual O&M cost, $M	Assume 95% of cost is energy	0.19	3.87	3.89	0.59	11.59	4.43	3.12	1.78	3.14	2.42	3.13	3.22
Dedicated Landfill	Wet solids, tons/day	50% for LS; 85% for crystallizer	39	781	781	117	2333	1,000	775	513	823	472	745	531
	Area, acres	10-ft pond depth, 143.6 lb/ft^3, 20 years	11	229	229	34	684	296	232	146	248	127	222	152
	Total mixed salts for disposal, tons/day		34	680	680	102	2030	870	674	810	707	699	648	462
	Capital cost, installed, $M	$0.25M/acre	2.9	57.2	57.2	8.6	170.9	74.1	57.9	36.5	61.9	31.6	55.5	37.9
	Annual O&M cost, $M	1% times capital+ $10/ton haul	0.2	3.4	3.4	0.5	10.2	4.4	3.4	2.2	3.6	2.0	3.3	2.3
Cost Summary														
Capital Cost, $M			10.2	154.8	123.0	18.3	330.2	147.0	117.8	90.8	122.0	88.5	118.6	102.2
Administrative, $M		25%	2.6	38.7	30.7	4.6	82.6	36.8	29.4	22.7	30.5	22.1	29.6	25.5
Contingency, $M														
	Project contingency	25%	2.6	38.7	30.7	4.6	82.6	36.8	29.4	22.7	30.5	22.1	29.6	25.5
	Process contingency	5%	0.5	7.7	6.1	0.9	16.5	7.4	5.9	4.5	6.1	4.4	5.9	5.1
Total Capital Cost for Scheme, installed, $M			15.8	240.0	190.6	28.4	511.8	227.9	182.6	140.8	189.1	137.2	183.8	158.4
Total Annual O&M Cost for Scheme, $M/year			1.5	29.8	25.9	3.7	73.8	37.1	21.0	19.4	22.8	12.4	26.7	21.0
Total Annual O&M Cost for Scheme, $/kgal			4.1	4.1	7.1	10.2	10.1	10.2	5.8	5.3	6.2	3.4	7.3	5.8
Annualized Capital, O&M Cost, $M/year		20 years @ 8% interest	3.1	54.3	45.3	6.6	125.9	60.4	39.6	33.7	42.0	26.4	45.5	37.2
Recovery of Water from Concentrate, %			100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0
MWs Needed for Brine Concentrator			0.2	4.3	4.3	0.7	13.2	5.1	3.3	5.1	3.3	5.0	4.4	4.8

Table A2.4. Performance, Design, and Cost Parameters for Process Scheme 3

CONCENTRATE ~> LIME SOFTENER ~> 2ND STAGE RO ~> EVAPORATION POND & LANDFILL														
Item	**Note**	**Cost Factor**	**Case No.**											
			1	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**
RO Concentrate	Concentrate flow, MGD		1	20	10	1	20	10	10	10	10	10	10	10
	Concentrate TDS, mg/L		4000	4000	8000	12,000	12,000	8000	8000	8000	8000	8000	8000	8000
Decarbonator	Capital cost, installed, $M											0.15		
	Annual O&M cost, $M											0.02		
Lime Softener	Feed flow, MGD		1	20	10	1	20	10	10	10	10	10	10	10
	Lime added, tons/day	Hydrated lime @ $75/ton	3.8	76.6	74.4	11.1	221.3	99.0	91.0	25.4	104.4	24.9	72.0	34.6
	Soda ash added, tons/day	58% @ $135 /ton	5.9	117.3	115.8	17.3	345.2	215.9	70.9	86.3	84.4	0.0	136.4	88.3
	Solids produced (dry), tons/day		10.57	211.4	210.4	31.5	630.3	295	243	107	268.2	44.76	227.5	116.4
	Capital cost, installed, $M	@ $0.7M/MGD	0.7	14	7	0.7	14	7	7	7	7	7	7	7
	Annual chemical cost, $M		0.6	12.1	11.9	1.8	35.4	21.1	8.5	8.0	10.0	0.7	13.6	8.4
	Other annual O&M cost, $M	@2% of capital	0.014	0.28	0.14	0.014	0.28	0.14	0.14	0.14	0.14	0.14	0.14	0.14
	Total annual O&M cost, $M		0.62	12.34	12.01	1.78	35.67	21.19	8.65	8.17	10.17	0.82	13.70	8.59
2nd Stage RO	Feed flow, MGD	Mimimal water loss with lime sludge	1.0	20.0	10.0	1.0	20.0	10.0	10.0	10.0	10.0	10.0	10.0	10.0
	Feed TDS, mg/L		3743	3743	7350	10,949	10,949	8370	5441	7964	5409	7511	7331	8181
	% Recovery		94.0	94.0	88.0	82.0	82.0	86.0	91.0	87.5	91.0	88.0	88.0	87.0
	Brine flow, MGD		0.1	1.2	1.2	0.2	3.7	1.5	0.9	1.3	0.9	1.3	1.2	1.4
	Brine TDS, mg/L		61,010	61,010	60,803	58,506	58,506	57,623	58,953	61,156	58,578	61,851	59,153	60,766
	Capital cost, installed, $M	@$2.5M/MGD product	2.35	47	22	2.05	41	21.5	22.75	21.875	22.75	22	22	21.75
	Annual O&M cost, $M	@ $1.00/kgal	0.3	6.9	3.2	0.3	6.0	3.1	3.3	3.2	3.3	3.2	3.2	3.2
Evaporation Pond	Feed flow, gpm		42	833	840	132	2596	1006	638	902	638	874	861	937
	Evaporation rate, gpm/acre	Depends on salinity	2.96	2.96	2.96	2.96	2.96	2.96	2.96	2.96	2.96	2.96	2.96	2.96
	Area, acres		14	281	284	45	877	340	216	305	216	295	291	317
	Total mixed salts for disposal, tons/day		15	305	306	46	912	348	226	331	225	325	306	342
	Final depth of solids, ft	74.9 lb/ft3 for 20 years	4.9	4.9	4.8	4.7	4.7	4.6	4.7	4.9	4.7	4.9	4.7	4.8
	No. of ponds required	Assume 12-ft pond depth	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0
	Total pond acreage required		14	281	284	45	877	340	216	305	216	295	291	317
	Capital cost, installed, $M	$0.30M per acre	4.2	84.4	85.1	13.4	263.1	102.0	64.7	91.4	64.7	88.6	87.2	95.0
	Annual O&M cost, $M	1% times capital	0.04	0.84	0.85	0.13	2.63	1.02	0.65	0.91	0.65	0.89	0.87	0.95
Dedicated Landfill	Wet solids, tons/day	50% solids for LS	21	423	421	63	1,261	590	486	214	536	90	455	233
	Area, acres	10-ft pond depth, 103 lb/ft3, 20 years	3	69	68	10	205	96	79	35	87	15	74	38
	Capital cost, installed $M	$0.25M/acre	0.9	17.2	17.1	2.6	51.3	24.0	19.8	8.7	21.8	3.6	18.5	9.5
	Annual O&M cost, $M	1% times capital+ $10/ton haul	0.1	1.7	1.7	0.3	5.1	2.4	2.0	0.9	2.2	0.4	1.8	0.9
Cost Summary														
Capital Cost, $M			8.1	162.6	131.2	18.7	369.3	154.5	114.2	129.0	116.3	121.4	134.7	133.2
Administrative, $M		25%	2.0	40.7	32.8	4.7	92.3	38.6	28.6	32.3	29.1	30.4	33.7	33.3
Contingency, $M														
	Project contingency	25%	2.0	40.7	32.8	4.7	92.3	38.6	28.6	32.3	29.1	30.4	33.7	33.3
	Process contingency	5%	0.4	8.1	6.6	0.9	18.5	7.7	5.7	6.5	5.8	6.1	6.7	6.7
Total Capital Cost for Scheme, installed, $M			12.6	252.0	203.4	28.9	572.5	239.5	177.1	200.0	180.2	188.2	208.8	206.4
Total Annual O&M Cost for Scheme, $M/year			1.1	21.8	17.8	2.5	49.4	27.7	14.6	13.1	16.3	5.3	19.6	13.7
Total Annual O&M Cost for Scheme, $/kgal			3.0	3.0	4.9	6.8	6.8	7.6	4.0	3.6	4.5	1.5	5.4	3.7
Annualized Capital, O&M Cost, $M/year		20 years @ 8% interest	2.4	47.4	38.5	5.4	107.7	52.1	32.6	33.5	34.7	24.5	40.9	34.7
Recovery of Water from Concentrate, %			93.9	94.0	87.9	81.0	81.3	85.5	90.0	87.0	90.8	87.4	87.6	86.5

APPENDIX 3

PROCESS STEP CONTRIBUTIONS TO CAPITAL AND OPERATING COSTS

A3.1 INTRODUCTION

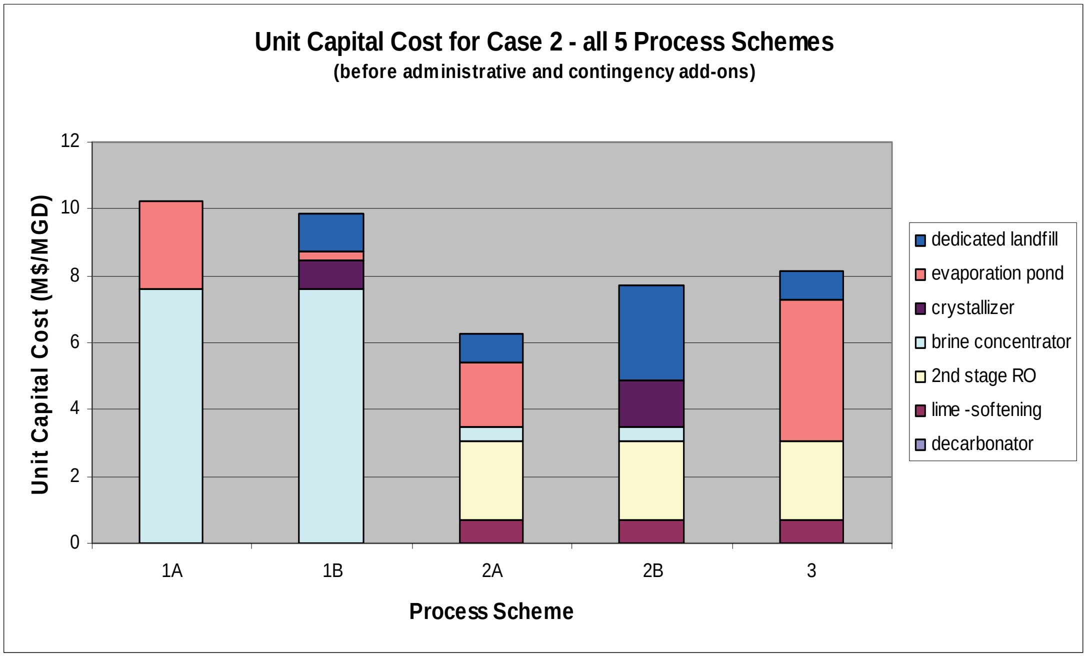
Total capital and operating costs include contributions from individual processing steps. The following bar charts break the equipment-related capital and operating cost values into process step contributions. Administrative and contingency costs are not included in the capital cost. To facilitate comparison of charts for different cases and different processing schemes, unit capital and unit operating costs are used.

The bar chart for Case 1 was given in Chapter 5 as Figure 5.3. Bar charts for Cases 2–5 are provided in this appendix as follows:

- Case 2: Fig. A3.1
- Case 3: Fig. A3.2
- Case 4: Fig. A3.3
- Case 5: Fig. A3.4

The bar charts for Cases 6–12 are given in terms of the various processing schemes. The bar chart for processing Scheme 1A (Cases 6–12) was given in Chapter 5 as Figure 5.6. Bar charts for the other processing schemes are provided in this appendix as follows:

- Processing Scheme 1B: Fig. A3.5
- Processing Scheme 2A: Fig. A3.6
- Processing Scheme 2B: Fig. A3.7
- Processing Scheme 3: Fig. A3.8



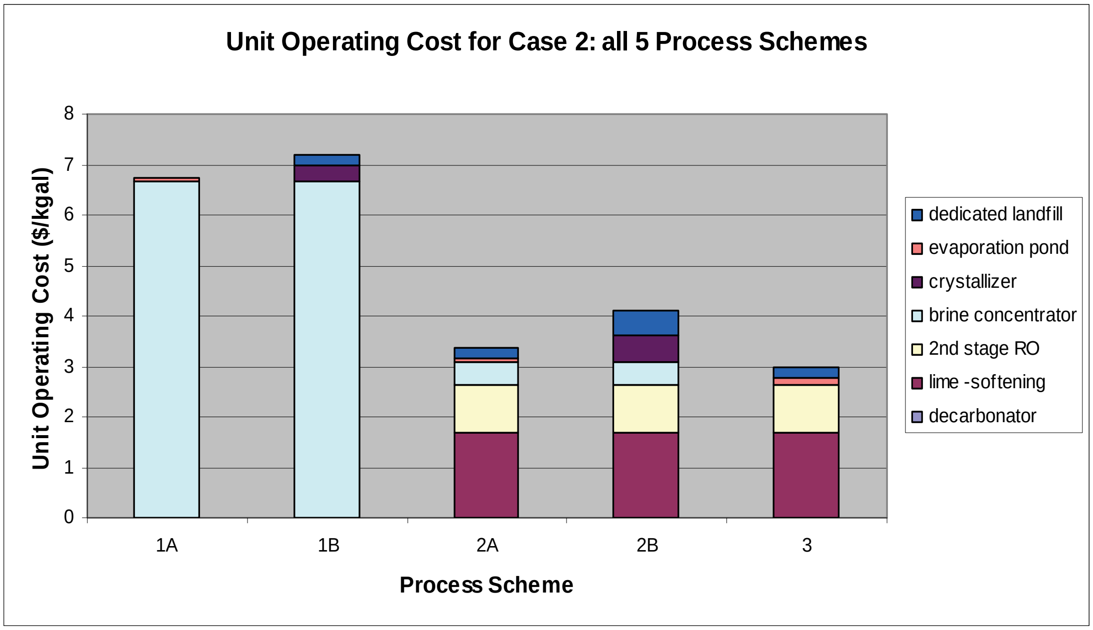
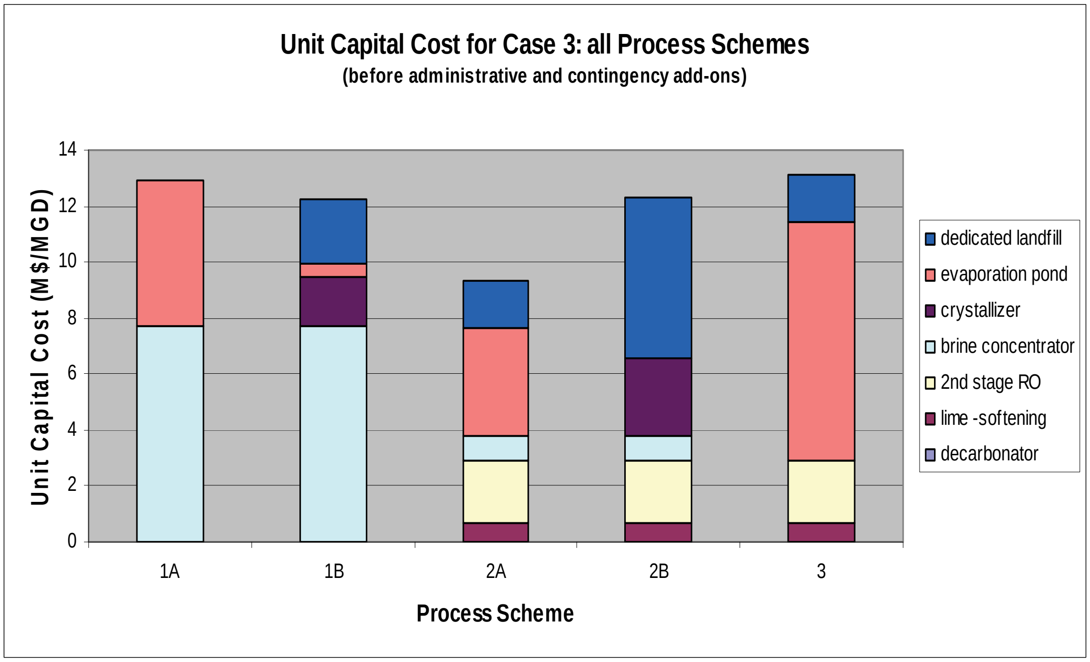


Figure A3.1. Process step contributions to unit capital and operating costs, Case 2.



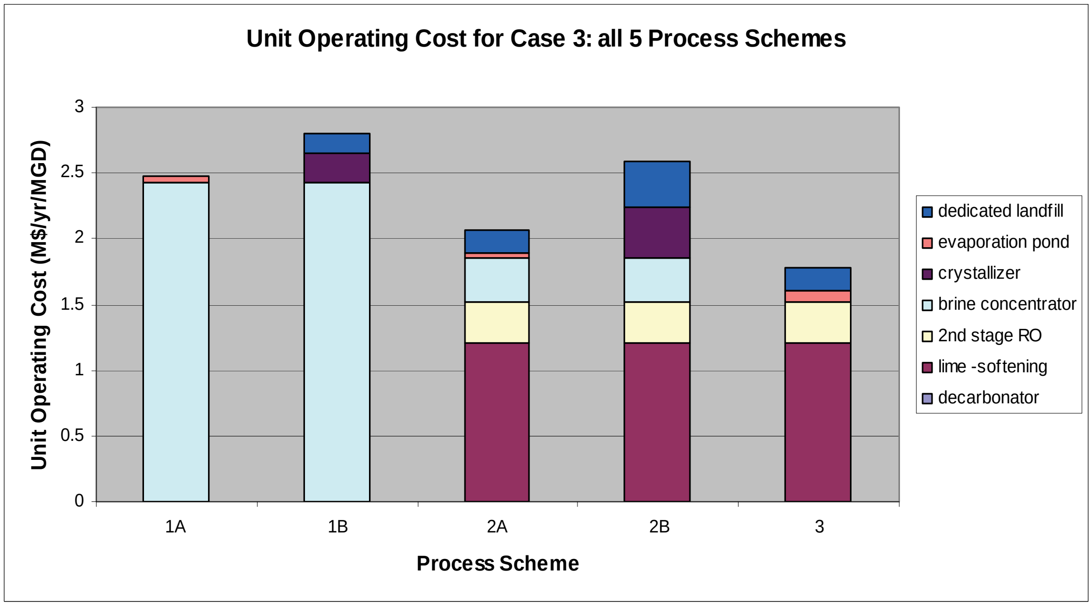
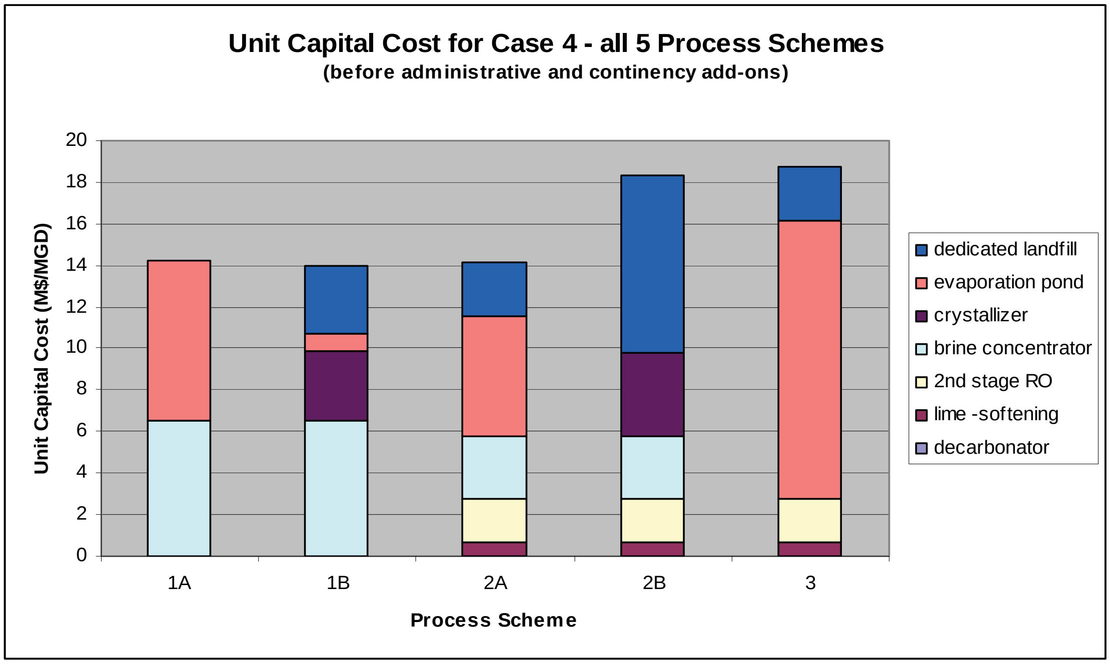


Figure A3.2. Process step contributions to unit capital and operating costs, Case 3.



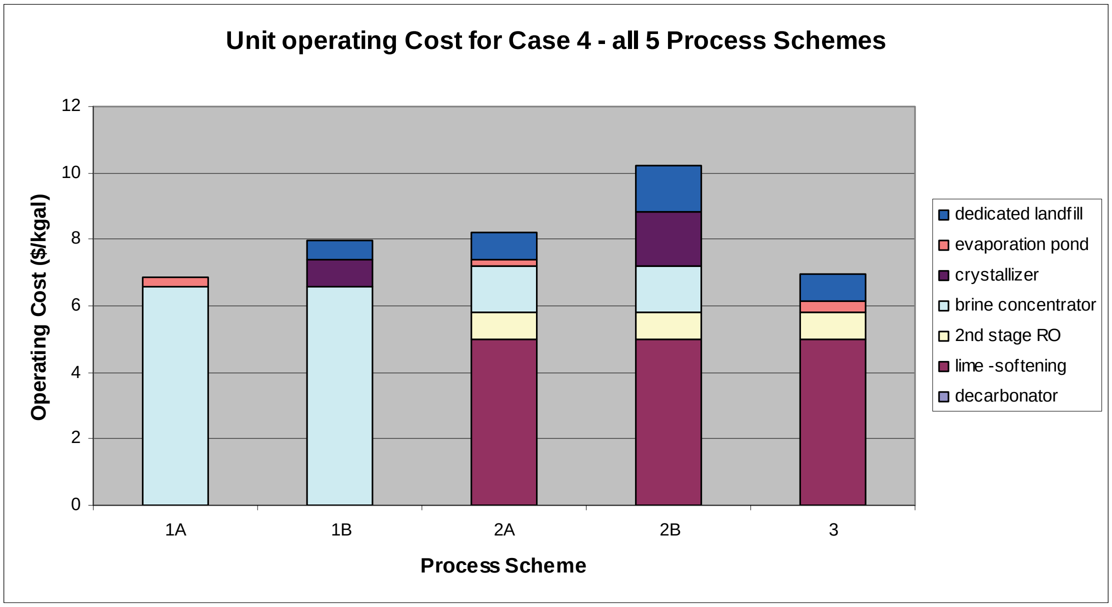
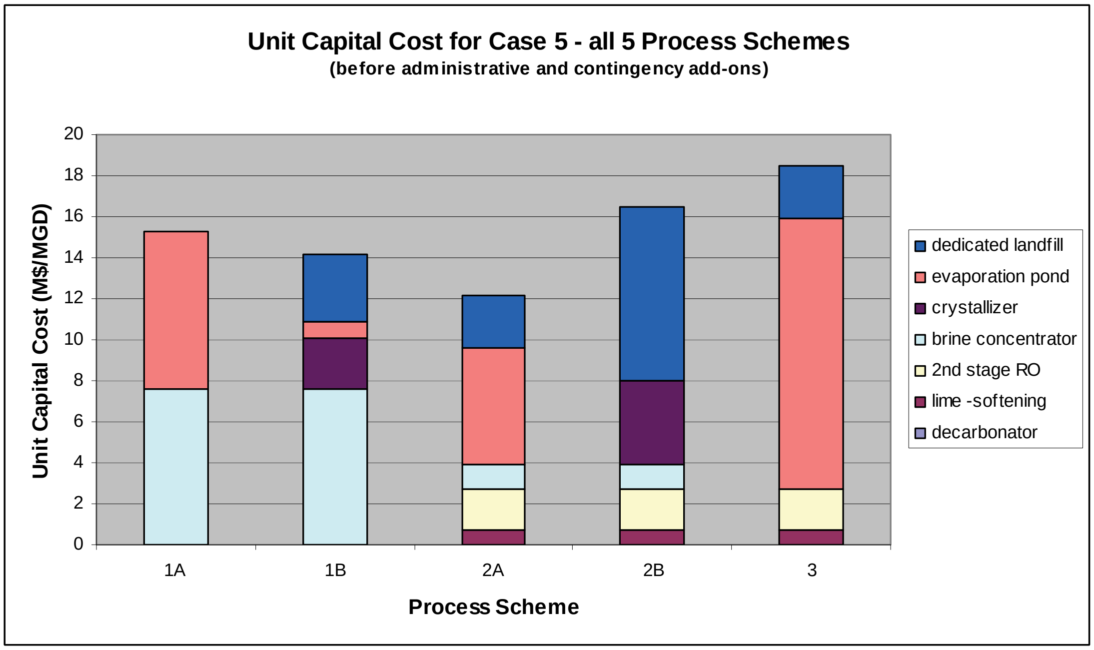


Figure A3.3. Process step contributions to unit capital and operating costs, Case 4.



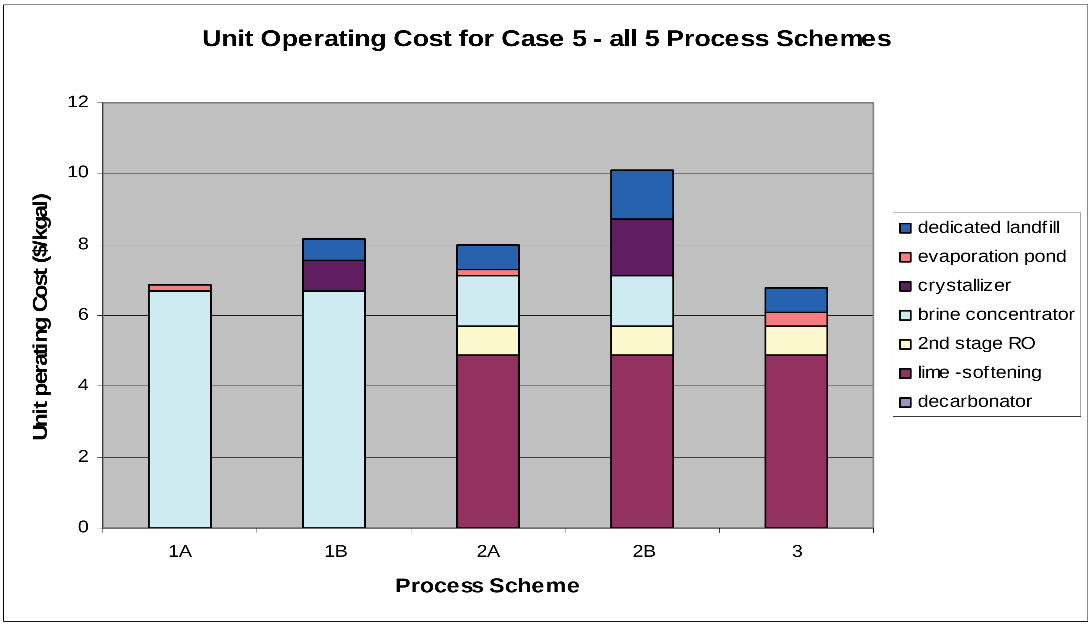
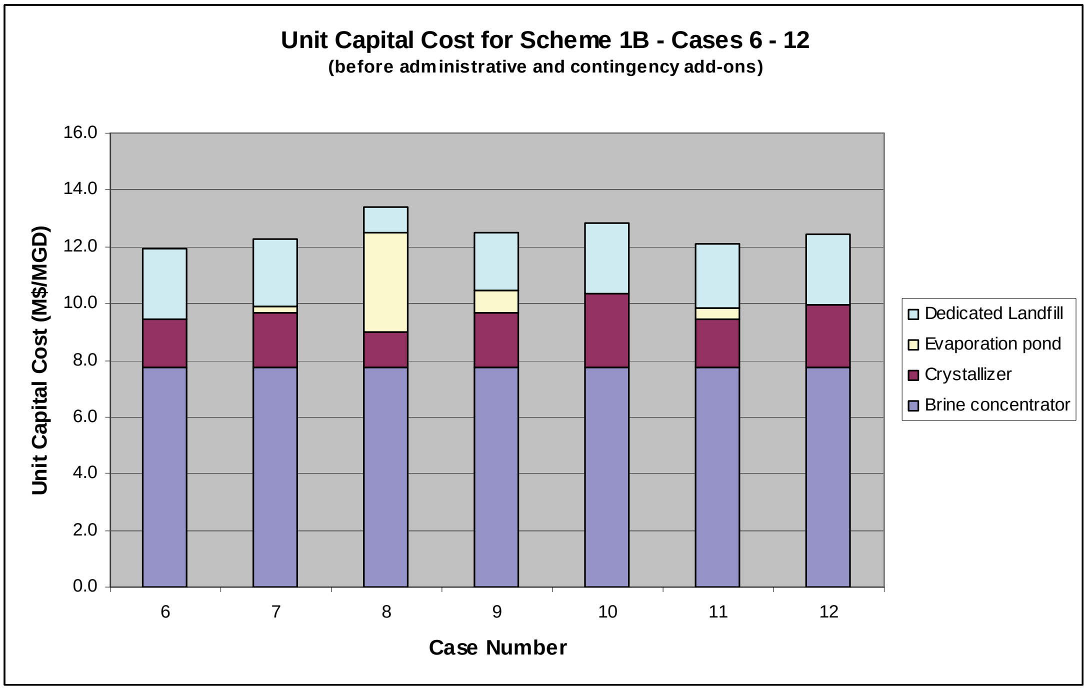


Figure A3.4. Process step contributions to unit capital and operating costs, Case 5.



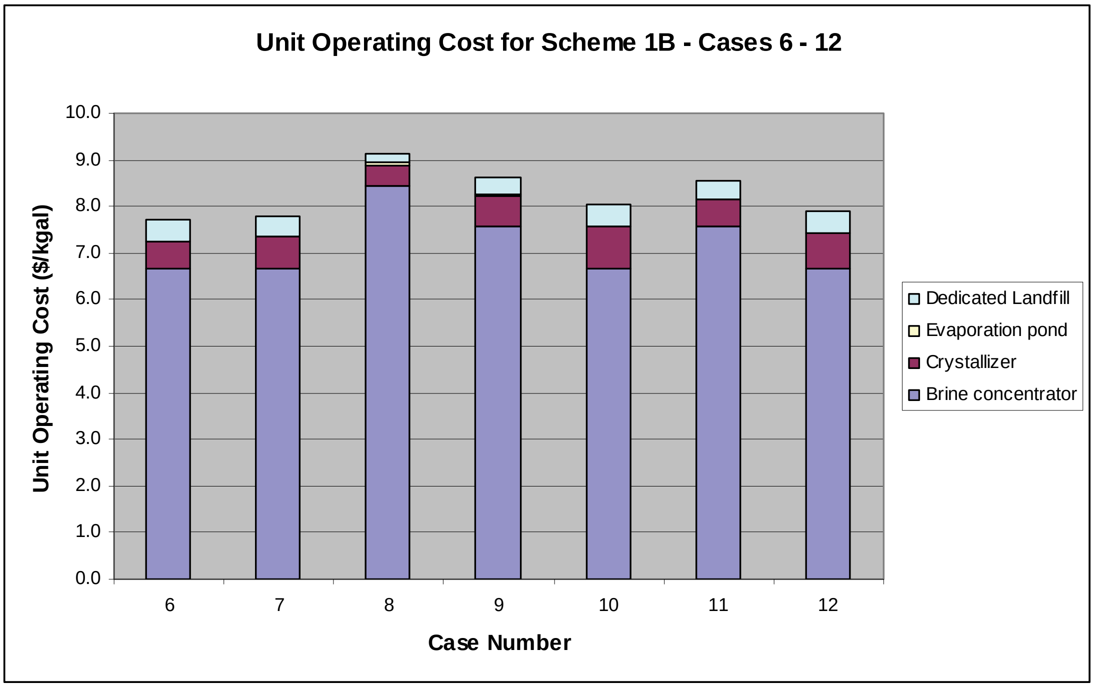


Figure A3.5. Process step contributions to unit capital and operating costs, Scheme 1B.

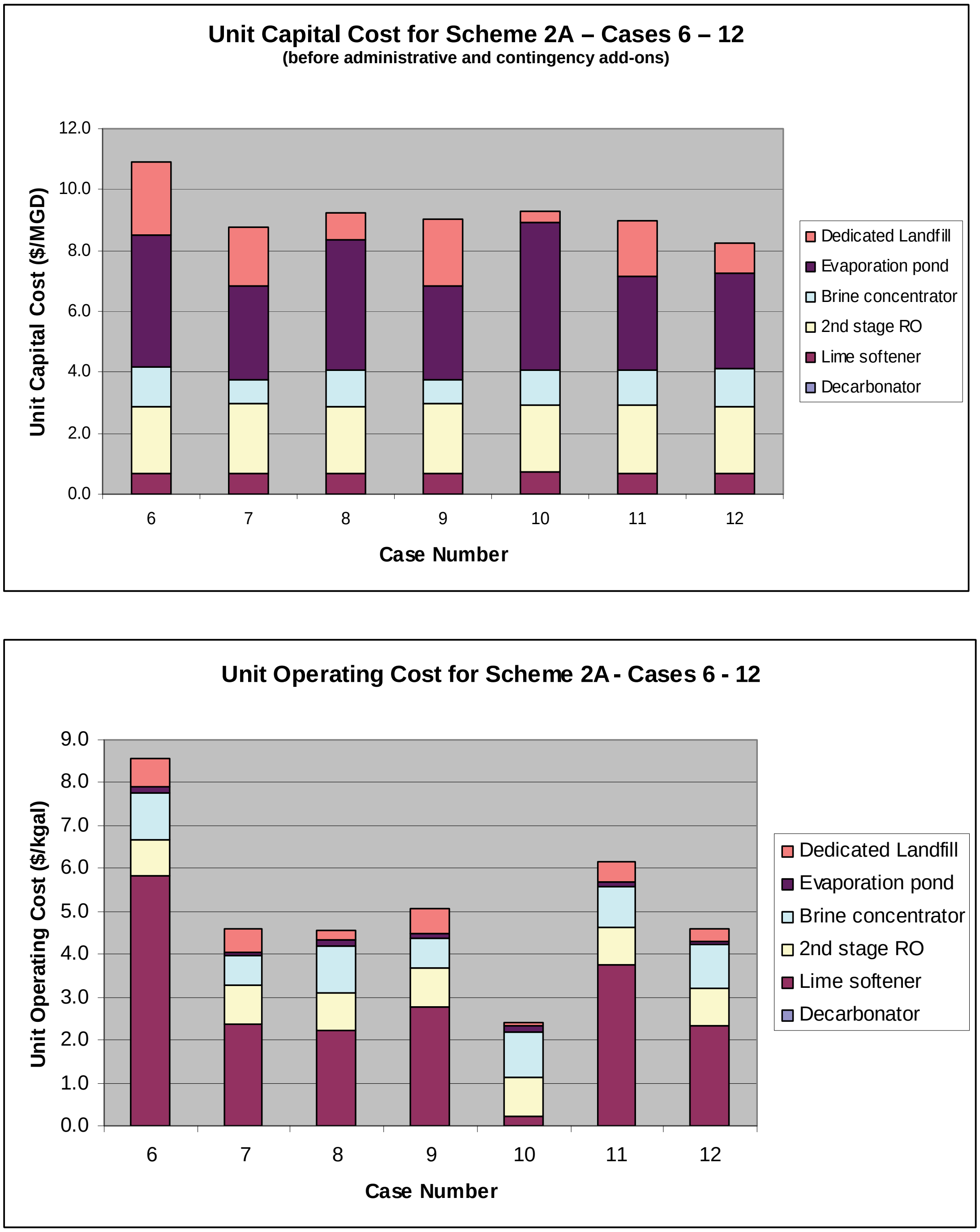
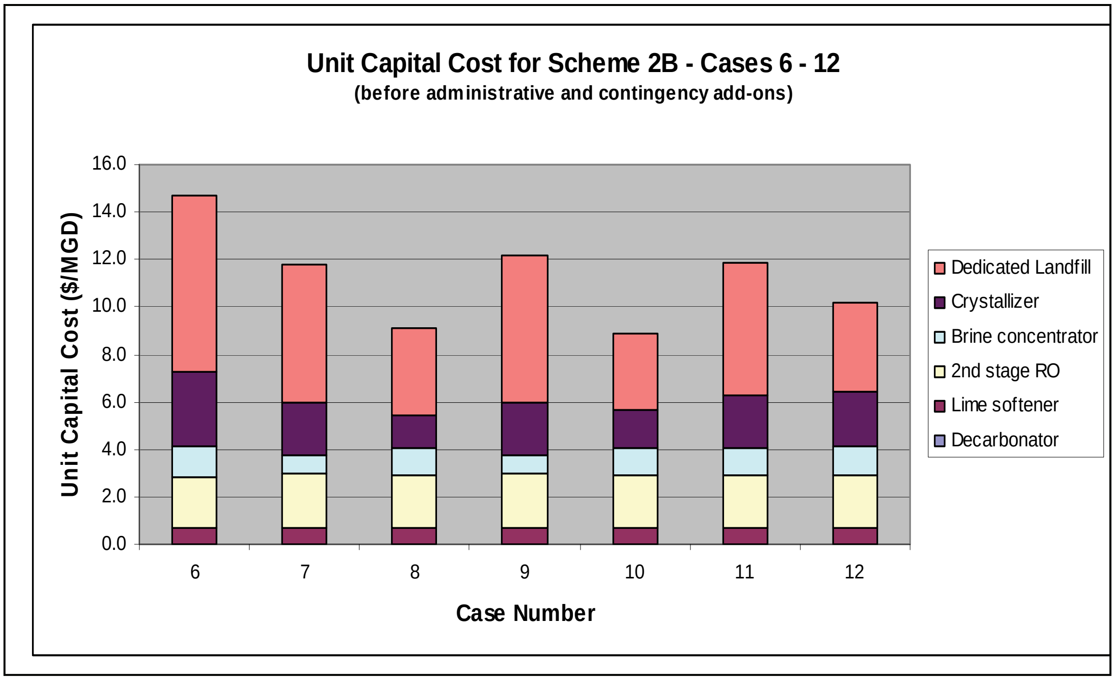


Figure A3.6. Process step contributions to unit capital and operating costs, Scheme 2A.



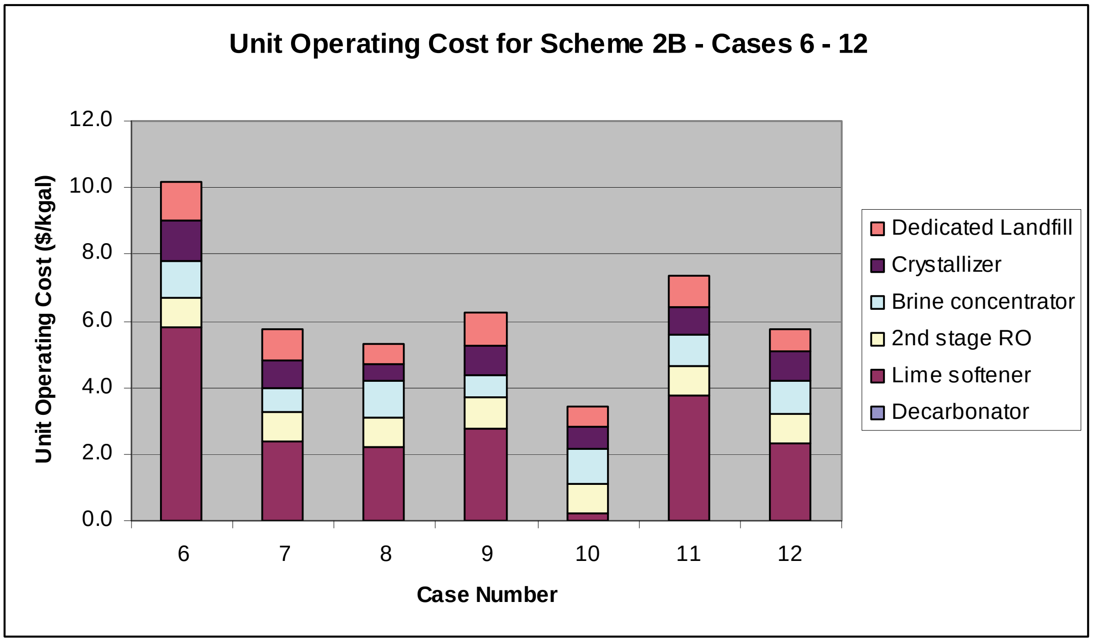
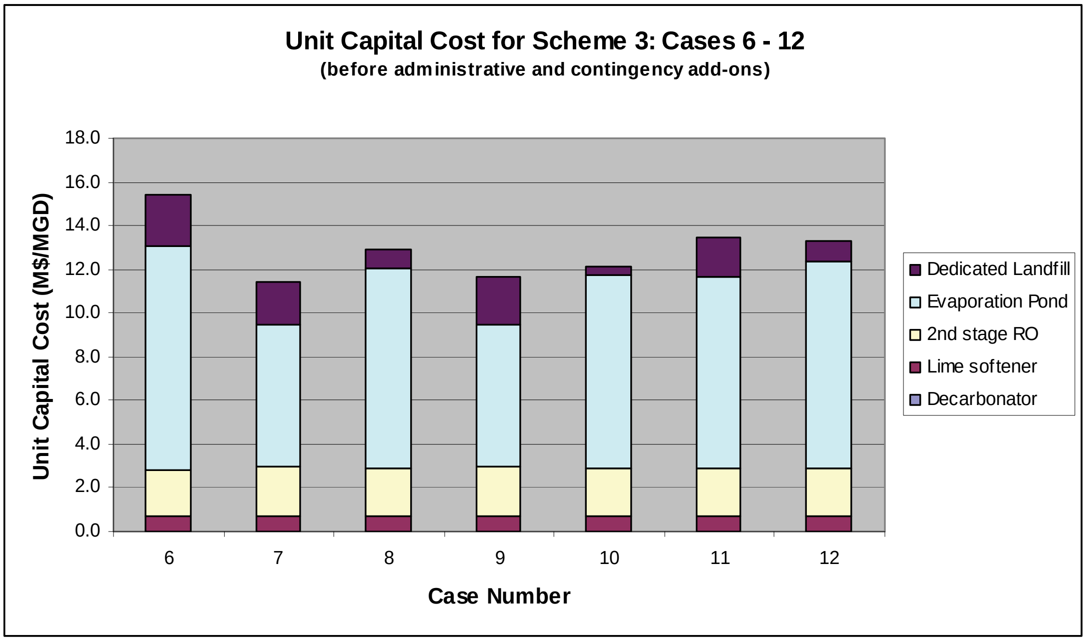


Figure A3.7. Process step contributions to unit capital and operating costs, Scheme 2B.



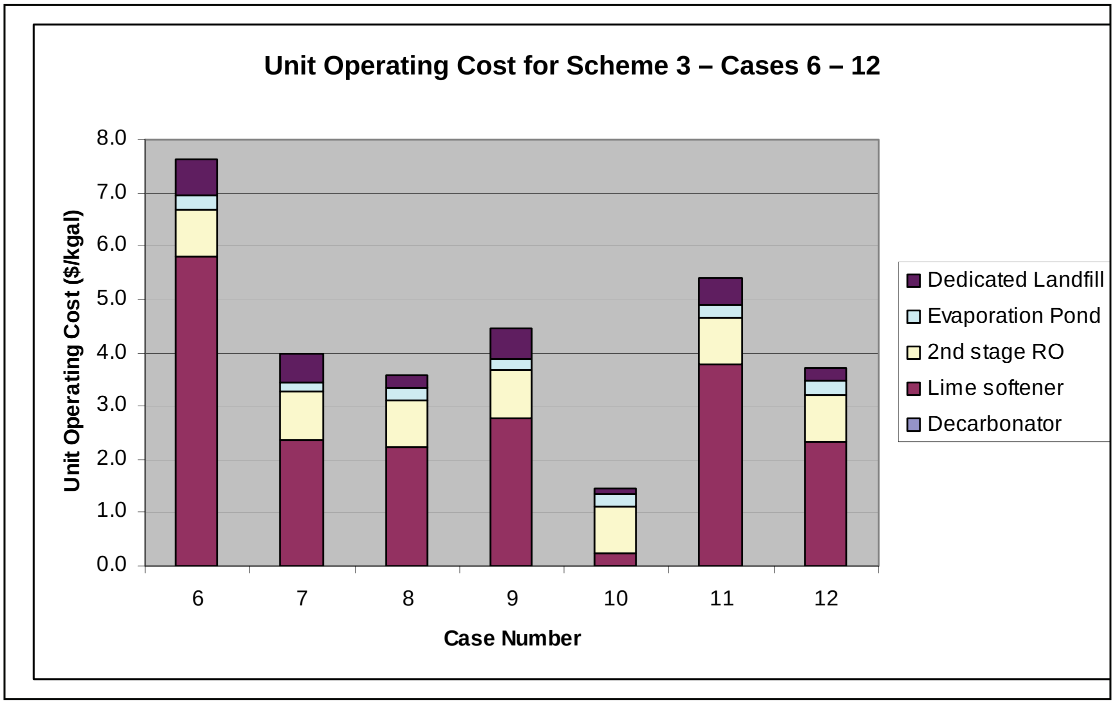


Figure A3.8. Process step contributions to unit capital and operating costs, Scheme 3.

APPENDIX 4

OBSERVATIONS AND EXPLANATIONS OF RESULTS FOR FIGURES 5.1, 5.2, 5.4, AND 5.5

A4.1 INTRODUCTION

The anomalies or variances from “average” values of capital, operating, and annualized costs in Figures 5.1, 5.2, 5.4, and 5.5 are explained by reference to the variables and the resulting effects on processing steps. This appendix provides a detailed analysis of each processing scheme and case. The analysis and explanations reflect that each processing step in a processing scheme can be affected in a different way and to a different degree by a change in salinity and flow (Cases 1–5) or by chemical composition (Cases 6–12). As the observations and explanations show, it is difficult to provide a simple explanation as to why the ups and downs of Figures 5.1, 5.2, 5.4, and 5.5 occur, and it is difficult to generalize and to provide simple “rules of thumb” to explain the results.

A4.2 OBSERVATIONS AND EXPLANATIONS FOR CASES 1–5

In Cases 1–5 the only variables are the salinity and flow volume while the chemical composition is the same for all five cases. The anomalies or variances from average values of capital, operating, and annualized costs in Figures 5.1 and 5.2 are explained by reference to how the variables affect the processing steps.

A4.2.1 Case 1

General observations

- Differs from base case (Case 3) conditions (8000 mg/L and 10 MGD): 4000 mg/L and 1 MGD
- Lowest salinity; lowest flow
- Second-lowest annualized cost among all cases for Schemes 1A, 1B, and 3
- Third-lowest annualized cost among all cases for Schemes 2A and 2B

Chemistry

- Relative to Cases 6–12, constituents are neither high nor low
- Chemistry is the same for Cases 1–6

Explanation

- (Scheme 1A) With low-salinity feed yet roughly the same concentration factor in the brine concentrator, volume reduction is much greater; thus, a smaller volume goes to evaporation ponds, and this leads to a lower capital cost for the evaporation pond.
- (Scheme 1B) For the same reason, a smaller volume goes to the crystallizer; the purge volume is lower.
- (Scheme 2A) For the same reason, a smaller volume goes to the RO, resulting in a higher concentration factor at the RO step and thus a lower concentrate volume going to the brine concentrator and subsequent processing steps; this leads to lower capital, operating, and annualized costs.
- (Scheme 2B) Similar reason as for 2A
- (Scheme 3) Similar reason as for 2A and 2B

A4.2.2 Case 2

General observations

- Differs from base case (Case 3) conditions (8000 mg/L and 10 MGD): 4000 mg/L and 20 MGD
- Lowest salinity; highest flow
- Lowest annualized cost among all cases for 1A, 1B, 2A, and 3
- Second-lowest annualized cost among all cases for 2B

Chemistry

- Relative to Cases 6–12, constituents are neither high nor low
- Chemistry is the same for Cases 1–6

Explanation

- Reasons are the same as for Case 1 with the additional slight economy of scale due to larger volumes

A4.2.3 Case 3 (Base Case)

General observations

- Average results for processing Schemes 1A, 1B, 2A, and 3
- Relatively high annualized cost for 2B

Chemistry

- Relative to Cases 6–12, constituents are neither high nor low
- Chemistry is the same for Cases 1–6

Explanation

- When compared to Cases 6–12 (same salinity and size), Case 3 has the second-highest flow into the crystallizer and thus the second-highest capital cost, third-highest operating cost, and second-highest annualized cost of these cases.

A4.2.4 Case 4

General observations

- Differs from base case (Case 3) conditions (8000 mg/L and 10 MGD): 12,000 mg/L and 1 MGD
- Highest salinity; lowest flow
- Second-highest unit annualized cost among all cases for 1A and 1B
- Highest unit annualized cost among all cases for 2A, 2B, and 3

Chemistry

- Relative to Cases 6–12, constituents are neither high nor low
- Chemistry is the same for Cases 1–6

Explanation

- Highest salinity means lower concentration factors in the brine concentrator (1A and 1B) and RO step (2A, 2B, and 3) and thus higher flows going to subsequent processing steps.
- Highest salinity means higher chemical cost per MGD treated in the lime softening step (2A, 2B, and 3).
- Highest salinity means larger evaporation ponds (1A), larger crystallizer (1B), and larger brine concentrator (2A, 2B, and 3).
- This results in higher unit capital and unit operating costs and thus unit annualized cost.

A4.2.5 Case 5

General observations

- Differs from base case (Case 3) conditions (8000 mg/L and 10 MGD): 12,000 mg/L and 20 MGD
- Highest salinity; highest flow
- Second-highest unit annualized cost among all cases for 1A, 1B, 2A, 2B, and 3

Chemistry

- Relative to Cases 6–12, constituents are neither high nor low
- Chemistry is the same for Cases 1–6

Explanation

- Similar to Case 4; the difference is a much higher volume
- Reasons are the same as for Case 4 with the additional slight economy of scale due to larger volumes

A4.3 OBSERVATIONS AND EXPLANATIONS FOR CASES 6–12

In Cases 6–12 the only variable is the chemical composition, as salinity and flow volume are the same for all seven cases. The anomalies or variances from average values of capital, operating, and annualized costs in Figures 5.4 and 5.5 are explained by reference to the chemical composition and the resulting effects on processing steps.

A4.3.1 Case 6

General observations:

- Average results for schemes 1A and 1B
- Highest capital, operating, and annualized costs for schemes 2A, 2B, and 3 for Cases 6–12

Chemistry:

- High Ca, low HCO_3 result in highest Ca and HCO_3 level of cases with high Ca
- Chemistry is the same for Cases 1–6

Explanation:

- Case 6 has the lowest solids reduction (TDS) at the lime softening step; this results in the highest feed TDS into the RO, which gives the lowest concentration factor in the RO and results in the largest flow into the brine concentrator; this leads to the largest flow into the crystallizer (2B).
- Case 6 has the highest soda ash dose at the lime softening step, which leads to the largest amount of solids produced.
- Case 6 has the highest total capital cost due to high brine concentrator, crystallizer, and landfill capital costs.
- Case 6 has a high operating cost due to the greatest amount of chemicals added.
- This leads to the highest annualized cost for Schemes 2A, 2B, and 3.

A4.3.2 Case 7

- Average results for all schemes

A4.3.3 Case 8

General observations

- Average results for Schemes 2A and 3
- Relatively low annualized cost for 2B
- Highest capital cost for Scheme 1B
- Highest operating cost for Scheme 1A and 1B
- Highest annualized cost for Schemes 1A and 1B

Chemistry

- Relatively low Ca and HCO_3
- Highest Na and Cl

Explanation

- The decreased capital cost is due to the highest final TDS from the brine concentrator and thus the lowest flow and size of the crystallizer (Scheme 2B).
- Case 8 results in the highest crystallizer purge due to high soluble chlorides; this leads to the highest evaporation pond cost.
- Case 8 has the highest energy (per kilogallon) for the brine concentrator.
- The high capital cost is due to increased evaporation pond acreage.
- The high brine concentrator operating cost for Schemes 1A and 1B leads to the highest annualized cost for Schemes 1A and 1B.

A4.3.4 Case 9

General observations

- Average results for Schemes 1A, 1B, 2A, and 3
- Relatively high annualized cost for Scheme 2B

Chemistry

- Second-highest amount of solids produced from lime softening
- Second-highest landfill solids from lime softening and crystallizer
- High Ca, high HCO_3

Explanation

- Case 9 has the lowest concentration factor in the brine concentrator; this leads to a large crystallizer; high landfill costs result from high solids from lime softening and crystallizer steps.

A4.3.5 Case 10

General observations

- Average results for Scheme 1B
- Lowest annualized cost for Schemes 1A, 2A, 2B, and 3

Chemistry

- Lowest Ca
- Lowest Ca/HCO_3 ratio
- Lowest SO_4, TDS
- Highest HCO_3
- Highest (Na + Cl)/Ca

Explanation

- For Scheme 1A, the low concentration factor in the brine concentrator leads to a high initial flow into evaporation pond. However, the low salinity of the brine going into

the evaporation pond leads to a low solids buildup rate and no need to cover ponds and replace them during the life of the plant (in many other cases, the evaporation pond acreage is increased significantly by the need to cover ponds and replace them during the life of the plant).
- For Schemes 2A, 2B, and 3 there is a low chemical need for the lime softening step and thus low solids produced, which results in the lowest operating cost for Schemes 2A, 2B, and 3.
- Schemes 2A and 3 have average capital costs; Scheme 2B has the lowest capital cost.
- The net result is the lowest annualized cost for 2A, 2B, and 3.

A4.3.6 Case 11

General observations
- Average results for Schemes 1A and 1B
- Relatively high annualized costs for Schemes 2A, 2B, and 3
- Average capital cost

Chemistry
- High Ca
- Medium HCO_3

Explanation
- Case 11 has the second-highest chemical need at the lime softening step, which leads to the second-highest operating cost.
- The result is the second-highest annualized cost.

A4.3.7 Case 12

General observations
- Average results

A4.4 OTHER OBSERVATIONS AND EXPLANATIONS

1. Schemes 2A, 2B, and 3 (which have lime softening steps) are much more sensitive to salinity than Schemes 1A and 1B.
    - *Explanation:* A higher salinity for the same chemical composition means higher calcium and magnesium levels. This leads to a greater softening load and thus greater use of chemicals and a greater amount of solids produced.

2. Schemes 2A, 2B, and 3 (which have lime softening steps) have a much larger range of cost values than Schemes 1A and 1B.
    - *Explanation:* This is due both to the effect of salinity (Cases 1–5) just mentioned in point 1 and to the effect of chemical composition (Cases 6–12). The latter results in a wide range of chemical costs and solids disposal cost (landfill). Thus, the difference is primarily due to the operating cost and solids produced at the lime softening step.

3. Scheme 2B undergoes a much wider swing in cost values than the other schemes.
    - *Explanation:* Scheme 2B is the only lime softening case (2A, 2B, and 3) that also has a crystallizer step. Both the lime softening step and the crystallizer produce solids for landfill, and the landfill cost for Scheme 2B is much higher than for other schemes.

4. Scheme 2A is consistently the lower annualized cost option among Schemes 2A, 2B, and 3.
   - *Explanation:* Scheme 2A has a lower capital cost (significantly lower in most cases) than Schemes 2B and 3. All three schemes have the same lime softening and second-stage RO costs. The additional capital cost of Scheme 2A is due to the brine concentrator and evaporation ponds. The additional capital costs of Schemes 2B and 3, respectively, are due to the (same) brine concentrator, a crystallizer, and a landfill (scheme 2B) or evaporation ponds and landfill (scheme 3). Unit capital costs for a brine concentrator or crystallizer are less than the unit capital cost for evaporation ponds. The high costs of brine concentrators and crystallizers come from operating costs. The operating costs for Schemes 2A and 2B are higher than that for Scheme 3. The operating cost for Scheme 2A is consistently second of the three. The annualized cost, dependent on both capital and operating costs, is lower for Scheme 2A than for Schemes 2B and 3.

5. Scheme 1A has a slightly lower annualized cost, in all cases, than Scheme 1B.
   - *Explanation*: Although capital costs for these two schemes are similar, they are slightly higher for Scheme 1A. This is because the capital cost for the (larger) evaporation ponds of Scheme 1A is slightly higher than the capital costs for the crystallizer, landfill, and (smaller) evaporation ponds of Scheme 1B. This, however, is compensated for by Scheme 1A having a lower operating cost than Scheme 1B. The operating cost for Scheme 1A is mostly due to the brine concentrator, and Scheme 1B has this same operating cost plus operating cost contributions from the crystallizer and the landfill, neither of which are part of the Scheme 1A operating cost. With a similar capital cost and slightly lower operating cost, the annualized cost for Scheme 1A is lower than for Scheme 1B.

6. Scheme 2B has a higher annualized cost in all cases than Schemes 2A and 3.
   - *Explanation:* In most cases the capital cost for Scheme 2B is lower or similar to that for Scheme 3. In all cases it is higher than for Scheme 2A. Scheme 2B has the highest operating cost of the three schemes. The high operating cost results from the high energy costs for the brine concentrator and crystallizer. Scheme 2A does not have the crystallizer contribution, and Scheme 3 has neither the brine concentrator nor crystallizer contribution. The resultant annualized cost for Scheme 2B is higher in all cases than that for Schemes 2A and 3.

7. Scheme 2A has the lowest unit annualized cost in 9 of the 12 cases.
   - *Explanation:* The schemes utilizing a crystallizer, Schemes 1B and 2B, never have the lowest unit annualized cost. This is due to the high energy cost of the crystallizer and the cost of landfilling the solids produced. Scheme 2A takes advantage of the second-stage RO prior to the brine concentrator while avoiding the high energy costs and solids disposal costs associated with the crystallizer (Schemes 1B and 2B). Over the wide range of chemical compositions (Cases 6–12), in only one case (Case 6) did Scheme 2A not have the lowest unit annualized cost.

8. What happened in the other three cases where Scheme 2A did not have the lowest unit annualized cost?
   - *Explanation:* Scheme 2A had the lowest capital cost in these three cases. In Case 2, Scheme 3 had a lower operating cost and a similar capital cost (but slightly higher) than Scheme 2A; Scheme 3 had the lowest unit annualized cost.

   - For Case 4, Scheme 2A had an average operating cost, with Schemes 1A and 3 having very similar and lower operating costs. The net result was that Scheme 1A had the lowest unit annualized cost.
   - For Case 6, both Schemes 1A and 1B had lower operating costs than Scheme 2A. Both Schemes 1A and 1B had lower unit annualized costs than Scheme 2A.
   - As to why these situations occurred, the reader is referred to the discussions of individual cases above.

9. The variability in operating costs is much greater than the variability in capital costs.
   - Capital cost is much less sensitive to feed water composition than operating cost.
   - The primary variable affecting unit capital cost is salinity, which determines the degree of concentration taking place in the desalination steps (second-stage RO, brine concentrator, crystallizer) and thus the volume of brine moving on to the subsequent process step. Higher salinity results in lower volume reductions in the desalination steps and larger equipment sizes for subsequent process steps.

APPENDIX 5

GEO-PROCESSORS' SELECTIVE SALT RECOVERY TECHNOLOGY

A5.1 INTRODUCTION

Selective salt recovery has received increasing attention in the past few years due to:

- The growing need for alternative concentrate management options
- The potential benefits of selective salt recovery in:
  - avoiding negative environmental impacts associated with concentrate, brine, and solids disposal
  - decreasing operating costs through the sale of recovered salts
  - providing a means of approaching the ideal of sustainability
  - decreasing the CO_2 footprint of the desalination process through removal of carbonate species

The timing of the attention is in large part associated with the visibility Geo-Processors USA, Inc. has attained since 2003 through presentations at various technical conferences. Geo-Processors is an Australia-based company established in Sydney in 1991; they opened an office in Los Angeles in 2004.

This Appendix provides details of Geo-Processors' approach to achieving zero liquid discharge (ZLD) outcomes from the treatment of concentrate, which involves sequential salt recovery using their patented SAL-PROC^{TM} process. The information provided is indicative only and should not be taken as a detailed evaluation of Geo-Processors' treatment technologies.

Due to the complexity and proprietary nature of the technology, it was necessary to engage Geo-Processors in providing a substantial portion of the analysis conducted and presented in this appendix. While this input was necessary, it means the information was not developed on the same basis and assumptions as the information presented in Chapter 5. Thus, the estimated costs cannot be compared directly with those of the ZLD processes presented in Chapter 5. However, the information provides considerable insight into the general nature and costs involved with this technology.

A5.2 ASSUMPTIONS AND STUDY APPROACH

The chemical compositions of the 12 cases presented in Chapter 4 were evaluated to establish the extent of similarity in water types using molar ratios of Cl, HCO_3, and SO_4 elements. Four cases were then selected (Cases 1, 6, 9, and 11) for assessing the techno-economics of a common ZLD process involving selective salt recovery steps, using the assumptions defined in Chapter 4. Case 1 was then used as the basis for a more in-depth preliminary conceptual process design, cost, and revenue estimate. Water quality data assigned to each case were used for the SAL-PROC^{TM} model simulation, on which a number of technically possible process steps for sequential extraction of salts were identified. The output data from the SAL-PROC^{TM} model were then used to assess and define a number of integrated treatment systems involving volume reduction and water and salt recovery steps leading to ZLD. All the treatment systems selected and discussed herein utilize SAL-PROC^{TM} process steps involving multistep chemical reactions and crystallization, with lime and soda ash reagents, followed by mechanical washing, dewatering, and drying components. Apart from the linkage of SAL-PROC^{TM} with conventional volume reduction methods (brine concentrator and crystallization ponds), secondary and tertiary RO units are used for initial volume reduction steps.

No particular attention has been paid to the indirect benefits from application of the SAL-PROC^{TM} technology, in addition to its economic features. These complementary benefits could be multifaceted, but overall they may be summarized as follows:

- Its effectiveness in removing the total salt load from the residuals cycle, largely in the form of useful products
- Its effectiveness in combined salt load and volume reduction
- Recovery of more finished water by using secondary RO desalination
- Reduction of operational and environmental footprints
- Reduction of the potential CO_2 footprint of desalination plants (and other plants) via conversion of HCO_3 to $CaCO_3$

Depending on the chemistry of the input water and scope of treatment, the number and arrangement of process components in Geo-Processors' treatment systems and output values may change from one concentrate type to another. However, to provide a comparative basis, a simplified block flow diagram of the most comprehensive ZLD treatment system, which applies to most cases and integrates SAL-PROC^{TM} process steps with a number of conventional water recovery and volume reduction methods, is presented in Figure A5.1.

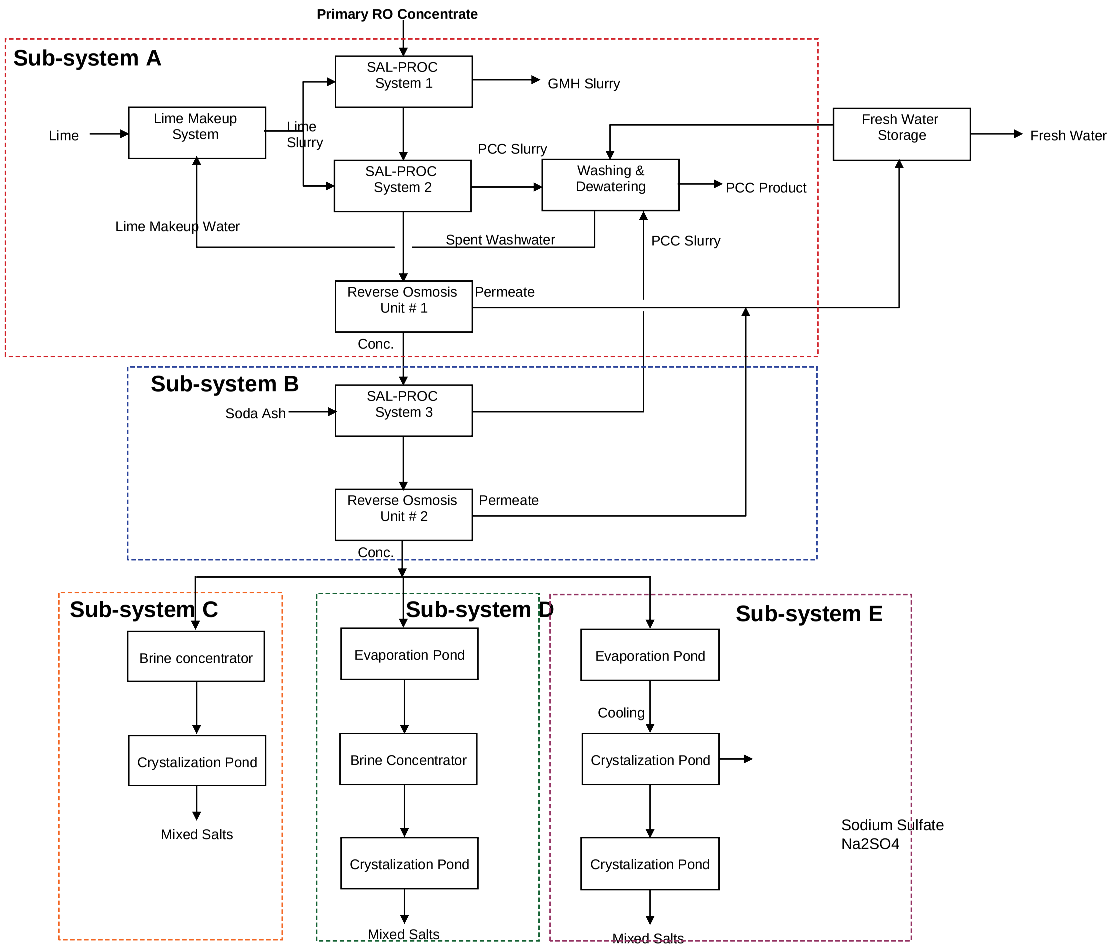


Figure A5.1. Generic block flow diagram for ZLD options ABC, ABD, and ABE for all cases.

A5.3 METHODS

A5.3.1 Desktop Modeling

RO concentrate quality data and flows were used for the SAL-PROC^{TM} process simulations, using a computer modeling software program developed by Geo-Processors for simulating the mineral precipitation phase with lime and soda ash reagents under variable TDS salinity and dissolved ionic (molar) ratios. SAL-PROC^{TM} is an innovative process for sequential or selective recovery of valuable by-products from saline streams and wastewater minimization. The technology includes components that are based on well-established chemical processing and mineral processing principles. Geo-Processors' licensed technologies have been extensively tested in a number of locations in Australia. Further description of the SAL-PROC^{TM} process and various linkage options with other volume reduction and salt recovery methods are available on the company's website, www.geo-processors.com.

The mineral recovery steps from the desktop modeling identified several process routes, with three selected as the most appropriate ZLD options for the nominated concentrate streams. As shown schematically in Figure A5.1, the selected treatment systems are basically comprised of two or three components (subsystems). All options include Subsystem A, which involves pretreatment by controlled lime dosing to selectively remove the dissolved sulfate and bicarbonate elements in the form of useful products. These steps are followed by secondary RO desalination to further recover water and reduce the effluent volume without a major change in the molar ratio of the dissolved ions. Downstream, each treatment system may incorporate one or two more components (identified as Subsystems B, C, D, and E) for additional salt recovery and water production, followed by thermo-mechanical volume reduction and crystallization steps. A block flow diagram showing the arrangement of the various subsystems in Geo-Processors' ZLD options is shown in Figure A5.2. The subsystems of Figure A5.1 are further described below with reference to the product streams expected from treatment of the nominated concentrate streams.

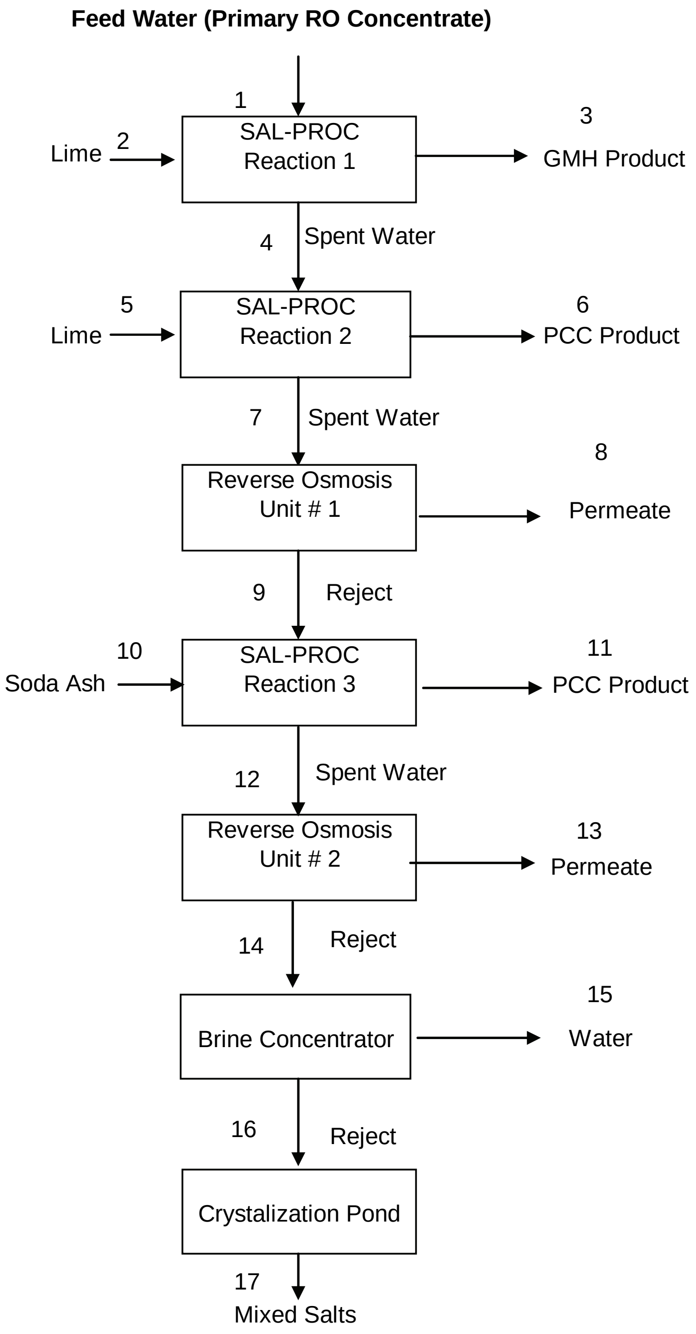


Figure A5.2. Generic block diagram showing the basic ZLD process steps, option ABC.

A5.3.1.1 Subsystem A

Subsystem A consists of taking the RO concentrate stream through two reaction steps according to the SAL-PROC^{TM} process. The first reaction step involves mixing lime slurry with the RO concentrate under controlled pH, dosing rates, and mixing conditions. From this reaction step, the treated concentrate is directed to a settling pond or a holding tank before dewatering and washing to produce a beneficial product consisting of either gypsum-magnesium hydroxide (GMH) or magnesium hydroxide (MH) product, both of which have numerous industrial, agricultural, and environmental applications.

Depending on the bicarbonate concentration level in the primary feed, the supernatant or spent water from the magnesium hydroxide pond or holding tank may be transferred to the second step of the reaction with lime slurry in a chemical reactor, under controlled dosing conditions. The slurry thus produced is then dewatered in thickeners and washed prior to further dewatering in a filter press or bag filter to produce a cake having between 40–50 % dry solids content. The filter cake is a commercial-grade precipitated calcium carbonate (PCC) having applications to paper, dye, paint, and polyvinylchloride manufacturing. Geo-Processors has extensive experience with the recovery of PCC products through previous projects dealing with bicarbonate-rich ground waters. Where the bicarbonate concentration level in the primary feed is low (and hence there is a reduced risk of calcium carbonate precipitation in the secondary RO), the PCC recovery step is not performed.

From either the first or second reaction step, the thickener overflow is led into a spent water tank, where the water is processed via an RO unit (RO1). The RO unit has a nominal 65% recovery rate, with the permeate stream recovered as fresh water and the concentrate stream further treated in a mixing tank with soda ash in Subsystem B.

A5.3.1.2 Subsystem B

Where further water recovery using RO desalination is possible, the reject stream from the RO1 flows into the next SAL-PROC^{TM} process component comprised of a reaction vessel for further reduction of dissolved calcium by recovery of PCC product using soda ash reagent. The water after this reaction is processed via thickening, with the thickened slurry transferred to the wash tanks in the first SAL-PROC^{TM} component for producing PCC. The overflow from the thickeners is processed by a second RO unit (RO2) to produce two streams, consisting of a permeate (50–65% recovery) which is reclaimed as fresh water and a concentrate stream which is further processed in a downstream subsystem to achieve ZLD.

A5.3.1.3 Subsystems C, D, and E

From Subsystem B there are three alternatives (Subsystems C, D, and E) for further treatment of the reject from RO2. These three subsystems are briefly described below.

- **Subsystem C:** This subsystem assumes the reject from RO2 will flow into a thermal brine concentrator. The concentrator will produce a high-purity freshwater stream and bittern stream, which will be transferred to a crystallization pond for harvesting mixed salts.
- **Subsystem D:** This subsystem assumes that the reject from RO2 will flow first into evaporation ponds to benefit from high natural evaporation rates before the concentrate in the pond is transferred to a brine concentrator and then into a crystallization pond for harvesting mixed salts.
- **Subsystem E:** This subsystem assumes favorable climatic conditions and land availability for establishing and operating evaporation and crystallizer ponds (i.e.,

instead of thermal–mechanical volume reduction methods). Accordingly, reject from RO2 will flow into evaporation ponds, with the concentrated brine then transferred to crystallization ponds for optional recovery of sodium sulfate salt and the residual bittern evaporated to dryness to recover mixed salts. Compared to other options, this subsystem requires the largest land area for establishment of evaporation and crystallizer ponds.

Following the initial comparison of all options (in terms of capital and operating costs, space requirements, and economic benefits from the by-products), the treatment option comprised of Subsystems A, B, and C (Figure A5.1) was selected as the best choice for further evaluation, as its requirements for land and energy are significantly lower than the other options.

A5.3.2 Conceptual Flow Diagrams

Based on the conceptual block diagrams for Subsystems A, B, and C, simplified process flow diagrams were prepared and material balances calculated for each of the cases. While flow rate and salt amounts differ for each case, the sequence of processing steps is identical to the generic block diagram of Figure A5.2 and is shown only for Case 1 in Figure A5.3. The differences in water quality make-up influence product yield, but the product recovery sequence stays the same.

As briefly explained above, the selection of option ABC as the preferred ZLD process offers a number of advantages, including the following:

- Minimal or no land use for setting up and operating evaporation ponds
- Maximum water recovery by applying feed treatment steps, which lead to by-product recovery
- Recovery of higher-value by-products with established markets in the Southwest and that would entail comparatively lower market entry costs
- Minimal environmental and operational footprints, including reduced brine concentrator and crystallizer pond sizes and reduced landfill disposal needs
- Conversion of dissolved HCO_3 to PCC–$CaCO_3$ solid product, avoiding potential CO_2 release (also true of the ABD and ABE options as well)

The flow diagram in Figure A5.3 shows the amount of chemicals required for treatment and the amount of product produced for a 2.5-MGD flow rate. Higher flows would involve additional parallel equipment modules. It should be noted that once higher throughput volumes are involved, materials handling becomes an important challenge. Furthermore, the layout for larger plants will be significantly different from smaller plants. Note: this study does not address the specifics related to scale-up and material handling.

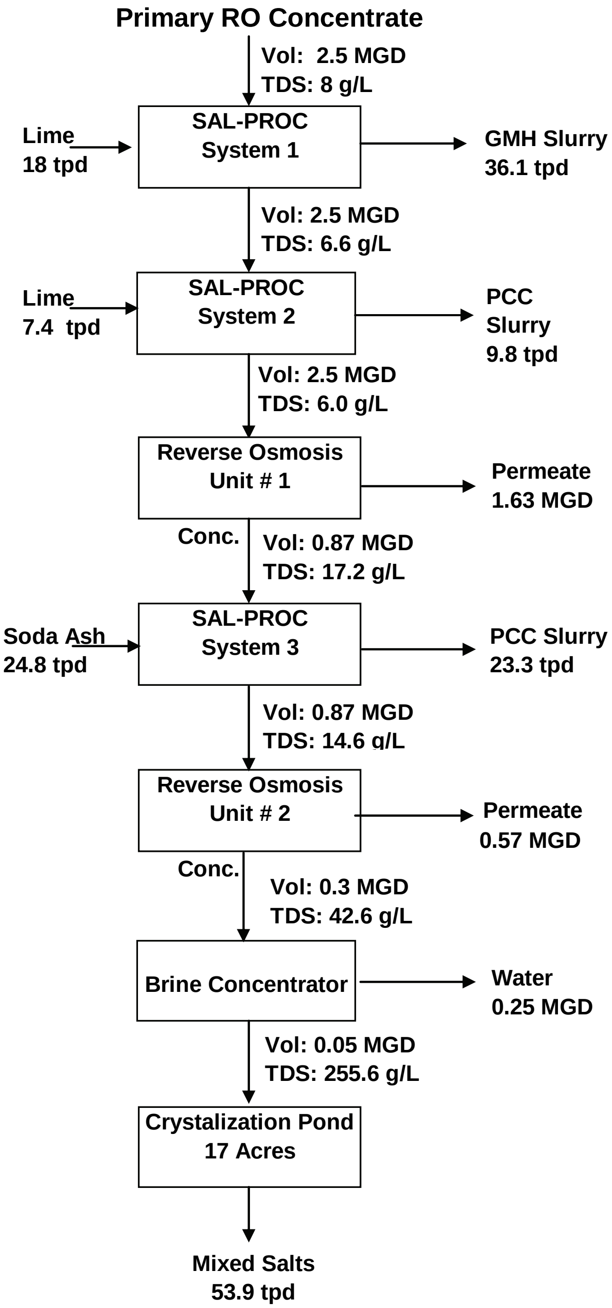


Figure A5.3. Process flow diagram for material balances, Case 1 ABC Subsystems, 2.5 MGD.

A5.4 BY-PRODUCTS AND OVERVIEW OF MARKETS

It is understood that the compositions of the RO concentrates selected for this study generally reflect the qualities of concentrates one can expect from RO treatment of a wide range of brackish groundwater resources in the southwestern United States. According to Geo-Processors' saline water classification scheme, these concentrates belong to water types 1, 2, and 3 and are characterized by elevated concentrations of calcium and sulfate elements at the expense of lower concentrations of sodium and chloride elements. Furthermore, most RO concentrates in this study contain elevated concentrations of dissolved silica; therefore, in the selection of ZLD processes, particular attention was paid to process steps leading to effective reduction of silica and sulfate ions prior to further RO steps for freshwater recovery. As schematically shown in Figure A5.3, the first treatment step in the SAL-PROC^{TM} process involves the recovery of GMH or MH products, which are known to provide effective means for the reduction of silica (by flocculation) or sulfate ions (by precipitation), respectively. Other by-products from SAL-PROC^{TM} include PCC, sodium sulfate, mixed salts, and fresh water. These are briefly described below.

A5.4.1 Water

A key product from the proposed treatment systems is fresh water. Considering the water demand in the region, the revenue potential from freshwater recovery in the proposed treatment systems offers significant economic incentives beyond the efficiency of the proposed processes in achieving ZLD. Revenue from water recovery is, however, not included in the following assessments.

A5.4.2 PCC

PCC is by far the largest revenue source for all cases studied. The largest use for PCC with an average particle size of <1.5 μm, a white color, and distinct brightness is by the paper manufacturing industry. Geo-Processors has extensive experience in recovery of such products. Other uses for PCC are in the manufacture of PVC and plastics, paints, and dyes and as a binder in pharmaceutical products.

A5.4.3 MH and GMH

MH has many applications; it is used widely in the treatment of water and wastewater, for neutralization of acidic effluents, and for the removal of dissolved heavy metals. The major advantage of MH is that it is a pH buffer and when it is used the treated wastewater will not exceed a pH of 10, even if excess MH is added. The sludge that is generated is known to be more compact and to require less dewatering prior to disposal. The addition of lime, caustic soda, or soda ash can raise the pH above 12, causing a potential environmental violation. Other applications are in the manufacture of magnesium metal and production of lightweight concrete and building materials. The SAL-PROC^{TM} process for producing high-grade $Mg(OH)_2$ for use in magnesium metal manufacturing has been licensed to a major car and computer parts manufacturer in Japan.

GMH has both neutralization as well as sodium ion replacement effects (i.e., reductions in the sodium adsorption ratio, or SAR) when applied as a soil conditioner to acidic and sodic soils. Another major application is in the manufacture of lightweight, fireproof building materials.

A5.4.4 Mixed Salts

The mixed salts harvested from crystallization ponds will consist of sodium sulfate, sodium chloride, and potassium chloride in various proportions and include minor impurities, such as silica and probably nitrate. In this report the analysis presumes that the mixed salts will be disposed of off-site in a landfill. However, in view of the significant tonnages produced, it would be advantageous to supply this material to a chemical factory at a reduced price as a feed source for the separation of sodium sulfate as a valuable by-product, using conventional thenardite manufacturing methods.

A5.4.5 Variance in Salts Produced

Table A5.1 shows the mass of solids in tons per day and flows (in MGD) corresponding to the numbered process steps of Figure A5.2 for each of the four cases. Table A.5.2 details the potential annual by-product recoveries and revenue base for the four cases.

The variance in revenue from a low of $3.5M (Case 11) to a high of $5.98M (Case 6) shows the strong dependence of revenue on the water composition.

Table A5.1. Solids (tons/day) and Flows (MGD) for the Four Cases Corresponding to the Numbered Process Steps in Figure A5.2 (2.5 MGD, 8000 mg/L TDS)

Step	Description	Case 1	Case 6	Case 9	Case 11
1	Feedwater	2.5	2.5	2.5	2.5
2	Lime	18.0	22.2	10.0	13.4
3	GMH product	36.1	47.6	7.8	14.5
4	Spent water	2.5	2.5	2.5	2.5
5	Lime	7.4	-----	57.1	23.7
6	PCC product	9.8	-----	38.4	7.9
7	Spent water	2.5	2.5	2.5	2.5
8	RO1 permeate	1.6	1.6	1.3	1.6
9	RO1 reject	0.9	0.9	1.3	0.9
10	Soda ash	24.9	38.1	-----	13.8
11	PCC product	23.3	36.8	-----	20.9
12	Spent water	0.9	0.9	-----	0.9
13	RO2 permeate	0.6	0.6	-----	0.6
14	RO2 reject	0.3	0.3	-----	0.3
15	BC water	0.3	0.3	1.1	0.3
16	BC reject	0.1	0.1	0.1	0.1
17	Mixed salts	53.9	54.5	49.7	74.0

Abbreviation: BC, brine concentrator.

Table A5.2. Potential Annual Product Recoveries and Revenue Base for the Four Cases[1]

Salt and Value	Parameter	Case 1	Case 6	Case 9	Case 11
GMH	Tons/year	12,978	17,132	-----	5216
GMH @ $150/ton	Annual revenue, $M	1.95	2.57	-----	0.78
MH	Tons/year	-----	------	2817	-----
MH @ $300/ton	Annual revenue, $M	-----	------	0.85	-----
PCC	Tons/year	11,916	13,248	13,840	10,393
PCC @ $250/ton	Annual revenue, $M	2.98	3.31	3.46	2.60
Mixed salts	Tons/year	19,413	19,607	17,893	26,649
Mixed salts <$5/ton	Nominal market price, $M	0.10	0.10	0.09	0.13
Total Revenue, $M/year		5.03	5.98	4.40	3.51

[1] For all four cases, the salinity was 8000 mg/L of TDS and the flow rate was 2.5 MGD.

A5.5 PRELIMINARY STREAMS, FLOWS, AND EQUIPMENT DESCRIPTIONS

Case 1 at a flow of 2.5 MGD was chosen as a convenient base to describe the streams, flows, and equipment involved in the Geo-Processors technology. The general nature of the process is substantially the same for other flows. Higher flows may involve additional parallel equipment modules.

Figure A5.3 shows the total mass (in tons per day) and flows for this case in a process flow diagram.

A more detailed look at the equipment arrangement and key stream descriptions are provided in Figure A5.4. The streams and equipment marked in Figure A5.4 are also described in Table A5.3.

NB: Stream numbers in BOLD, equipment numbers in plain text.

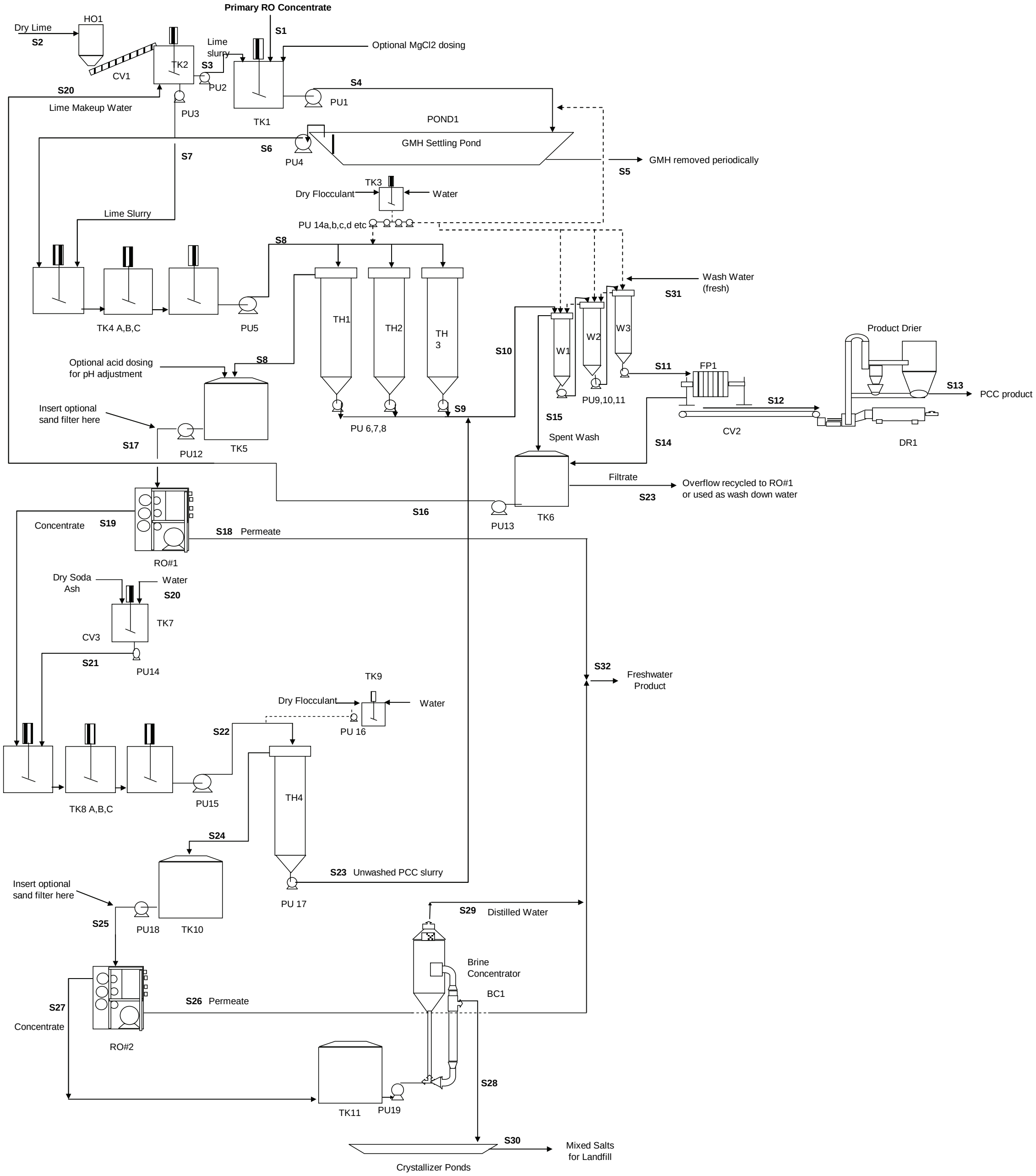


Figure A5.4. Equipment arrangement with key streams for total ZLD process, Case 1 ABC.

Table A5.3. Streams and Equipment Descriptions

Streams and Equipment Descriptions

Note: All Tables to be Read in Conjunction with Figure A5.4

(A) Preliminary Streams

Stream	Description
S1	Process feed water
S2	Dry lime [Ca(OH)2]
S3	Lime slurry to GMH rxn
S4	Process slurry w Mg(OH)2
S5	GMH removed periodically
S6	Clarified solution to PCC rxn
S7	Lime slurry to PCC rxn
S8	Process slurry w PCC
S8A	Clarified solution (spent water)
S9	Thickened PCC slurry
S10	Combined thickened PCC slurry
S11	Washed PCC slurry
S12	Dewatered PCC filter cake
S13	Dry PCC product
S14	Filtrate
S15	Spent wash water
S16	Lime makeup water

Stream	Description
S17	RO1 feedwater
S18	RO1 permeate
S19	RO1 concentrate
S20	Dilution water for soda ash soln
S21	Soda ash solution @ 30% strength
S22	Reactor slurry product
S23	Thickened PCC slurry from Stage D
S24	Spent water, clarified solution
S25	RO2 feedwater
S26	Permeate
S27	RO2 concentrate
S28	Brine concentrator reject
S29	Distilate from BC
S30	Salts may be further refined. Mixture of NaCl, Na2SO4 and some KCl
S31	Wash water
S32	Combined freshwater from all streams

(B) Preliminary Equipment

Ponds

Name	Notes
POND1	GMH settling/pptn pond
POND2	Crystallization pond. May be split into 3-4 subsections. Salts harvested annually

Pumps

Name	Function	Stream
PU1	Process slurry pump	S4
PU2	Lime slurry pump	S3
PU3	Lime slurry pump	S7
PU4	Pond O/F pump	S6
PU5	Reactor soln pump	S8
PU6	Thickener underflow pump	S9
PU7	Thickener underflow pump	S9
PU8	Thickener underflow pump	S9
PU9	Washer underflow pump	S11
PU10	Washer underflow pump	S11
PU11	Washer underflow pump	S11
PU12	RO2 feed pump	S17
PU13	Lime make up water pump	S16
PU14a.b.c	Poly/flocc diaphragm metering pumps	-
PU15	Soda solution pump	S21
PU16	Reactor product pump	S22
PU18	underflow pump	S23
PU17	Polymer pump	-
PU19	RO2 feed pump	S27
PU20	RO2 reject to BC	

Tanks

Name	Function
TK1	Flash mixer/reactor
TK2	Lime makeup vessel
TK3	Flocc/Poly make up tank
TK4 A,B,C	Reactors
TK5	Spent water tank/RO2 feed tank
TK6	Spent wash tank
TK7	Soda dissolution tank
TK8 a.b.c	Reactor vessels
TK9	Flocc/Poly make up tank
TK10	RO2 feed water tank
TK11	BC feed homogenization tank

Thickeners and Washers

Name	Function
TH1	Primary thickener
TH2	Primary thickener
TH3	Primary thickener
TH4	Primary thickener
W1	Washer
W2	Washer
W3	Washer

Other Equipment

Name	Comments
CV1	Screw conveyor to suit (based on continuous lime makeup, not batch)
CV2	Belt conveyor to suit.
CV3	Screw conveyor to suit (based on continuous or batch soda makeup)
FP1	Suggest belt filter press, not plate and frame. Should get 40 -50% solids in cake
DR1	As per manufacturer specs. To handle 50% solids filter cake & produce dry PCC powder
RO#1	Reverse osmosis unit 1
RO#2	Reverse osmosis unit 2
BC3	Brine concentrator

A5.6 CONCEPTUAL CAPITAL COST

Considering the scope of this study, the capital cost estimate presented and further elaborated in the following table is a Class V estimate. A Class V estimate (as defined by the American Association of Cost Engineers) has an accuracy of –30 to +50%.

Table A5.4 shows the effects of the final processing sequence on costs. In the previous analysis, treatment option ABC was used (see Figure A5.1). Table A5.4 provides capital and operating costs for the ABC, ABD, and ABE treatment options for Case 1 at a flow of 2.5 MGD.

Table A5.4. Component-Based Cost and Revenue Estimates for Treatment Options ABC, ABD, and ABE

Process Component	Description	Cost Factor	ABC	ABD	ABE
Evaporation Ponds	TDS salinity, mg/L		8000	8000	8000
	Brine flow rate, MGD		2.50	2.50	2.50
	Area, acres		-----	70.00	70.00
	Capital installed cost, $M	$0.30M/acre	-----	21.00	21.00
	Operating cost, annual $M	1% of capital cost	-----	0.21	0.21
Crystallization Ponds	Area, acres		16.75	16.75	63.50
	Capital installed cost	$0.28M/acre	4.69	4.69	17.78
	Operating cost, annual $M	1.0% of capital cost	0.05	0.05	0.18
Landfill	Mixed salts for land disposal, kton/year		19.41	19.41	9.35
	Area, acres		19.50	19.50	9.50
	Capital installed cost	$0.25 M/Acre	4.88	4.88	2.38
	Operating annual cost, $M	1.0% of capital cost	0.05	0.05	0.02
RO1	Feed flow rate, MGD		2.50	2.50	2.50
	Permeate, MGD		1.63	1.63	1.63
	Permeate recovery, %		65.00	65.00	65.00
	Concentrate, MGD		0.88	0.88	0.88
	Capital installed cost, $M	$2.4M/MGD	6.00	6.00	6.00
	Operating annual cost, $M	$1.00/kgal ($0.36M/year/MGD)	0.90	0.90	0.90
RO2	Feed flow rate, MGD		0.88	0.88	0.88
	Permeate, MGD		0.58	0.58	0.58
	Permeate recovery, %		65.00	65.00	65.00
	Concentrate, MGD		0.30	0.30	0.30
	Capital installed cost, $M	$3.6M/MGD	3.15	3.15	3.15
	Operating annual cost, $M	$1.00/kgal ($0.36M/year/MGD)	0.32	0.32	0.32
Brine Concentrator	Feed flow rate, MGD		0.30	0.08	-----
	Capital installed cost, $M	$7.72 M/MGD	2.32	0.58	-----
	Operating annual cost, $M	$2.43 M/year/MGD	0.73	0.18	-----
By-products Recovery Plant	Plant recovery stages		3.00	3.00	3.00
	Volume to be treated, MGD		2.50	2.50	2.50
	Daily salt load to be handled, tons/year		84.50	84.50	84.50
	Lime reagent usage, tons/year		9135	9135	9135
	Soda ash reagent usage, tons/year		8946	8946	8946
	Capital cost for single-stage plant	$3.1 M /MGD	7.75	7.75	7.75
	Oper. ann. cost, $M single-stage plant	$1.5 M /MGD	3.75	3.75	3.75
	Capital cost, two-stage plant	$5.8 M /MGD	14.50	14.50	14.50
	Oper. ann. cost, $M two-stage plant	$2.2 M /MGD	5.50	5.50	5.50
	Capital cost, three-stage plant	$7.7 M /MGD	19.25	19.25	19.25
	Operating annual cost, $M	$3.2 M /MGD	8.00	8.00	8.00
	FOR THREE STAGE PLANT	TOTAL CAPITAL COST	40.28	59.55	48.56
		TOTAL OPERATING COST	10.04	9.70	9.63

Explanations:
(1) Based on 360 day-year operation
(2) Lower recovery because of elevated Si content

			ABC	ABD	ABE
Potential Annual By-products Recoveries & Revenue Base	GMH, tpa		12978.00	12978.00	12978.00
	$150/ton		1.95	1.95	1.95
	PCC, tpa		11916.00	11916.00	11916.00
	$250/ton		2.98	2.98	2.98
	Sodium Sulfate				11491.25
	$110/ton				1.27
	Mixed salts, tpa		19413.00	19413.00	9353.50
	$5/ton		0.10	0.10	0.05
	Total annual revenue		**5.03**	**5.03**	**6.24**

A5.7 DISCUSSION

Geo-Processors' selective salt recovery technology has been developed, extensively piloted, and licensed outside of the United States, but domestic efforts have been minimal. Although information has been presented at conferences and can be found on Geo-Processors' website (www.geo-processors.com), a more detailed characterization of the technology was warranted. In addition to describing the technology, the potential role this technology might play in municipal membrane concentrate processing was evaluated.

The unique aspect of the technology is that it has the potential to provide both environmental and economic benefits that far outweigh those of conventional treatment methods. By recovering commercial-grade salts from the RO concentrate and either marketing the salts directly or transforming them into value-added products, the salts are no longer wastes and their sale can provide an income stream to offset operating costs, thus improving the treatment economics.

In summary:

- The technology is proven and has been patented, piloted, and licensed. Project-related information is available at the website www.geo-processors.com.
- The general processing approach utilizes conventional processing equipment to implement the proprietary in-depth understanding of the chemical reactions and physical behavior of salts.
- The general processing scheme is tailored to address each site-specific challenge.
- Products of commercial value can be obtained from the concentrates.
- This is the only ZLD option with a revenue stream other than permeate or distillate.
- The nature and composite value of salts from a concentrate depend on the concentrate chemical make-up.
- Comparison of potential salt revenues from the four different 8000-mg/L concentrates (see Table A5.2) show that they range from \$3.5–6.0 M/year for a 2.5-MGD facility.
- The other cases demonstrate the significant revenue potential for waters common to the arid Southwest: waters with low sodium and chloride and high alkalinity.
- The costs presented for the Geo-Processors' technology relied on information from the technology developer and marketer. The study author attempted to assign the same cost basis to RO, brine concentrator, evaporation ponds, and other equipment items but could not do this for the SAL-PROC^{TM} equipment. Consequently, the author cannot apply the same confidence in the cost estimates. That being said, the costs developed in this chapter are likely within the Class V estimate of –30 to +50%.
- The more detailed consideration of Case 1 provides some insight into the nature of the processing and revenue costs. In rough figures, the projected capital cost for a 2.5-MGD treatment process is \$40M, the operating cost is \$10M/year, and the potential revenue is \$5M/year.
- Comparison of these costs with those of Chapter 5 suggests the selective salt recovery treatment is at least cost-competitive if not more attractive than the non-salt recovery treatments.
- The different final processing options and costs are illustrated in Table A5.4 for a 2.5-MGD system.
- This potential attractive economic situation perhaps obscures the important benefit that environmental impacts are being lessened, including reducing the potential CO_2

footprint of the desalination plants through removal of dissolved CO_2 via product formation.

- The Geo-Processors technology represents an important step toward the goals of sustainability and environmentally sound water resource management.

Advancing the Science of Water Reuse and Desalination

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