Use of Produced Water in Recirculated Cooling Systems at Power Generating Facilities

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ABSTRACT

Tree ring studies indicate that, for the greater part of the last three decades, New Mexico has been relatively "wet" compared to the long-term historical norm. However, during the last several years, New Mexico has experienced a severe drought. Some researchers are predicting a return of very dry weather over the next 30 to 40 years. Concern over the drought has spurred interest in evaluating the use of otherwise unusable saline waters to supplement current fresh water supplies for power plant operation and cooling and other uses.

The U.S. Department of Energy's National Energy Technology Laboratory sponsored three related assessments of water supplies in the San Juan Basin area of the four-corner intersection of Utah, Colorado, Arizona, and New Mexico. These were (1) an assessment of using water produced with oil and gas as a supplemental supply for the San Juan Generating Station (SJGS); (2) a field evaluation of the wet-surface air cooling (WSAC) system at SJGS; and (3) the development of a ZeroNet systems analysis module and an application of the Watershed Risk Management Framework (WARMF) to evaluate a range of water shortage management plans.

The study of the possible use of produced water at SJGS showed that produce water must be treated to justify its use in any reasonable quantity at SJGS. The study identified produced water volume and quality, the infrastructure needed to deliver it to SJGS, treatment requirements, and delivery and treatment economics. A number of produced water treatment alternatives that use off-the-shelf technology were evaluated along with the equipment needed for water treatment at SJGS.

Wet surface air-cooling (WSAC) technology was tested at the San Juan Generating Station (SJGS) to determine its capacity to cool power plant circulating water using degraded water. WSAC is a commercial cooling technology and has been used for many years to cool and/or condense process fluids. The purpose of the pilot test was to determine if WSAC technology could cool process water at cycles of concentration considered highly scale forming for mechanical draft cooling towers. At the completion of testing, there was no visible scale on the heat transfer surfaces and cooling was sustained throughout the test period.

The application of the WARMF decision framework to the San Juan Basis showed that drought and increased temperature impact water availability for all sectors (agriculture, energy, municipal, industry) and lead to critical shortages. WARMF-ZeroNet, as part of the integrated ZeroNet decision support system, offers stakeholders an integrated approach to long-term water management that balances competing needs of existing water users and economic growth under the constraints of limited supply and potential climate change.

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REPORT SUMMARY

Due to severe drought conditions in the Southwest in recent years, EPRI and the U.S. Department of Energy's National Energy Technology Laboratory have sponsored three related assessments of water supplies in the San Juan Basin area of the four-corner intersection of Utah, Colorado, Arizona, and New Mexico. Two of the studies assess the use of saline waters in power plants. The third describes the adaptation of a deterministic watershed model to forecast the impact of climate change on river hydrology in the San Juan Basin.

Background

Tree ring studies conducted by climate researchers at the University of Arizona have shown that the last thirty years in New Mexico have been relatively wet as compared to the norm. Historically, wet-dry-wet cycles have occurred every 60 to 80 years; and the recent wet period in New Mexico was coincident with economic development—expansion of agriculture, extensive oil and gas production, and the construction and operation of two large coal-fired power plants in the Four Corners area. However, New Mexico has been suffering from a severe drought in recent years, and researchers are predicting a return of very dry weather over the next 30 to 40 years. Concern over the drought has spurred interest in better understanding the effects of climate change on regional water budgets and in evaluating the use of otherwise unusable saline waters to supplement existing limited supplies of fresh water in the region to supplement water supply for power plant operation and cooling and other uses.

Objectives

- To assess the use of produced water, a byproduct of oil and gas production, as a supplemental supply for the San Juan Generating Station (SJGS)
- To conduct a field evaluation of the use of degraded water in the wet-surface air cooling (WSAC) system at SJGS
- To develop a ZeroNet systems analysis module and application of the watershed analysis risk management framework (WARMF) decision support framework to evaluate alternative management plans for water sharing under drought conditions.

Approach

In the assessment of the use of produced water, the project team addressed the volume and quality of available produced water, the infrastructure that would be needed to deliver it to SJGS and to treat it for plant use, and the economics of delivery and treatment. In the field evaluation of WSAC, the team field-tested the technology at SJGS to determine its capacity to cool power plant circulating water using degraded water. In the development of a framework to evaluate alternative drought management plans, the project team adapted a deterministic watershed model to forecast the impact of climate change on river hydrology in the San Juan Basin.

Results

The study of the possible use of produced water at SJGS showed that produced water must be treated to justify its use in any reasonable quantity at SJGS. The study identified produced water volume and quality, the infrastructure needed to deliver it to SJGS, treatment requirements, and delivery and treatment economics. The study evaluated a number of produced water treatment alternatives that use off-the-shelf technology as well as the equipment needed for water treatment at SJGS.

Pilot testing of WSAC technology showed that degraded water could cool process water at impurity concentrations considered highly scale forming for mechanical draft cooling towers. At the completion of testing, there was no visible scale on the heat transfer surfaces and cooling was sustained throughout the 147-day test period.

The application of the WARMF decision framework to the San Juan Basis showed that drought and increased temperature impact water availability for all sectors (agriculture, energy, municipal, industry) and lead to critical shortages. The WARMF application extended the analysis beyond its traditional deterministic roots by implementing a Monte Carlo sampling technique to construct climate scenarios that account for the uncertainty of wet, normal, and dry years. WARMF-ZeroNet, as part of the integrated ZeroNet decision support system, offers stakeholders an integrated approach to long-term water management that balances competing needs of existing water users and economic growth under the constraints of limited supply and potential climate change.

EPRI Perspective

EPRI developed the WARMF to help watershed and government agencies prepare for significant changes in water resource management policy. WARMF calculates total maximum daily load (TMDL) analyses and thus facilitates allocation of TMDLs among stakeholders. More information about WARMF can be found in EPRI Report TR-110709.

Keywords

Drought
Produced water
Wet surface air cooling
WARMF (Watershed Analysis Risk Management Framework)
Climate change
Power plant cooling
Watersheds

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1
INTRODUCTION

1.1 Introduction

New Mexico has been suffering from a severe drought in recent years. Tree ring studies conducted by climate researchers at the University of Arizona[1] have shown that the last thirty years in New Mexico have been relatively "wet" as compared to the norm. Historically, wet-dry-wet cycles have occurred every 60 to 80 years. The current wet period in New Mexico is coincident with economic development – expansion of agriculture, extensive oil and gas production and the construction and operation of two large coal-fired power plants in the Four Corners area. Researchers are predicting a return of very dry weather over the next 30 to 40 years. Concern over the drought has spurred interest in evaluating the use of otherwise unusable saline waters to supplement existing limited supplies of fresh water in the region to supplement water supply for power plant operation and cooling and other uses.

The U.S. Department of Energy's National Energy Technology Laboratory sponsored three related assessments of water supplies in the San Juan Basin area of the four-corner intersection of Utah, Colorado, Arizona, and New Mexico. They included (1) an assessment of using water produced with oil and gas as a supplemental supply for the San Juan Generating Station (SJGS); (2) a field evaluation of the wet-surface air cooling (WSAC) system at SJGS; and (3) development of a ZeroNet module and application of the WARMF decision framework to evaluate the impact of climate change on water supplies in the San Juan Basin.

SJGS is operated by Public Service of New Mexico (PNM) and is located about 15 miles northwest of Farmington, New Mexico. It has four units with a total generating capacity of about 1,800 MW. The coal-fired plant uses 22,400 acre-feet of water per year from the San Juan River, most of its demand coming from cooling tower make-up. SJGS is a zero liquid discharge facility and, as such, it is well practiced in efficient water use and reuse.

1.2 Objective and Scope

The objectives were to (1) evaluate the use of produced water from oil and gas wells as a supplemental source of water for the San Juan Generating Station (SJGS); (2) assess the use of cooling tower blow down from SJGS as makeup water in the West Surface Air Cooling System; and (3) develop and apply a ZeroNet module to assess produced water volume and quality,

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[1] F. Ni, T. Cavazos, M. K. Hughes, A. C. Comrie, and G Funkhouser, "Cool-Season Precipitation in the Southwestern USA Since AD 1000: Comparison of Linear and Nonlinear Techniques for Reconstruction, *International Journal of Climatology,* Volume 22, Issue 13, pp. 1645 - 1662, November 15, 2002.

infrastructure to deliver it to SJGS, treatment requirements to use it at the plant, delivery and treatment economics, etc.

1.3 Organization of this Report

Following Section 1, Introduction and Executive Summary, Sections 2 through 9 present the produced water project assessment, analyses, and results. Section 2 provides background information on produced water production with oil and gas in the San Juan Basin and the San Juan Generating Station. Section 3 addresses produced water infrastructure, availability, and transportation resources. Section 4 assesses produced water use, treatment and disposal. Section 5 evaluates a ceramic membrane system to reduce dissolved solids and chloride content of produced water. Section 6 assesses the compatibility of treated produced water as a supplement to or replacement of fresh water at the San Juan Generating Station. Section 7 performs a cost-benefit analysis of gathering, conveying, and treating produced water for use at San Juan Generating Station. Section 8 describes a two phase implementation plan for use of produced water at SJGS. Section 9 addresses applicability of the produced water analysis to other regions in the U.S.

Section 10 describes the Wet Surface Air Cooling test at San Juan Generating station, which assessed the capability to cool power plant circulating water using degraded water from the blowdown of SJGS unit 3.

Section 11 describes adaptation of a deterministic watershed model to forecast the impact of climate change on river hydrology in the San Juan Basin.

1.4 Executive Summary

1.4.1 Produced Water Assessment

Produced water is generated nationally as a byproduct of oil and gas production. Seven states generate 90 percent of the produced water in the continental US. About 37 percent of the sources documented in the US Geological Survey's Produced Waters Database were deemed to be treatable.

The produced-water assessment addressed produced water volume and quality, infrastructure to deliver it to SJGS, treatment requirements to use it at the plant, and delivery and treatment economics.

A baseline of produced water generation was developed at the outset to establish the study area. Oil and gas production, produced water handling and disposal, and produced water quantities and chemistry are introduced. Legislative efforts to enable the use of this water at SJGS are also described. As shown in Table 1-1, the estimated total water resource in the study area is about 55,000 barrels per day (BPD).

The largest obstacle to produced water reuse in the San Juan Basin is the lack of pipeline infrastructure for its transport. Pipeline infrastructure is almost exclusively used for oil, gas and product transport. Most of the produced water in the Basin is stored in tanks at the well head and must be transported by truck to salt water disposal (SWD) facilities prior to injection. Produced water transportation requirements from the well head to SJGS and the availability of existing infrastructure to transport the water are discussed.

Table 1-1
Total Water Resources in San Juan Basin – 2006

	BPD	gpm	AF/yr
Fairway	22,600	659	1060
Tri-City	3,020	88	142
Close-in	13,680	399	644
Backflow	10,000	292	470
Total Produced Water	49,300	1,438	2,316
Prax Air – Cooling Tower Blowdown	300	9	14
BHP Billiton – Mine Water	1,700	50	80
Purge Water – SO_2 Absorber Bleed Stream	3,430	100	161
Total Other Water	5,430	159	255
Total Water Resource	54,730	1,597	2,571

A simplified water balance was developed and water quality constraints were established for each major water user. It is shown that the produced water must be treated to justify using it in any reasonable quantity at SJGS. A number of produced water treatment alternatives are evaluated utilizing off-the-shelf technology. Water treating equipment at SJGS is also incorporated into the evaluation.

In conjunction with this study, bench-scale testing was conducted by CeraMem Corporation to evaluate ceramic membrane filtration. The process could be used to reduce the level of certain forms of contamination in produced water, i.e. oil and particulate matter. A benefit of this technology is that ceramic membranes could last for a significant period of time, thereby reducing the operating cost of pretreatment.

The compatibility of treated produced water is also assessed. Potential water users at SJGS are assessed for flow capacity and chemistry, i.e. constituents of concern and corrosion and deposition potential. Costs associated with the use of treated produced water in each area are estimated and summarized.

The costs of gathering, conveying and treating produced water for use at SJGS are evaluated. Life-of-project projections are developed for the produced water resource in the Study Area and a number of scenarios are assessed to determine reasonable recoverable volumes of water. PNM and producer revenue sharing, in the form of reduced produced water disposal costs, is also incorporated into the economic analysis.

The development of the produced water project was evaluated in two phases. The first phase would consist of a 10.8-mile pipeline to convey water from close-in producers to a new water treating facility located at SJGS. In Phase 2, a collection center would be built and the pipeline would be extended to its full length – about 28.5 miles. A major producer in New Mexico would install satellite collection stations using existing/unutilized pipeline infrastructure.

A methodology was developed to readily estimate capital and operating costs for produced water treatment. Two examples are presented to show how the methodology can be used to evaluate the cost of treatment of produced water at power plants close to oil and gas production. As shown in Table 1-2, the estimated capital cost for gathering, transporting and treating the produced water at the San Juan Generating Station is about $43 million.

Table 1-2
Total Produced Water Project Capital Costs

Total Project Capital Costs		
BR	Gathering system to Collection Center	$5,000,000
Dugan	Inject into pipeline	$100,000
Richardson	Inject into pipeline	$100,000
PNM	Collection Center, pipeline & treatment	$37,900,000
Total Project		$43,100,000

Lastly, possible test configurations for produced water demonstration projects at SJGS are described. The ability to host demonstration projects would enable the testing and advancement of promising produced water treatment technologies. Testing is described for two scenarios – with and without a produced water treatment system at SJGS.

1.4.2 Wet-Surface Air Cooling Field Assessment at SJGS

Wet surface air cooling (WSAC) technology was tested at the San Juan Generating Station (SJGS) to determine its capacity to cool power plant circulating water using degraded water. WSAC is a commercial cooling technology and has been used for many years to cool and/or condense process fluids.

Figure 1-1 is a schematic of the WSAC system. In the WSAC, water is applied in dense spray patterns to the exterior of tubes carrying the liquid or gas to be cooled.

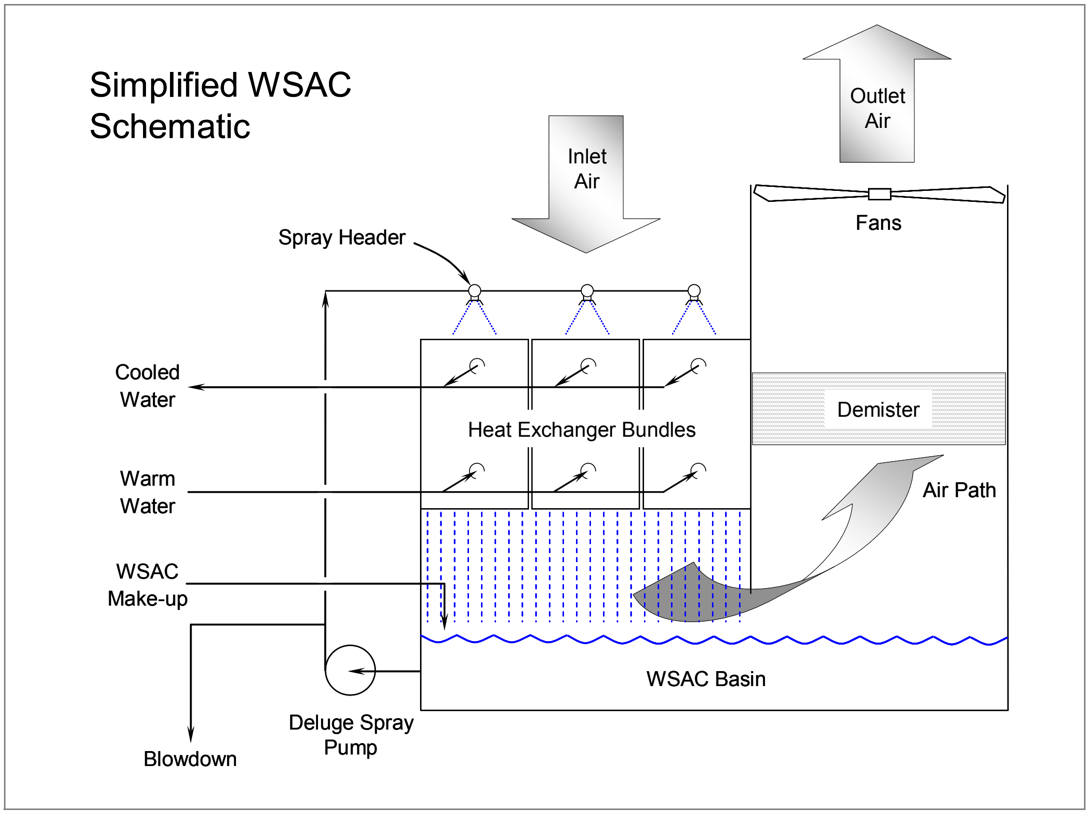

Figure 1-1
Schematic of the Wet-Surface Air Cooling Pilot Unit (WSAC)

At the same time, air is also drawn down and around the perimeter of the tubes in the same direction as spray water.

The purpose of the pilot test was to determine if WSAC technology could cool process water at cycles of concentration considered excessive (highly scale forming) for mechanical draft cooling towers.

The WSAC pilot was online for a total of 147 days – from July 5 to November 29, 2005, during which, it was in service for 2,898 hours. It was configured to cool circulating water from Unit 3, and at the same time, use Unit 3 circulating water for make-up. In this mode, the WSAC operated at an equivalent of 24 to 70 cycles of concentration (based on freshwater fed to SJGS cooling towers). Ten cycles of concentration is considered a safe limit at this plant to control mineral scale formation. An additional benefit of operating in this mode is that WSAC could also function as a concentrating device by reducing the volume of a large plant wastewater stream. At the completion of testing, there was no visible scale on the heat transfer surfaces (tube externals) and cooling was sustained throughout the test period.

1.4.3 WARMF Decision Support Framework Application to the San Juan Basin

Well-designed and useful decision support systems (DSS) are critical tools for managing water resources. As part of the U.S. Department of Energy's ZeroNet Water-Energy Initiative, the ZeroNet DSS was developed to evaluate effects of climate change on water budgets. The ZeroNet DSS incorporates the existing Watershed Analysis Risk Management Framework (WARMF), which was enhanced through development of a ZeroNet Module. This tool allows stakeholders to evaluate water management scenarios under different climate conditions, and visualize flow conditions and water shortages and surpluses within a basin.

The WARMF application to 16,000 mi^2 (42,000 km^2) of the San Juan Basin (CO, NM) extended the analysis beyond its traditional deterministic roots by implementing a Monte Carlo sampling technique to construct climate scenarios that account for the uncertainty of wet, normal, and dry years. The application estimated the mean and standard deviation of the Navajo Reservoir level in the San Juan Basin vs. time for each of six scenarios representing each combination of three- and five-year drought conditions and mean ambient temperature increases of 0, 1, and 2 deg C.

Figure 1-2 illustrates the cumulative probability distribution of the surface elevation of the Navajo Reservoir at the end of year five for one of the seven-year scenarios, which assumed a three-year drought and a one-degree C surface temperature increase (Scenario d3t1). To maintain the reservoir level above the 1825 meter level established for the Navajo Reservoir, it was necessary to reduce the prescribed reservoir releases by 65% during specific drought years. The probability distribution covers the range from about 1770 to 1855 meters and the 1825-meters limit corresponds to about 40% cumulative probability, e.g. the reservoir level would be below the minimum 40% of the time under this scenario. However the mean reservoir level remains above the minimum 1825-meter level during entire seven-year period as shown in Figure 1-3.

Table 1-3 summarizes the results of all six scenarios. In each scenario, it was necessary to limit reservoir releases during at least one year of the drought period and the limits ranged from 18% for scenario d3t0 (3-year drought, zero temperature increase) and 70% for scenarios d3t2 and d5t2 (three- and five-year droughts, 2 deg C temperature increase). The table also presents the water volume held back in the reservoir and the total and average downstream water shortage for each scenario.

The application of the WARMF decision framework to the San Juan Basis demonstrates its application for water supply management. The scenario simulations showed that drought and increased temperature impact water availability for all sectors (agriculture, energy, municipal, industry) and lead to critical shortages. WARMF-ZeroNet, as part of the integrated ZeroNet DSS, offers stakeholders an integrated approach to long-term water management to balance competing needs of existing water users and economic growth under constraints of limited supply and potential climate change.

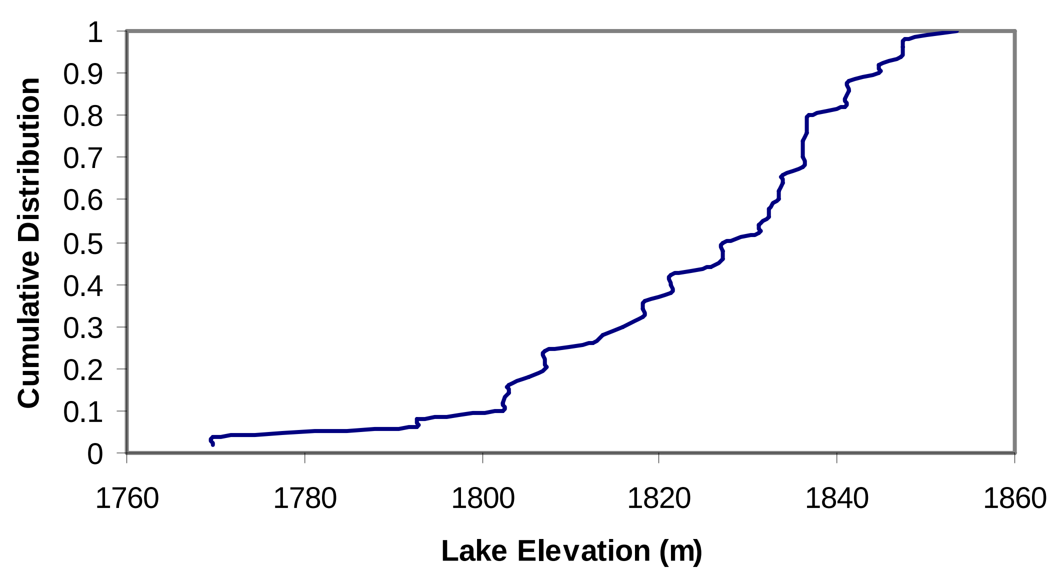

Figure 1-2
Projected cumulative distribution of surface elevation of Navajo Reservoir at the end of year 5 for a 3-year drought and 1-degree temperature increase (Scenario D3T1)

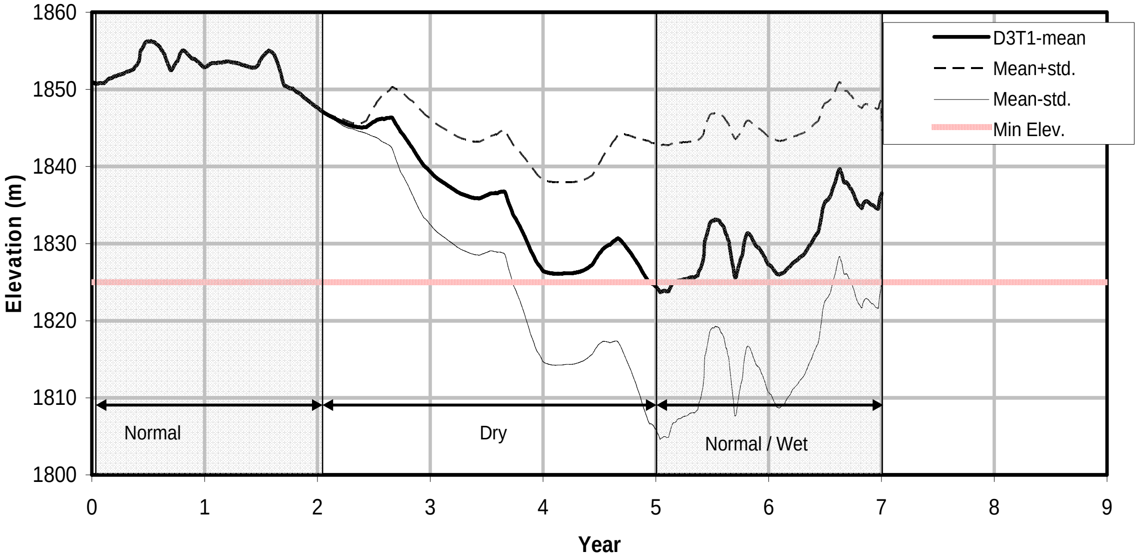

Figure 1-3
Projected mean surface elevation of Navajo Reservoir under a 3-year drought and 1-degree temperature increase (Scenario D3T1) with ±1 standard deviation

Table 1-3
Projected Water Shortages for Drought / Climate Scenarios

Scenario	Reservoir Release Adjustment	Volume Held Back in Reservoir (AF)	Number of Years with Reservoir Release Adjustment	Total Shortage Downstream of Reservoir During Drought Period (AF)	Average Shortage Downstream of Reservoir (AF/yr)[1]
D3T0	18%	86,786	1	266	266
D3T1	65%	314,010	1	45,050	45,050
D3T2	70%	689,548	2	129,173	64,586
D5T0	45%	656,699	3	27,767	9,256
D5T1	62%	896,268	3	109,713	36,571
D5T2	70%	1,358,127	4	248,282	62,071

[1] Averaged for years when reservoir release

2

ASSESSMENT OF PRODUCED WATER FROM OIL AND GAS WELLS

2.1 Introduction

The objective of the produced water assessment was to evaluate produced water from oil and gas wells as a supplemental source of water for the San Juan Generating Station (SJGS). The study addressed produced water volume and quality, infrastructure to deliver it to SJGS, treatment requirements to use it at the plant, delivery and treatment economics, etc. This section addresses produced water points of generation, quantity and quality.

SJGS is operated by Public Service of New Mexico (PNM) and is located about 15 miles west of Farmington, New Mexico. It consists of four units with a total generating capacity of about 1,800 MW. The plant uses 22,400 acre-feet of water per year from the San Juan River, of which most is used for cooling tower make-up. The plant is a zero liquid discharge facility and, as such, is well practiced in efficient water use and reuse.

For the past few years, New Mexico has been suffering from a severe drought. Tree ring studies conducted by climate researchers at the University of Arizona[1] have shown that the last thirty years in New Mexico have been relatively "wet" compared to the norm. Historically, wet-dry-wet cycles have occurred every 60 to 80 years. The current wet period in New Mexico is coincident with economic development – expansion of agriculture, extensive oil and gas production and the construction and operation of two large coal-fired power plants in the Four Corners area. Researchers are predicting a return of very dry weather over the next 30 to 40 years. Concern over an impending drought has spurred interest in evaluating the use of otherwise unusable saline waters.

2.2 San Juan Basin

The San Juan Basin (the Basin) is designated as Geologic Province 22 by US Geological Survey (USGS) and is rich in oil, gas and coal as well as minerals. New Mexico ranks 4th in natural gas and 7th in crude oil production in the nation. The Basin is located in the northwest corner of New Mexico with a small portion in southwest Colorado. Refer to Figure 2-1. At its greatest dimensions, the Basin is 130 miles by 160 miles and is comprised of a number of producing

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[1] F. Ni, T. Cavazos, M. K. Hughes, A. C. Comrie, and G. Funkhouser, "Cool-Season Precipitation in the Southwestern USA Since AD 1000: Comparison of Linear and Nonlinear Techniques for Reconstruction, *International Journal of Climatology,* Volume 22, Issue 13, pp. 1645 - 1662, November 15, 2002.

geologic units. The Fruitland Petroleum System (the Fruitland) generates the produced water assessed in this study. SJGS is situated on the western edge of the Fruitland, which is the coal source for the plant.

As oil or gas is produced, the fluid brought to the surface typically contains oil and water, gas and water or all three components. In oil production for example, it is not unusual to obtain nine barrels of water for every barrel of oil. Produced water salinity is quite variable and depends upon the hydrologic conditions of the producing zone, e.g. saline native waters from an ancient seabed or a hydrologic connection to a freshwater aquifer. In the San Juan Basin, produced water salinity measured as total dissolved solids (TDS) can vary from 100 mg/l to 60,000 mg/l.

There are two types of oil and gas reserves in the Basin:

- Conventional/continuous oil and gas
- Coal bed methane (CBM)

In conventional and continuous production, a well is drilled into a formation and oil and/or gas are extracted. Conventional formations are well defined from a geologic perspective with clear-cut reserve boundaries. Continuous formations, in contrast, have poorly-delineated boundaries and generally defined reserves. In coal bed methane (CBM) production, methane gas is extracted directly from coal seams. Conventional and continuous wells can range from 3,500 to 8,000 feet in depth in the Fruitland. CBM wells are usually shallow – 1,000 to 3,000 feet – and typically produce a significant amount of water.

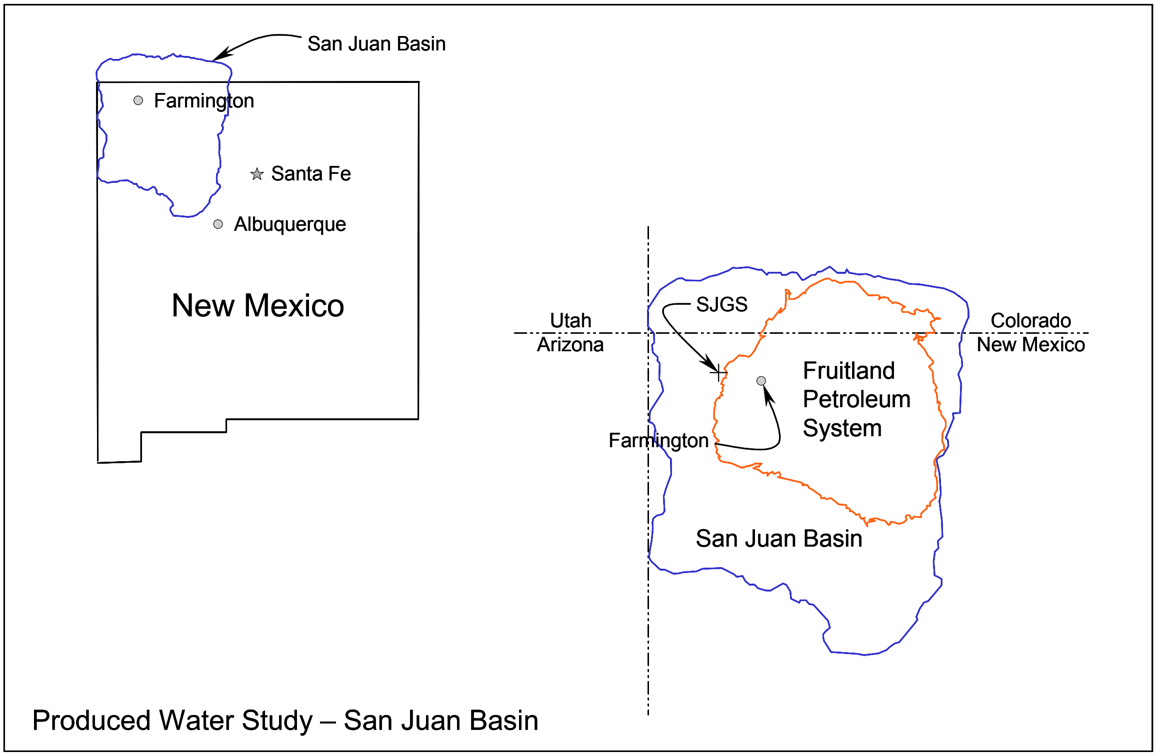

Figure 2-1
Region of Produced Water Study in New Mexico

2.3 Regulatory Framework

The Oil Conservation Division[2] (OCD) regulates all oil and gas production in the state. In New Mexico (as in many other states), produced water is designated a waste byproduct of oil and gas production. Shortly after produced water is brought to the surface, it is de-oiled, filtered and disposed of via injection wells. There are several underlying formations in the Basin that are routinely used for produced water injection, e.g. the Mesa Verde, Dakota and Entrada. Injection wells range from 2,000 to 8,000 feet deep and operate at fairly high injection pressures – from 1,000 to 2,500 psi. Production and injection zones are described as "tight" formations in the San Juan Basin and require fracturing to break or crack formation rock to provide flow paths for production fluids.

There have been several attempts to make use of produced water (e.g. for dust suppression or road construction) rather than dispose of it via injection. In New Mexico this action is defined as a beneficial use of the state waters and is regulated by the Office of the State Engineer (OSE). Under this designation, a right to use the water must be obtained and its use must comply with all applicable environmental regulations. Also, it must be demonstrated that the produced water being considered has no hydrologic connection to other waters of the state, i.e. rightful water assigned to others has not been appropriated. The regulatory and environmental protection afforded by the OCD (designating the water as a byproduct of oil and gas production) would be lost with beneficial use. It is for this reason that producers would prefer to inject the water rather than use it for another purpose.

2.4 Legislative Remedies

PNM endeavored to address this regulatory issue involving produced water reuse by supporting a bill in the New Mexico legislature in January of 2004 that would specifically allow the "disposal" of produced water at electric generating facilities. This would designate produced water reuse as an alternate method of disposal. As a result, a beneficial use would not be created and the regulatory jurisdiction of the OCD would not be invoked. The bill attempted to accomplish two goals:

- Allow producers to dispose of produced water at SJGS. The plant could treat and utilize the water for cooling tower make-up, scrubber make-up, ash wetting, etc. Most of the water would be consumed as evaporative losses or waters of moisture in scrubber sludge or ash. Any residual produced water (wastes from treatment) would be disposed of in the permitted and regulated evaporation ponds at SJGS.
- PNM would receive a tax credit to compensate for the cost of conveying and treating the water that would otherwise be too costly to consider as economically viable.[3] The amount of the proposed tax credit was $1,000 per acre-foot of produced water delivered to SJGS not to

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[2] OCD is a division of the Energy, Minerals and Natural Resources Department of the state of New Mexico.

[3] This is similar to an existing tax credit for treating and discharging produced water into the Rio Grande River in southeast New Mexico to meet interstate water compacts with Texas.

exceed $3 million annually. Also, there would be a maximum payable life-of-the-project cap equal to 50 percent of the capital cost of the project.

The bill was introduced into the January-February 2004 state legislative session and the provision allowing produced water disposal at electric generation facilities such as SJGS was signed into law in March 2004. The tax credit was not included in the bill and, if it is to be pursued, it will have to be reintroduced in an upcoming legislative session.

2.5 Produced Water Quantity

There are 19,090 oil and gas wells (categorized as active wells by OCD in 2003) in the San Juan Basin and they generate approximately 68,500 BPD (barrels per day) of produced water in an area covering about 3,200 square miles. Refer to Figure 2-2 for a map of the "Study Area". The Study Area[4], which covers about 2,400 square miles, was selected based on its proximity to:

- High-volume areas of produced water generation in the Basin
- Existing east-west gas transmission lines and their associated rights of way.

The gas transmission lines generally bisect the Study Area and run parallel to state Highway 64. In Kirtland, the lines branch off in different directions westward – some head in a northwest direction just past SJGS.

Refer to Figure 2-2 for a map of the extent of oil and gas production. The township grids are included in Study Area map because they delineate the areas of production activity that OCD[5] uses to locate oil, gas, CBM and injection wells.

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[4] The Study Area was reduced to 1,500 square miles later in the report to focus on specific high-volume areas of produced water generation.

[5] Raw data was provided by OCD and can be found at their website, emnrd.state.nm.us/OCD/. Producers must report oil and gas production as well as produced water generation and disposal to OCD.

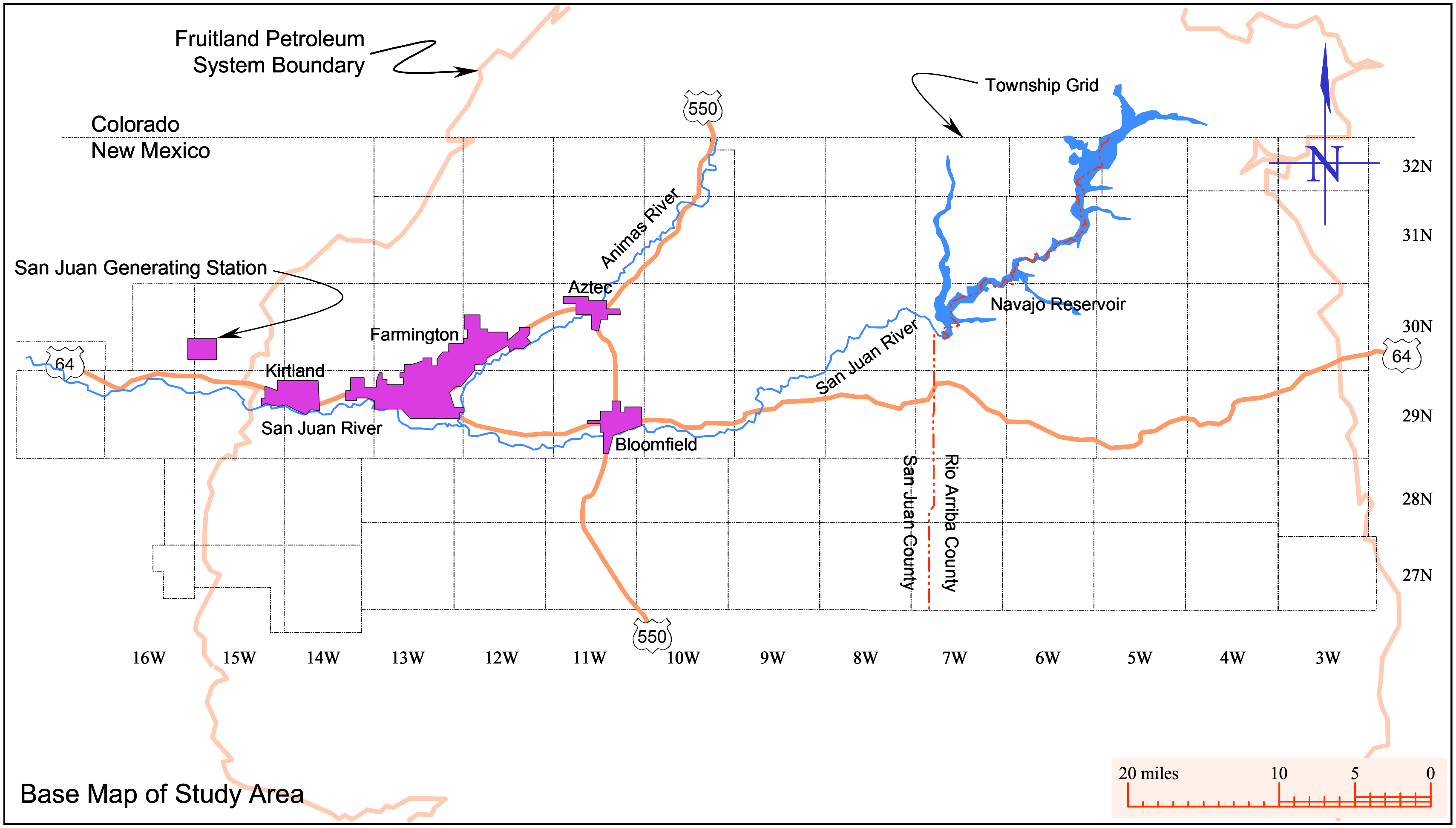

Figure 2-2
Base Map of Produced Water Study Area

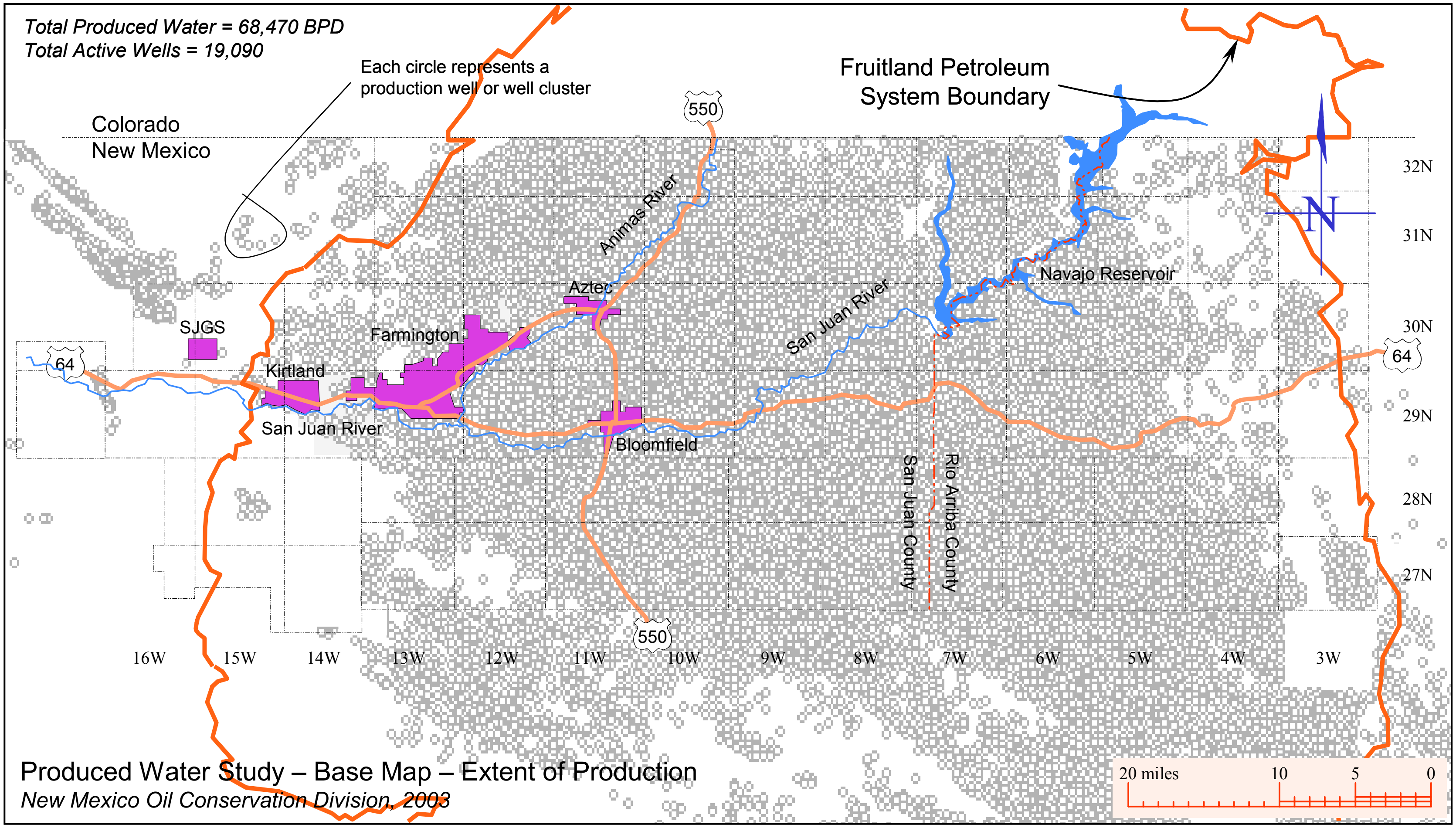

Figure 2-3
Locations of Producing Oil and Gas Wells in Produced Water Study

The wells are generally located in low-density patterns, i.e. one well every 160 to 320 acres, with little interconnecting piping and infrastructure to gather produced water. Well density will increase in New Mexico with the recent approval from the Bureau of Land Management (BLM) allowing for production infilling, i.e. one well every 80 acres on federal lands. While some producers have installed water gathering lines, most wells are not connected to any type of collection system.

The Study Area was established to identify produced water that is proximate to SJGS as well as existing infrastructure that could be used to convey the water, e.g. underutilized or abandoned pipelines. The Study Area generated 53,900 BPD (average daily) of produced water in 2003 – about 79 percent of all the water produced in the San Juan Basin in New Mexico.

Produced water gathering strategies are discussed in detail in Section 3, Produced-Water Infrastructure Availability and Transportation Analysis. Produced water generation patterns for the Study Area (based on 2003 data) are summarized below:

- 48,100 BPD or 89 percent of the produced water in the Study Area was generated north of township grid 28N. Refer to Table 2-1. Figure 2-4 highlights in blue the townships that generate more than 500 BPD of produced water in the Study Area. These townships generated approximately 46,223 BPD or 86 percent of the produced water in the Study Area. With the exception of four townships, all are located at or north of Highway 64.
- Refer to Table 2-2 for a sensitivity analysis of produced water generation in the Study Area versus township volume. Of the 78 townships in the Study Area, 44 townships did not generate more than 300 BPD. One township generated less than 1 BPD and seven had no production. As the production-per-township quantity is increased, the number of townships starts to drop dramatically.
- The two largest clusters in the Study Area generated 39,200 BPD (largest highlighted areas).
- Two of the townships on the western edge of the Fruitland (CBM production) – 29N14W and 30N14W – generated the most produced water of any of the townships in the Study Area, 12,516 BPD.

Table 2-1
Produced Water Generation by Township Grid

Township Grid (Range 3W-16W)	Produced Water BPD	Produced Water Pct of Total	Produced Water Cum Pct
32N	8,475	15.7%	15.7%
31N	14,051	26.0%	41.8%
30N	16,651	30.9%	72.6%
29N	8,896	16.5%	89.1%
28N	2,605	4.8%	93.9%
27N	3,269	6.1%	100.0%

Table 2-2
Produced Water Generation vs. Township Volume

Townships with Volumes Greater than	Number of Townships	Total Produced Water Generation	Produced Water Pct of Total
<1 BPD	8	<1 BPD	0.0%
1 – 299 BPD	36	5,022 BPD	9.3%
300 – 699 BPD	14	6,680 BPD	12.4%
700 – 1,499 BPD	11	10,619 BPD	19.7%
1,500 – 2,999 BPD	6	14,504 BPD	26.9%
3,000+ BPD	3	17,122 BPD	31.7%

Three of the high-volume townships are split by the San Juan River and four townships are south of it. Transporting produced water via pipeline from south of Highway 64 is complicated by the fact that the San Juan River flows parallel to the highway in the Study Area, which would necessitate a river crossing. Also, produced water south of Highway 64 comprises a small fraction of available water in the Study Area and is generally more saline than average (discussed later). Lastly, refer to Figure 2-4 for a summary of produced water generation by township in the Study Area.

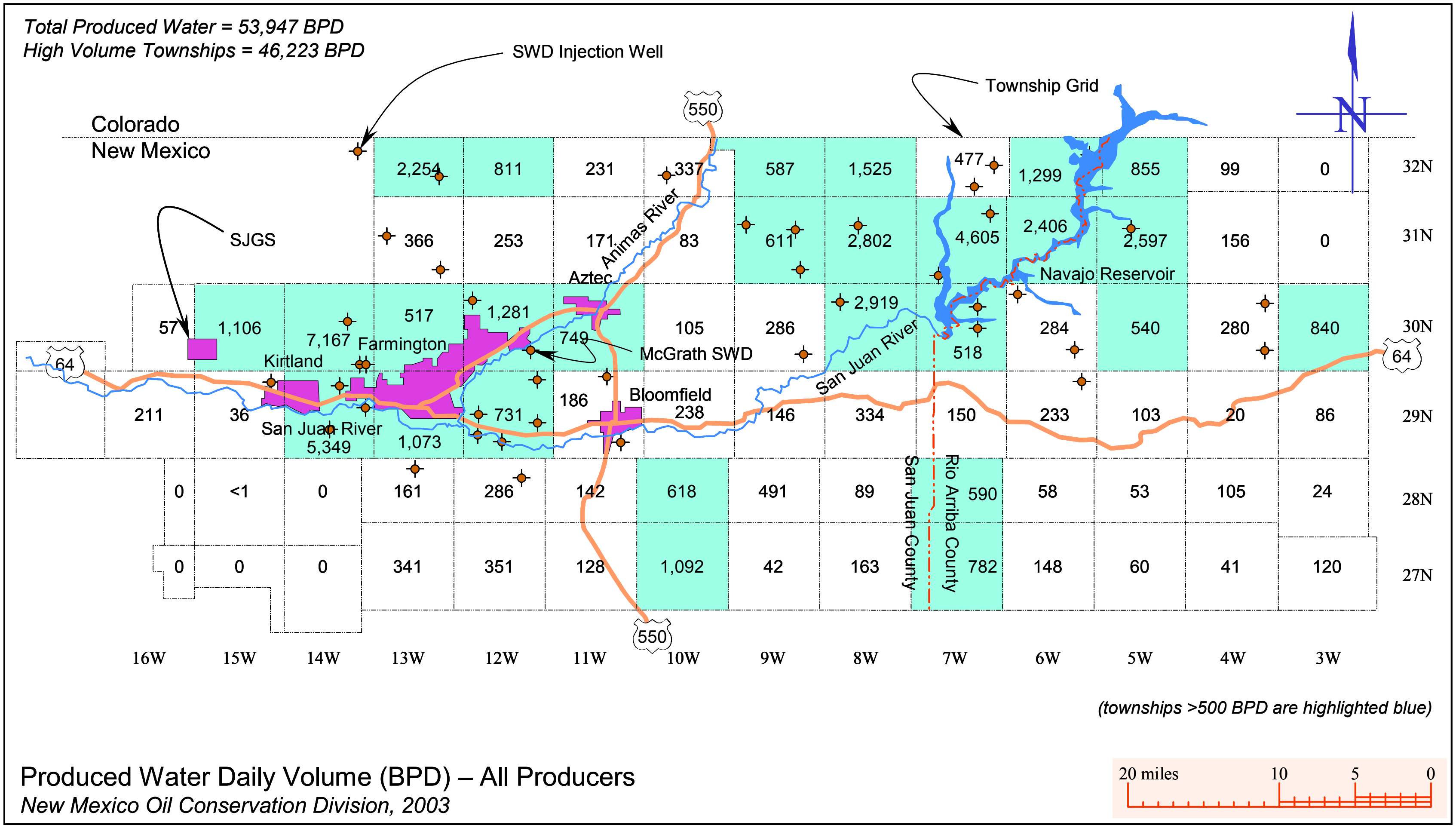

Figure 2-4
Produced Water Daily Volume - All Producers

2.6 Salt Water Disposal Facilities

Produced water is separated from oil and/or gas and stored in a covered atmospheric tank at the wellhead. The water is then transported via tanker truck to a salt water disposal facility (SWD) where it is treated before final disposal by way of deep-well injection. There are 61 SWDs listed as active injection wells (by OCD in 2003) in the Basin in New Mexico. They are operated by 30 entities – large and small oil companies, one refinery and several private treatment and disposal operations. Of these, 44 are in the Study Area and are operated by 20 entities. Figure 2-4 also shows the locations of active SWDs in the Study Area.

Water delivered to a SWD is first passed through an API[6] oil separator to remove solid material (e.g. sand and gravel), oily sludge and floatable oil. After oil removal, the water is filtered to remove fine particulate matter (cartridge-type filtration). A non-oxidizing biocide is usually added to the filtered water to prevent downhole biological fouling just prior to injection into the formation.

SWDs are clustered in areas of high produced water generation to minimize transportation costs of hauling produced water from the wellhead to the disposal well. Hauling frequency depends on the amount of water a well produces (new wells generally produce more water initially – this is especially true for CBM production). Hauling is the largest cost component of produced water disposal. Depending on distance, hauling costs range from $1.00 to $2.00 per barrel and up. Disposal costs vary from $0.25 to $1.00 per barrel.

2.7 Produced Water Generated in Colorado

The focus of this section of the report applies only to produced water generated in the Basin in New Mexico. A significant amount of CBM water is produced in Colorado along the northern edge of the Fruitland. Compacts established between Colorado and New Mexico bar interstate transfers of water without the approval of their respective OSEs. Therefore, this water is considered outside of the scope of this project.

2.8 Future Produced Water Quantities

When a conventional oil or gas well is developed, initial volumes of produced water can be high with a gradual decline over time. Some wells, depending on the formation, generate produced water without a drop-off in volume. CBM wells typically generate high initial volumes of produced water that decline at a greater pace than conventional wells. No effort has been made by any of the producers to predict the decline of produced water generation in any parts of the Study Area. A large producer in the Basin) felt that their water volume might fall by an annual factor of $e^{-0.05}$ *to* $e^{-0.1}$ (equivalent to 4.9% to 9.5%) at current levels of production, i.e. rates of

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[6] The API separator was developed over 70 years ago in a joint effort by the American Petroleum Institute (API) and the Rex Chain Belt Company (currently known as US Filter Envirex Products). The first API separator was commissioned in 1933.

extraction remain the same with no new well installations. Several CBM producers on the western edge of the Fruitland have not seen any falloff in their wells and do not expect to see any in the near future.

All producers are planning more well installations. Accelerated installation of new wells, as a result of denser infill drilling permitted by BLM, will increase near-term produced water generation. On the other hand, stepped up withdrawal will more quickly deplete water in the producing zones. However, many oil field operators do not see a decline in produced water generation in the next 10 to 20 years.

Also, there is a potential to back-flow SWD injection wells to extract previously injected water. Several producing companies have offered this idea as another means of generating produced water. One large producer felt they could generate at least 10,000 BPD by back flowing several of their SWD injection wells. Also, back-flowing could easily be incorporated into a project where produced water is being gathered and conveyed to SJGS.

2.9 Produced Water Chemistry

A sampling and analysis program was conducted to identify the geochemical characteristics of produced water at the McGrath SWD, which is central to conventional oil and gas and CBM production in the Study Area (Figure 2-4). McGrath SWD is owned and operated by Burlington Resources – the largest producer in the Basin in New Mexico. Thirty samples were taken over a 30-day period – one per day at random times. The water quality analysis includes:

- General mineral chemistry – Na^{+1}, Ca^{+2}, Mg^{+2}, alkalinity, Cl^{-1}, etc.
- Heavy metals
- TDS, electrical conductivity and pH
- Ammonia, sulfide and boron
- Total Suspended Solids (TSS) and Total Petroleum Hydrocarbons (TPH)

Refer to Figure 2-5 for a summary of TDS results and Table 2-3 for a summary of produced water chemistry at the McGrath SWD. TDS varied from 6,400 mg/l to 22,600 mg/l. Low TDS water likely was from CBM production to the north and high TDS water from conventional gas production to the west. Other chemistry of interest includes:

- Sodium, chloride and bicarbonate alkalinity predominate the chemistry. This is typical of produced water.
- Relative to total iron content, calcium and magnesium hardness are low.
- Barium and strontium levels averaged 3.1 mg/l and 19 mg/l, respectively.
- Sulfate levels ranged from 168 to 884 mg/l.
- Total and dissolved iron levels were high. Most of the iron comes from above ground carbon steel pipe used to convey produced water.

- Copper, chrome and lead ranged from non-detectable levels to less than 0.050 mg/l. Selenium ranged from non-detectable levels to 0.080 mg/l. Arsenic and mercury were not detected.
- Silica levels were relatively low for produced water – from 12.2 to 27.6 mg/l[7].
- Ammonia levels ranged from 7.0 to 23.0 mg/l.
- Boron levels were typical of many oil field operations – from 1.00 to 3.00 mg/l.
- Sulfide levels were very low – almost always non-detectable. This is characteristic of the Fruitland.
- Total petroleum hydrocarbons (TPH) ranged from 23 to 520 mg/l. High levels of TPH are assumed to be from conventional oil and gas wells. CBM produced water typically has very low levels of TPH – usually <10 mg/l.

There is a significant amount of CBM produced water that is generated near SJGS in townships 29N14W and 30N14W (Figure 2-4). Refer to Table 2-4 for a summary of chemistry for three

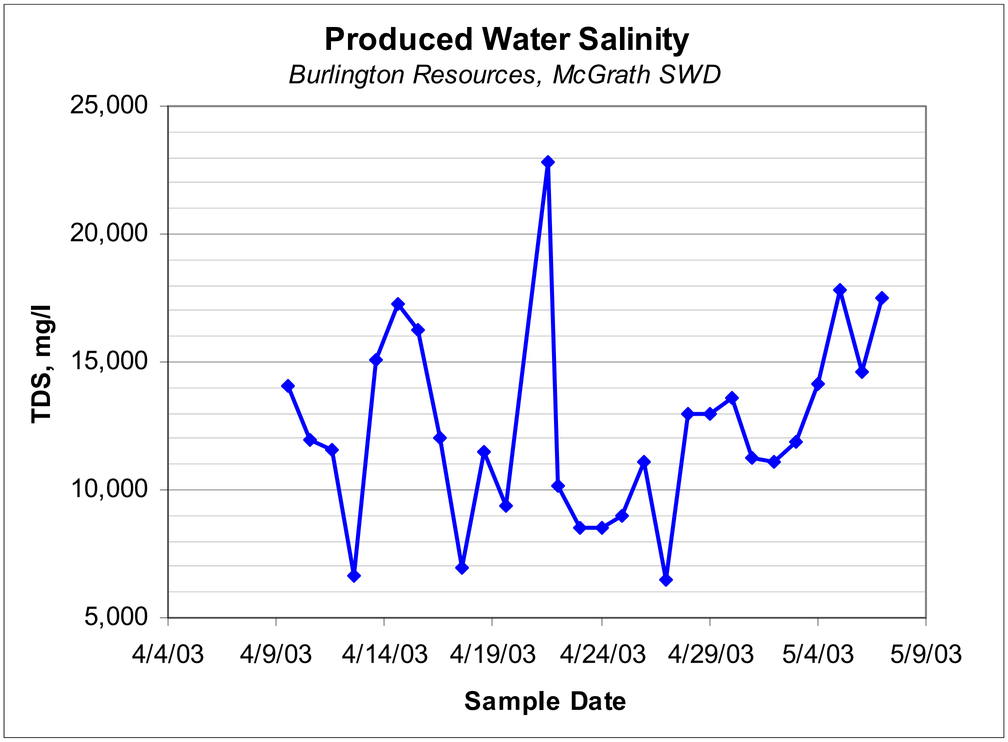

Figure 2-5
Produced Water Salinity, Burlington Resources, McGrath SWD

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[7] Silica can range as high as 150 to 250 mg/l in areas where enhanced oil recovery (steam injection into the producing formation) is practiced.

SWDs. Noteworthy of these chemical analyses is the fact that TDS varies considerably, from 6,300 to 26,100 mg/l. This is due in part to local geology, i.e. the proximity of the wells to the edge of the Fruitland Petroleum System. Many of the chemistry observations cited above hold for this water as well. Produced water chemistry is discussed in more detail in Section 4, Produced Water Use, Treatment, and Disposal Analysis.

The Petroleum Recovery Research Center (PRRC)[8] is currently developing a database of produced water chemistry for the San Juan Basin (as well as other producing units). Current information shows variations in produced water chemistry from north-to-south and east-to-west within the Study Area. Refer to Figure 2-6.

In the east, where CBM extraction predominates, produced water TDS ranges from 8,400 to 13,800 mg/l. Within this area, note how TDS falls as production nears the state border to the north. The highest TDS is south of Highway 64 – approaching 60,000 mg/l. A cluster of data north of Farmington is representative of both conventional and CBM production. TDS of produced water to the west (in Farmington) is higher than that of produced water directly to the east.

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[8] PRRC is a division of New Mexico Institute of Mining and Technology.

Table 2-3
McGrath SWD Chemistry, 30-Day Random Sampling Program

McGrath SWD Chemistry

30-Day Random Sampling Program

		Avg	Min	80th Percentile	90th Percentile	Max
Na (1)	mg/l	4,201	1,862	5,148	6,040	8,055
K	mg/l	177	55.1	282	368	434
Calc'd NH_4	mg/l	16.3	8.93	20.5	24.4	29.5
Ca	mg/l	143	59.8	178	200	311
Mg	mg/l	34.1	12.3	42.7	48.2	88.2
Ba	mg/l	3.08	0.72	4.70	5.54	7.98
Sr	mg/l	19.4	7.19	24.2	31.3	54.7
Dissolved Fe	mg/l	33.1	1.1	42.0	80.4	187.0
Cu	mg/l	ND	ND	NC	NC	0.019
Zn	mg/l	0.230	ND	NC	NC	0.564
As	mg/l	ND	ND	NC	NC	ND
Cr	mg/l	ND	ND	NC	NC	0.035
Pb	mg/l	ND	ND	NC	NC	0.031
Se	mg/l	ND	ND	NC	NC	0.080
Hg	mg/l	ND	ND	NC	NC	ND
Ag	mg/l	NA	NA	NA	NA	NA
U	mg/l	NA	NA	NA	NA	NA
TC	mg/l_{CaCO3}	9,970	4,348	12,316	14,528	19,661
HCO_3	mg/l	764	319	973	1,075	1,298
CO_3	mg/l	0.64	0.10	1.24	1.68	17.3
Cl (1)	mg/l	6,219	2,771	7,601	9,071	12,507
Br	mg/l	14.5	7.13	17.9	19.5	21.8
F	mg/l	ND	ND	ND	ND	ND
NO_3	mg/l	ND	ND	4.43	4.80	5.67
NO_2	mg/l	ND	ND	ND	ND	ND
SO_4	mg/l	544	168	758	810	884
TA	mg/l_{CaCO3}	9,970	4,348	12,316	14,528	19,661
SiO_2	mg/l	18.5	12.2	20.3	24.0	27.6
Total Fe	mg/l	41.3	5.19	69.5	84.7	187
Total Alkalinity	mg/l_{CaCO3}	697	320	868	931	1,100
Total NH_3	mg/l_N	12.8	7.02	16.0	19.1	23.0
B	mg/l_B	2.05	1.00	2.39	2.64	3.00
O-PO_4	mg/l_P	ND	ND	2.33	2.51	2.70
Total Sulfide	mg/l_S	ND	ND	ND	ND	1.60
pH		7.05	6.41	7.25	7.30	8.23
EC	µS/cm	19,880	10,300	23,740	26,690	35,900
TDS (Calc'd)	mg/l	12,210	5,290	15,130	17,820	23,950
TSS	mg/l	108	26	160	211	240
TPH	mg/l	163	23	258	310	520

Notes…..

1. Na and Cl values adjusted (as required) to achieve ionic balance.
2. NA = not analyzed, ND = not detectable, NC = not calculable.

Table 2-4
CBM Chemistry, Close-in Fruitland, Townships 29N/4W and 30N/4W

CBM Chemistry - Close-In Fruitland

Townships 29N14W and 30N14W

		Salty Dog 2/3	Turk's Toast	Taber Locke
Na (1)	mg/l	9,563	2,119	6,848
K	mg/l	149	6.45	25.0
Calc'd NH_4	mg/l	12.4	2.16	121
Ca	mg/l	128	6.27	66.6
Mg	mg/l	87.4	4.34	32.1
Ba	mg/l	20.8	1.86	13.6
Sr	mg/l	20.6	1.73	18.3
Dissolved Fe	mg/l	0.84	<0.01	<0.01
Cu	mg/l	ND	ND	ND
Zn	mg/l	0.298	ND	ND
As	mg/l	ND	ND	ND
Cr	mg/l	ND	0.005	ND
Pb	mg/l	0.036	ND	ND
Se	mg/l	0.017	ND	ND
Hg	mg/l	ND	ND	ND
Ag	mg/l	NA	ND	ND
U	mg/l	NA	ND	ND
TC	mg/l$_{\text{CaCO3}}$	21,697	4,649	15,557
HCO_3	mg/l	1,440	1,952	1,050
CO_3	mg/l	5.51	34.2	0.68
Cl (1)	mg/l	14,518	2,089	10,418
Br	mg/l	15.6	2.74	3.17
F	mg/l	ND	2.30	1.47
NO_3	mg/l	2.55	ND	ND
NO_2	mg/l	ND	ND	ND
SO_4	mg/l	24.9	37.4	ND
TA	mg/l$_{\text{CaCO3}}$	21,697	4,649	15,557
SiO_2	mg/l	9.67	12.2	32.5
Total Fe	mg/l	0.78	4.05	9.08
Total Alkalinity	mg/l$_{\text{CaCO3}}$	1,180	1,910	1,050
Total NH_3	mg/l$_N$	10.6	1.90	94.0
B	mg/l$_B$	2.87	1.60	2.40
O-PO_4	mg/l$_P$	ND	ND	ND
Total Sulfide	mg/l$_S$	ND	17	NA
pH		8.23	8.82	7.40
EC	µS/cm	40,300	9,160	29,900
TDS (Calc'd)	mg/l	26,010	6,300	18,660
TSS	mg/l	42	16	18
TPH	mg/l	ND	17	2.3

Notes…..

1. Na and Cl values adjusted (as required) to achieve ionic balance.
2. NA = not analyzed, ND = not detectable.

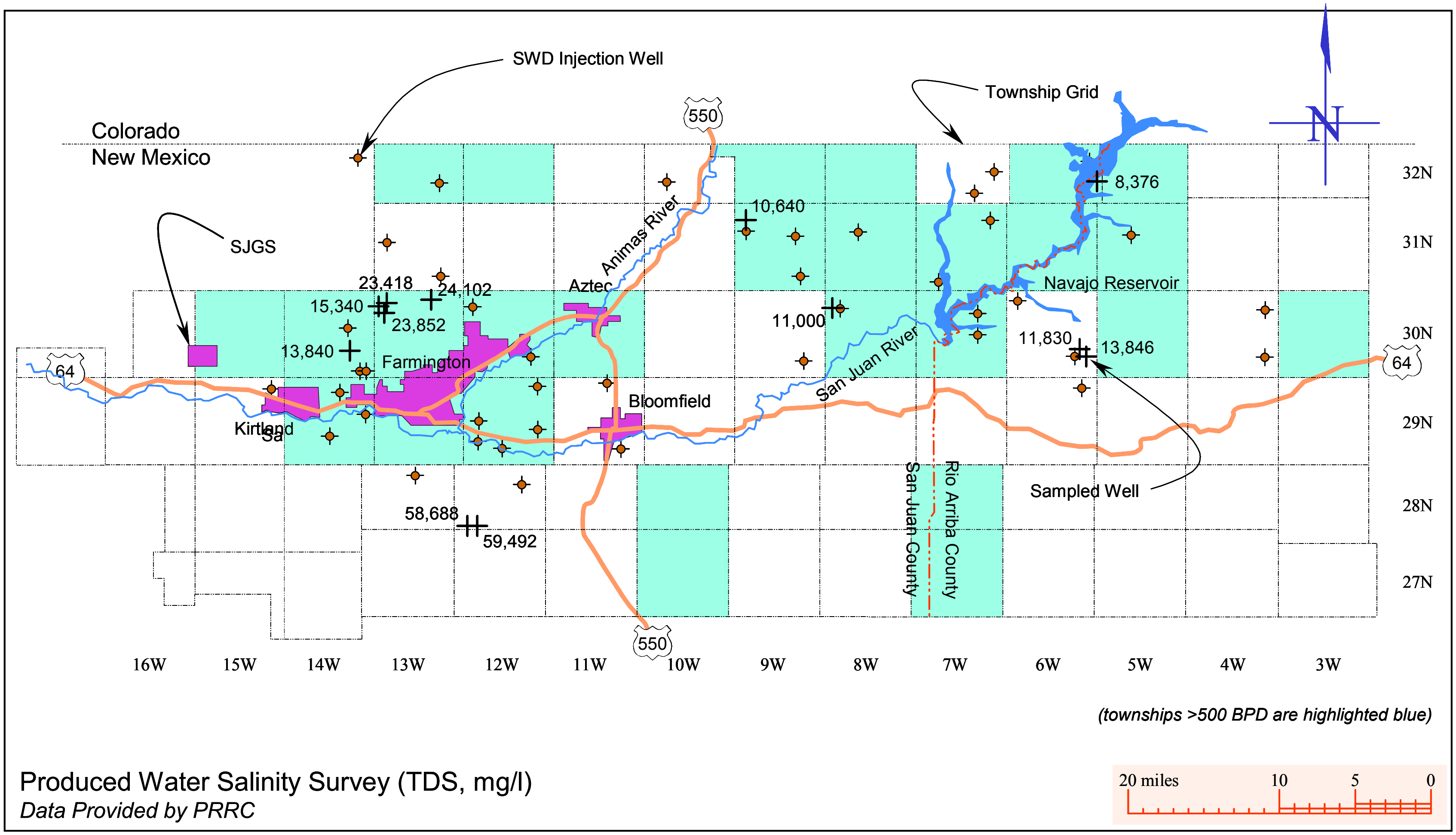

Figure 2-6
Produced Water Salinity Survey (TDS, mg/l)

3

PRODUCED-WATER INFRASTRUCTURE AVAILABILITY AND TRANSPORTATION ANALYSIS

3.1 Introduction

Transportation is the largest obstacle to produced water reuse in the San Juan Basin (the Basin). Most of the produced water in the Basin is stored in tanks at the wellhead and must be transported by truck to salt water disposal (SWD) facilities prior to injection. Depending on the location of a well, one-way transport can exceed several hours. Also, relative to other producing areas in the United States, water generation in the Basin is spread over a large area, i.e. wells are on 80-acre parcels in the Basin[1] as compared to 100-foot centers (or less) in parts of Texas, Oklahoma and California. Produced water transportation requirements from the wellhead to San Juan Generating Station (SJGS) and the availability of existing infrastructure to transport the water are covered in this section.

There is a broad network of oil and gas gathering and transmission pipelines in the San Juan Basin. In a typical operation, oil and/or gas are separated from produced water at the wellhead. The majority of hydrocarbon production in the Basin is natural gas. Gas is compressed at the wellhead and fed to a network of gathering lines where it can be stored and pre-treated (at an intermediate facility) and transported to a gas treating facility in Bloomfield, New Mexico or outside the Basin. Gas treatment consists of water and CO_2 removal, de-sulfurization and the separation and/or blending of different hydrocarbon constituents, e.g. methane, ethane, propane, etc. From Bloomfield, treated gas is transported to points north, south or west via gas transmission lines. Oil is usually stored at the wellhead and trucked to a central location for delivery by truck or pipeline to the Giant Refinery[2] in Bloomfield where it is de-sulfurized and processed into gasoline, diesel, heavy fuel oils, etc.

SJGS is located about 18 highway miles west of the center of Farmington, New Mexico and 30 miles west of Bloomfield. Refer again to Figure 2-2, for a depiction of the Study Area. Also, SJGS is located on the western edge of the Fruitland Petroleum System (the Fruitland), placing it just outside of areas of oil and gas production. As described in Section 2, *Assessment of Produced Water from Oil and Gas Wells*, the Study Area is 31 miles wide by 84 miles long at its

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[1] The Bureau of Land Management (BLM) just increased the limit from one well per 160 acres to one per 80 acres on federal lands.

[2] The Giant refinery in Gallup, New Mexico, which is 100 miles SSW of Bloomfield, may also receive oil from the Basin.

greatest dimensions. Produced water is generated in 69 of the 78 townships in the Study Area. Refer to Figure 3-1 for produced water generation by township.

3.2 Produced Water Generation

The Study Area was established to identify produced water that is reasonably close to SJGS. There were 19,090 oil and gas wells listed as active in the Basin in New Mexico in 2003[3] (13,600 wells in the Study Area). Wells in the Basin generated about 68,500 BPD of produced water – 53,900 BPD in the Study Area.

A significant amount of produced water is generated in the Colorado portion of the San Juan Basin (just north of the Study Area). Compacts established between Colorado and New Mexico bar interstate transfers of water without the approval of their respective Offices of the State Engineer (OSE). Therefore, Colorado produced water collection, conveyance and reuse are not included this evaluation.

Refer to Table 3.1 for a summary of produced water generation in the Study Area.

Table 3-1
Summary of Produced Water in the Study Area

Produced WaterGeneration	Townships	Township Volume (barrels)	Percent of Total	Cumulative Volume (barrels)
0 BPD	8	0	0.0%	0
1 to 299 BPD	36	5,022	9.3%	5,022
300 to 699 BPD	14	6,680	12.4%	11,702
700 to 1,499 BPD	11	10,619	19.7%	22,321
1,500 to 2,999 BPD	6	14,504	26.9%	36,825
3,000+ BPD	3	17,122	31.7%	53,947
Total	78	53,947	100.0%	

Produced water generation patterns for the Study Area are summarized below:

- About 42,000 BPD or 80 percent of the produced water in the Study Area is generated north of or at Highway 64.
- 20 townships produce more than 700 BPD of water in the Study Area and generate 29,400 BPD or 70 percent of the produced water in the Study Area. Fifteen of these townships are located at or north of Highway 64.

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[3] Reported by the Oil Conservation Division (OCD). Oil and gas production statistics are compiled and made available to the public on their website at www.emnrd.state.nm.us/ocd.

- Two townships on the western edge of the Fruitland generate the most produced water in the Study Area – 12,516 BPD or 23 percent of the daily volume in the Study Area.
- Two of the high-volume townships are split by the San Juan River and two townships are south of it. Produced water south of Highway 64 comprises a small fraction (about 10 percent) of available water in the Study Area and is generally more saline.

3.2.1 Areas of High-Volume Production

Figure 3-1 identifies three areas of high-volume produced water generation in the Study Area. They are designated in this section as:

- Close-in production (12,520 BPD) in two Kirtland area townships – 30N14W and 29N14W (which is bisected by the San Juan River) – about 5 to 10 miles from SJGS
- Fairway production (17,760 BPD) in ten townships from 31N to 32N and 5W to 9W
- Tri-City production (2,760 BPD) in three townships in the Aztec-Bloomfield-Farmington area.

These areas were selected for several practical reasons:

- Close-in production is in the vicinity of SJGS (the plant can be seen from many of the wellheads).
- There is pipeline infrastructure that runs west and north of Bloomfield to Fairway production.
- Produced water that is trucked to SWDs in the Tri-City Area (as well as surrounding low-volume areas) could easily be re-routed to existing (or new) infrastructure.

It is noteworthy that most of the SWDs in the Study Area are situated in areas of high water production – shorter distances to injection wells reduce transportation costs[4]. Additionally, there are only a handful of SWDs south of Highway 64 (mostly because of reduced water production). This water must be transported north at substantial cost to the producers. Therefore at this point in the report, the Study Area has been reduced from 2,400 square miles as described in Section 2, Assessment of Produced Water from Oil and Gas Wells, to 1,500 square miles to focus on high-volume areas of produced water generation. It is delineated by townships – 32N5W (northeast corner) to 29N14W (southwest corner).

Infrastructure and the rationale for transporting produced water from high-volume areas are discussed in more detail later in this section.

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[4] Transportation by tanker truck (250 barrel capacity) accounts for 50 to 80 percent of produced water handling costs depending on the wellhead distance to SWDs.

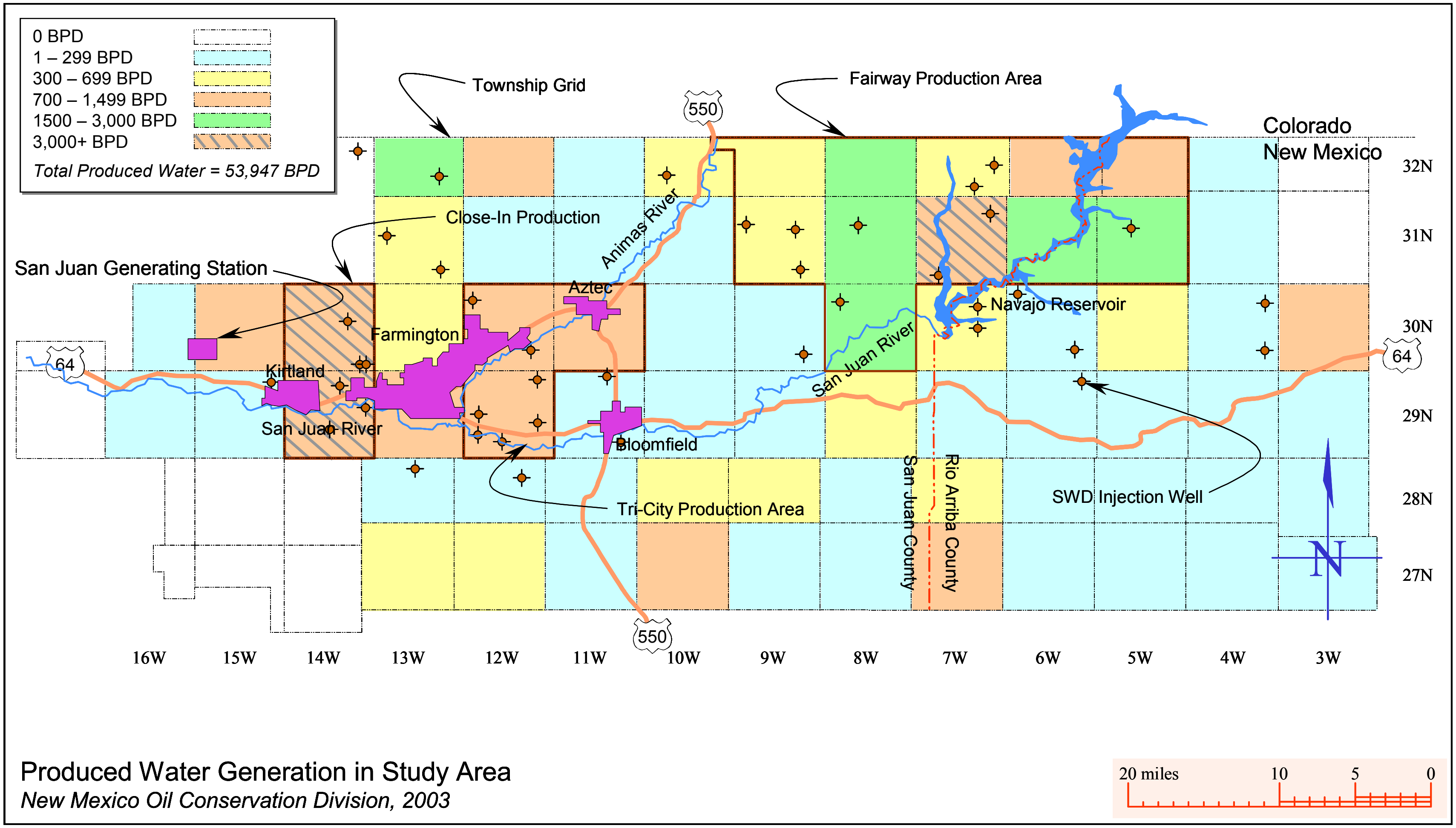

Figure 3-1
Produced Water Generation in Study Area

3.3 Major Gas Transmission Infrastructure

Bloomfield is the hub of oil and gas production and processing in northwest New Mexico and is home to five gas processing plants and one oil refinery. Consequently, there are a number of major gas transmission lines in the Study Area. Refer to Figure 3-2. At the start of this project, it was assumed that abandoned or underutilized gas transmission lines could provide an ideal means of conveying produced water from areas of high-volume production to SJGS. Gas transmission rights-of way could also provide an established path for a new produced water pipeline to SJGS.

In the past few years, however, the demand for natural gas has skyrocketed in the western United States. Numerous gas-fired combined cycle power plants[5] have come on line in California, Arizona and Nevada. Increased demand and the need to move greater volumes of natural gas have created a shortage of gas transmission infrastructure. This has spurred new pipeline construction to transport gas from Texas, Oklahoma, New Mexico and Colorado to western states. The demand for new pipelines has eliminated any heretofore excess capacity that may have existed. A number of major natural gas pipeline companies were contacted to determine the availability of abandoned or underutilized pipeline – all existing pipeline assets are fully utilized. Even older/low pressure lines are being kept in service and used for gathering purposes.

As stated previously, natural gas transmission line rights-of-way could provide established pathways to SJGS and several gas pipeline companies have stated (in principle) that their rights-of-way could be made available for a produced water pipeline.

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[5] Combined cycle plants utilize a gas turbine to drive an electric generator and a heat recovery steam generator (utilizing the hot exhaust from the gas turbine) to drive a steam turbine/electric generator. Natural gas is the primary fuel source for gas turbines.

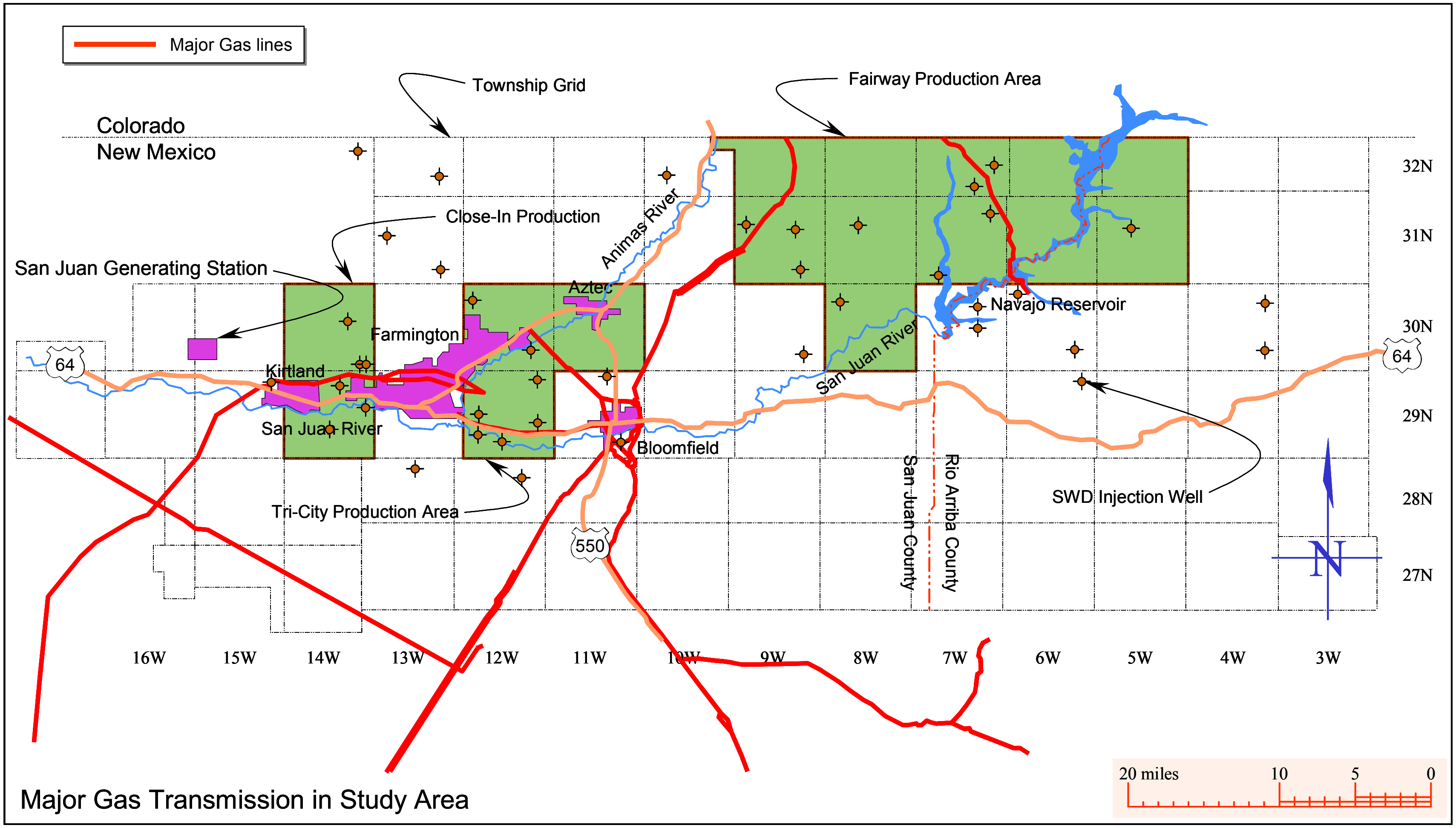

Figure 3-2
Major Gas Transmission Lines in Study Area

3.4 Other Pipeline Infrastructure

Burlington Resources, the largest producer in the Basin, was also consulted to determine the extent of pipeline infrastructure in the Basin that could be used for transporting produced water. They were also instrumental in identifying how produced water is handled, i.e. separated from oil and/or gas at the wellhead, transported to SWDs and treated prior to deep-well injection.

Burlington Resources identified two abandoned pipelines that could be used to gather produced water:

- CO_2 Gas Line – 4” high-pressure carbon steel line originally constructed to transport CO_2 to evaluate a production technique to displace methane from coal
- Hart Canyon Line – 4” high-pressure carbon steel line previously used to transport produced oil to the refinery in Bloomfield.

Both lines are owned by Burlington Resources and are preserved-in-place for possible future service. Refer to Figure 3-3. The CO_2 Gas Line originates close to Bloomfield and threads its way past a number of SWDs and terminates close to the New Mexico-Colorado border in the center of the Fairway Production Area. The Hart Canyon line extends north from Bloomfield and is situated between the Tri-City and Fairway Production Areas. As discussed next, both lines are well situated and could be used for produced water gathering.

Lastly, discussions with other large producers (by way of introductions from Burlington Resources) did not yield any other significant infrastructure. Many lines have been abandoned and not preserved-in-place so the condition of this buried pipe is presumed poor. Some lines have been cut and the remaining sections re-routed. Many of the unused segments are short and of little use.

3.5 Produced Water Gathering, Staging and Conveyance

Given the orientation of the three high production areas in the Study Area and the orientation of the CO_2 Gas Line and the Hart Canyon Line (refer to Figure 3-3), four gathering, staging and conveyance strategies emerged:

- Use the CO_2 Gas Line and the Hart Canyon Line to gather produced water from the Tri-City and Fairway Areas.
- A Collection Center could be constructed in Bloomfield to accept and pretreat produced water prior to conveyance to SJGS.
- A new pipeline could be constructed to convey produced water from the Collection Center in Bloomfield to SJGS.
- Gather produced water directly from two or more Close-in Area producers using the new Bloomfield-to-SJGS produced water pipeline.

Refer to Figure 3-4 for a schematic of the gathering, staging and conveyance strategies.

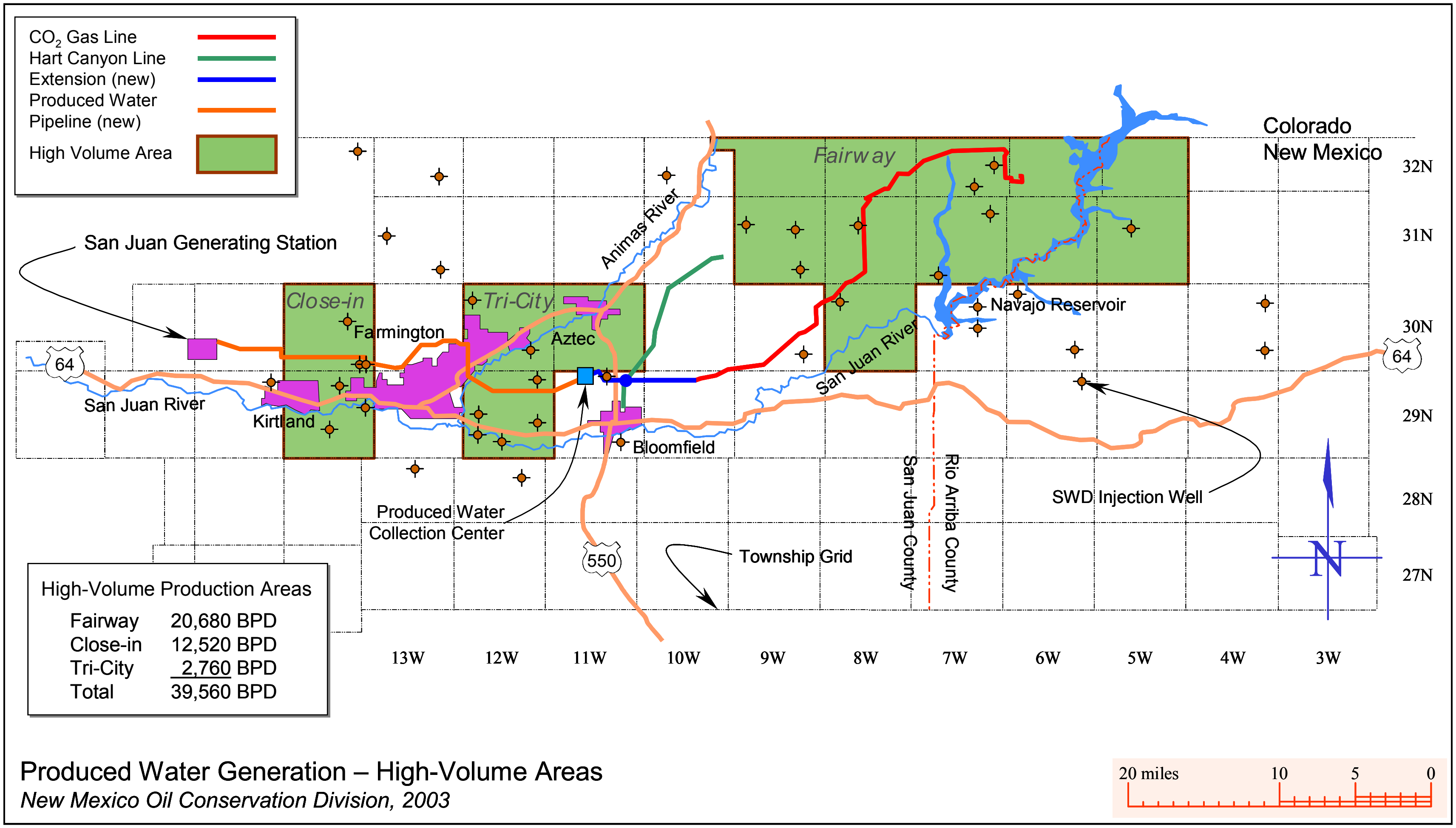

Figure 3-3
Produced Water Generation in High-Volume Areas

3.5.1 CO_2 Gas Line and Hart Canyon Line

The CO_2 Gas Line and the Hart Canyon Line provide a convenient and direct means of gathering produced water from the Tri-City and Fairway Areas. Given their relative orientation, they could either be tied together and routed (as a new line) to a collection point or, depending on the location of the collection point, they could be routed to it separately. After discussions with Burlington Resources, it was determined that it would be more practical to combine the produced water flow of the CO_2 Gas Line and the Hart Canyon Line and route a new extension line to the Collection Center.

The CO_2 Gas Line and the Hart Canyon Line would receive produced water from a series of injection points – possibly three to four in each line (the CO_2 Line could have more because of its greater length). Refer to Figure 3-4. Two means of injection surfaced in discussions with Burlington Resources:

- Satellite collection stations would receive produced water from transport trucks. They would be located in areas of high traffic to optimize daily volume. Each station would have a receiving tank, transfer pumps and filters and on a pre-programmed schedule would inject filtered produced water into either the CO_2 Gas Line or Hart Line. Each truck would be given an electronic identification card to track who used the system. In the event there was a problem with either vandalism or improper disposal of a waste product, specific operators would be prohibited from disposing of produced water.
- SWD direct injection would be used for a number of nearby injection wells. SWD operations remove oil and grit from produced water and filter it before injection to protect the well and receiving formation. Filtered water would be injected into either the CO_2 Gas Line or Hart Line.

3.5.2 Collection Center in Bloomfield

The Collection Center in Bloomfield would be used to:

- Receive produced water via the intertie/extension of the CO_2 Gas Line and Hart Line.
- Remove oil and grit using a three-step process – API gravity/coalescing separation, dissolved air flotation and walnut shell filtration.
- Equalize chemistry via storage to reduce variations in produced water salinity.
- Monitor water quality prior to charging the conveyance pipeline – oil content, suspended solids, pH, salinity, etc.

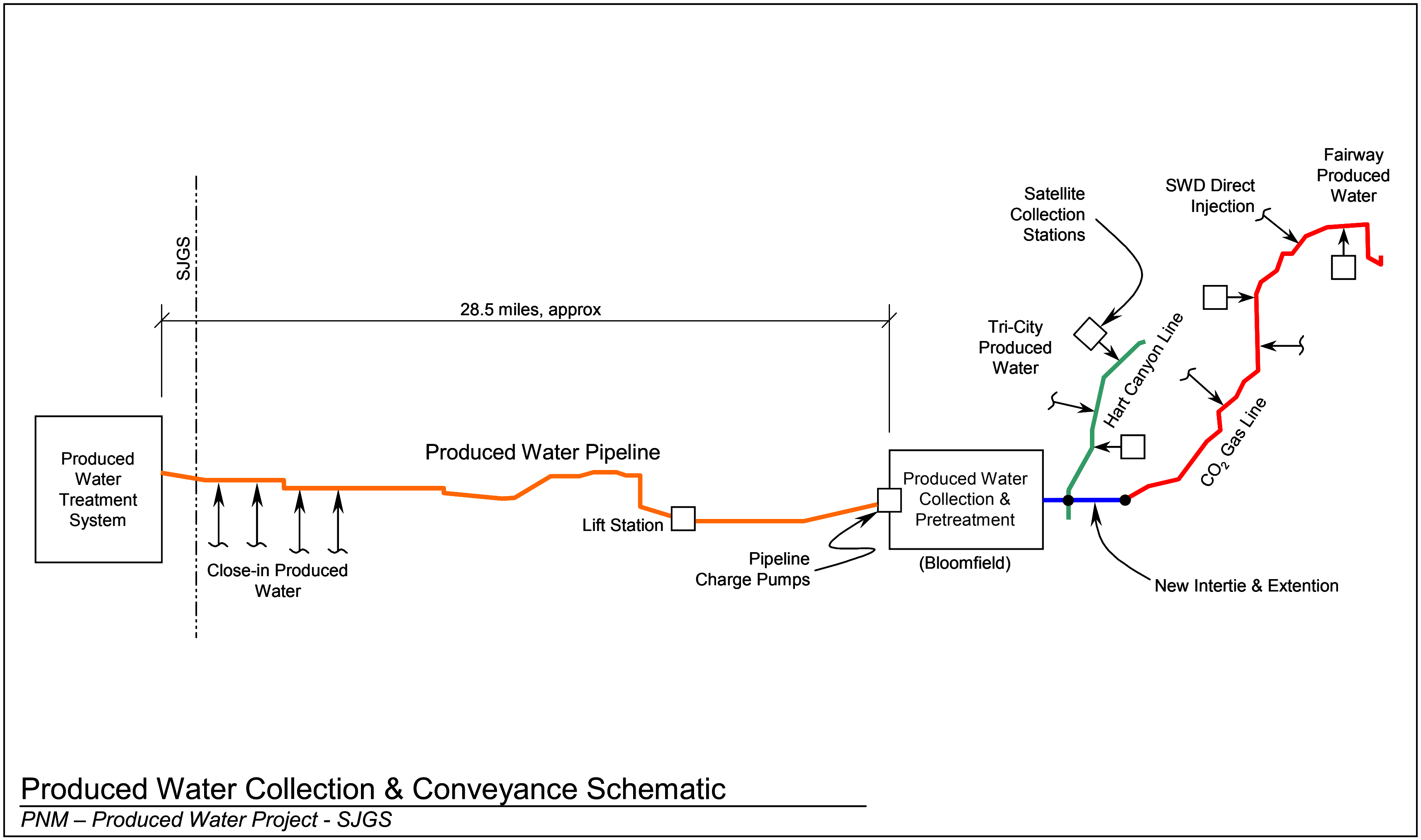

Figure 3-4
Produced Water Collection and Conveyance System Schematic

Water to be transported in the pipeline from Bloomfield to SJGS must be free of oil, grit and suspended matter to protect its integrity. Final water treatment (desalinization) could be also done at the Collection Center in Bloomfield rather than SJGS; however, the environmental and economic issues with associated brine and sludge disposal will likely preclude this. Produced water handling, pretreatment, conveyance and treatment alternatives are discussed in Section 4, *Treatment and Disposal Analysis*.

Lastly, rerouting produced water transport trucks to the Collection Center in Bloomfield (or to SJGS) was not considered feasible. Trucking is the largest cost component of handling, treating and injection produced water, so hauling water extra distances would only raise the cost of disposal for the producers (and discourage participation in produced water reuse). Also, there are times when trucks deliver fluids to SWDs that cannot be injected. It was felt that the SWD operators were better equipped to monitor/control this activity.

3.5.3 Produced Water Conveyance

A 28.5-mile, 14-inch pipeline would be required to convey produced water from the Collection Center in Bloomfield to SJGS. Public Service of New Mexico (PNM) enlisted the services of a local engineering firm to evaluate pipeline routes and costs as well as identify locations for the Collection Center. The most cost effective route is shown in Figure 3-3. The route selected is the shortest and takes advantage of PNM transmission right-of-way the last third of the pipeline length. The pipeline would be constructed of high-density polyethylene (HDPE) and would consist of charging pumps, a mid-length lift station and clean-out stations along its length. The elevation change of the line is predominantly downhill but there are several lifts that must be overcome. The line would be designed for an operating pressure of approximately 200 psi. The line was purposely sized large to accommodate up to 60,000 BPD of produced water (44,700 BPD during peak collection years is the likely flow rate) in the event additional water is available from future increased gas production. Refer to Table 3-2 for the pipeline design basis and Table A.1 in the Appendix for installation and operating costs.

Table 3-2
Pipeline Design Basics

Design Flow Rate	60,000 BPD
	2,823AF/yr
	1,750 gpm
Pipeline Length	28.5 miles
Pipeline Diameter	14 inches
Pipeline Material	HDPE
Cleanout Stations	10
Charge pressure	300 psi
Lift Pressure	300 psi
Charge/Lift Power	328 kw

3.5.4 Close-in Area Produced Water

Close-in Area CBM (coal bed methane) production from the Kirtland area would be collected directly by the new 28.5-mile pipeline. Typically, CBM water has no measurable free oil[6] content. In comparison, conventionally produced oil and gas can have very high levels of free oil products (in excess of 500 mg/l). Simple filtration is all that is necessary to pretreat the CBM produced water before it is injected into the pipeline. Producers already filter the water prior to injection so they would not have to perform any special treatment prior to the pipeline. Refer to Figures 3-3 and 3-4. Produced water would be collected from the two major gas producers in townships 29N14W and 30N14W, Dugan Production Corporation and Richardson Operating Company. In addition to produced water, cooling tower blowdown from Prax Air (a small industrial operation in Kirtland, New Mexico) and water from the BHP Billiton mine (coal supplier to SJGS) can readily be picked up by the pipeline[7] because it passes both of these operations. Prax Air and BHP Billiton would also pump their water into the pipeline.

3.6 Other Sources of Produced Water

There are two additional sources of produced water in the Study Area that should be investigated. Both could further enhance produced water recovery volume.

There is a large independent disposal operation in the vicinity of the Collection Center in Bloomfield (about three miles east) that injects approximately 10,000 BPD of produced water. Many small and intermediate-sized producers utilize their services in lieu of installing their own injection facilities. Typical of SWDs, the facility pretreats produced water (oil and grit removal followed by and filtration) prior to injection. This operation should be considered a possible resource and investigated further.

Those SWDs that can be utilized to pump filtered produced water to the CO_2 Gas Line, Hart Line or directly into the 28.5-mile pipeline may also have the ability to backflow formations that formerly accepted produced water. Oil company geologists[8] feel that many injection wells (not all) can be used for this purpose. To accomplish this, a pump would have to be inserted into an injection well (capable of backflowing) to extract produced water. It is felt that the McGrath SWD (operated by Burlington Resources) could be converted to a backflow well, capable of generating up to 5,000 BPD of previously-injected produced water. An additional 10,000 BPD of produced water may be available in the Study Area from backflowing.

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[6] CBM gas is collected directly from coal bed seams. The seams are fractured to allow trapped gas and water to escape. Separable hydrocarbons in the coal are usually in the form of methane gas. It is rare to find higher molecular weight hydrocarbons (in liquid form) such as butane or pentane.

[7] Their contribution to would amount to 1,400 BPD of a possible 40,000 BPD project, about four percent of total project flow.

[8] Discussions with geologists at Burlington Resources and Dugan Production Corporation.

3.7 Recent Legislative Changes and Phased Implementation

As summarized in Section 1, Produced Water Assessment, a bill allowing the “disposal” of produced water use at electric generating facilities was proposed in the 2004 New Mexico legislative session. The bill had two provisions. First, produced water reuse would be designated as an alternate method of disposal (rather than a beneficial use). Second, tax credits would be granted for using produced water at a power plant. Beneficial use of produced water was a major obstacle to oil and gas producer participation in any water reuse plan[9]. Also, tax credits are required to help pay for the new infrastructure[10] necessary to convey produced water from Bloomfield to the SJGS. The provision allowing disposal at a power plant passed, however, the tax credit did not. If the tax credit provision is to be pursued, it must be reintroduced in an upcoming legislative session.

As a result of these legislative actions and given the cost of new infrastructure, PNM is evaluating a phased approach to using produced water at SJGS:

Phase 1. Build a new 11-mile pipeline to gather and convey Close-in production from the Kirtland area to SJGS. The pipeline would be either be sized to just accommodate Close-in produced water daily volume to minimize front-end project costs, or sized to accommodate full-project throughput.

Phase 2. Gather Fairway and Tri-City production utilizing the CO_2 Gas Line and the Hart Canyon Line. This alternative would involve Burlington Resources as a project participant. A new Collection Center would be built in the Bloomfield area to pre-treat (and possibly treat for end use) produced water. The Phase 1 portion of the pipeline would be extended an additional 17.5 miles or a new 28.5-mile pipeline would be built from the Collection Center to SJGS. The implementation of this phase will be influenced by passage of the tax credit legislation.

Specific project details are discussed in Section 8, *Implementation Requirements*.

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[9] Under beneficial use, a right to use the water must be obtained. Also, it must be demonstrated that the produced water being considered has no hydrologic connection to other waters of the state. The regulatory and environmental protection afforded by the OCD (designating the water as a byproduct of oil and gas production) would have been lost to producers with beneficial use.

[10] Infrastructure includes the produced water collection and treatment center in Bloomfield and the 28.5-mile pipeline.

4

PRODUCED WATER USE, TREATMENT, AND DISPOSAL ANALYSIS

4.1 Introduction

Produced water use at San Juan Generating Station (SJGS) is evaluated in this section. Previous sections addressed the produced water resource in the San Juan Basin and the infrastructure required to deliver it to SJGS.

Two approaches are employed to evaluate the use of produced water at SJGS:

- Use produced water "as is" by feeding it directly to major process area(s) in the plant, e.g. take advantage of significant dilution by blending produced water with plant freshwater and using it for make-up to the cooling towers.
- Treat produced water and use it with minimal restrictions in the plant.

Before evaluating these approaches, a simplified water balance is presented to show how water is used and reused at the plant. Water quality constraints are then established for each major water user and produced water chemistry is assessed against these constraints. It is shown in this evaluation that produced water must be treated to justify using it in any reasonable quantity at SJGS.

Produced water treatment alternatives are evaluated utilizing off-the-shelf technology. Water treating equipment at SJGS is also incorporated into the evaluation. The economics of produced water treatment is assessed and a produced water treatment alternative is selected[1].

4.2 Water Use at SJGS

High quality water from San Juan River is withdrawn and stored in a 30-day pond on the plant site. SJGS uses 22,400 acre-feet of water per year (equivalent to 13,890 gpm) of San Juan River – the only source of water for the plant. The plant is a zero liquid discharge facility and, as such, is well practiced in the efficient use and reuse of water. The plant recycles most of its wastewater and uses evaporation ponds for final disposal.

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[1] A full-project economic analysis for produced water collection, pipeline and treatment is developed in Section 7, *Cost/Benefit Analysis*.

4.2.1 Plant Water Use

The plant uses, reuses and treats water for reuse, consumes water in the form of non-recoverable losses of water to process, and eventually disposes of wastewater. Table 4-1 identifies five categories of plant water, including the major process streams. Each stream is designated with a number, description, category, annual average flow rate and applicable water quality constraint(s). Figure 4-1 presents a simplified schematic of the plant water system at SJGS. Stream numbers in Table 4-1 correspond to the stream numbers in Figure 4-1.

Table 4-1
Plant Water System Flow Rates of Major Streams, San Juan Generating Station

Stream	Description	Type (1)	Flow Rate gpm (2)	Major Water Quality Constraints (3)
1	Total Plant Feed	FW	13,890	
2	Cooling Tower Make-up	FW	12,480	chloride, calcium, sulfate, silica
3	Ash System Make-up	FW	100	TDS
4	Limestone Prep Fresh Make-up	FW	1,210	chloride, magnesium
5	CT Evaporation & Drift	Lost	11,640	
6	CT Blowdown	RW	1,000	
7	Boiler Blowdown	RW	430	
8	Plant Drains	RW	100	
9	Process Pond Recycle	RW	1,530	
10	Recycle to LS Prep	RW	730	chloride, magnesium
11	Brine Concentrator Feed	RW	800	chloride, boron
12	BC Distillate to CT	TRW	165	
13	BC Distillate to Demineralizers	TRW	620	
14	BC Brine to Evaporation Ponds	WW	15	
15	Boiler Feed Water	TRW	620	
16	Spent Regenerant (5)	RW	<2	
17	Limestone Prep Total Make-up	FW/RW	1,940	chloride, magnesium
18	Absorber Purge Water	WW	100	
19	Water Lost to Absorber Cake	Lost	140	
20	Water Lost to Ash System	Lost	100	
21	Water Lost to Flue Gas	Lost	1,700	
22	Steam Losses	Lost	190	
23	Plant Service Water	FW	100	TDS

Notes:

FW = freshwater (San Juan River), RW = recyclable wastewater, TRW = treated recycled wastewater, WW = non-recyclable wastewater, Lost = water lost to process (not recoverable).

Flow rates are best estimates based on a variety of plant data sources. Flow rates are based on an annual average plant operating capacity of 79.5% for a total consumption of 22,400 AF/year. Flows are rounded to the nearest 10 gpm except for BC brine and distillate recycled to cooling tower.

Some systems, such as the cooling tower, have numerous constraints. The constraints identified in the table are considered major water quality concerns relative to the use of produced water at SJGS.

Refer to Figure 4-1, Simplified Water Balance.

Demineralizers regenerate very infrequently because they receive low-TDS distillate.

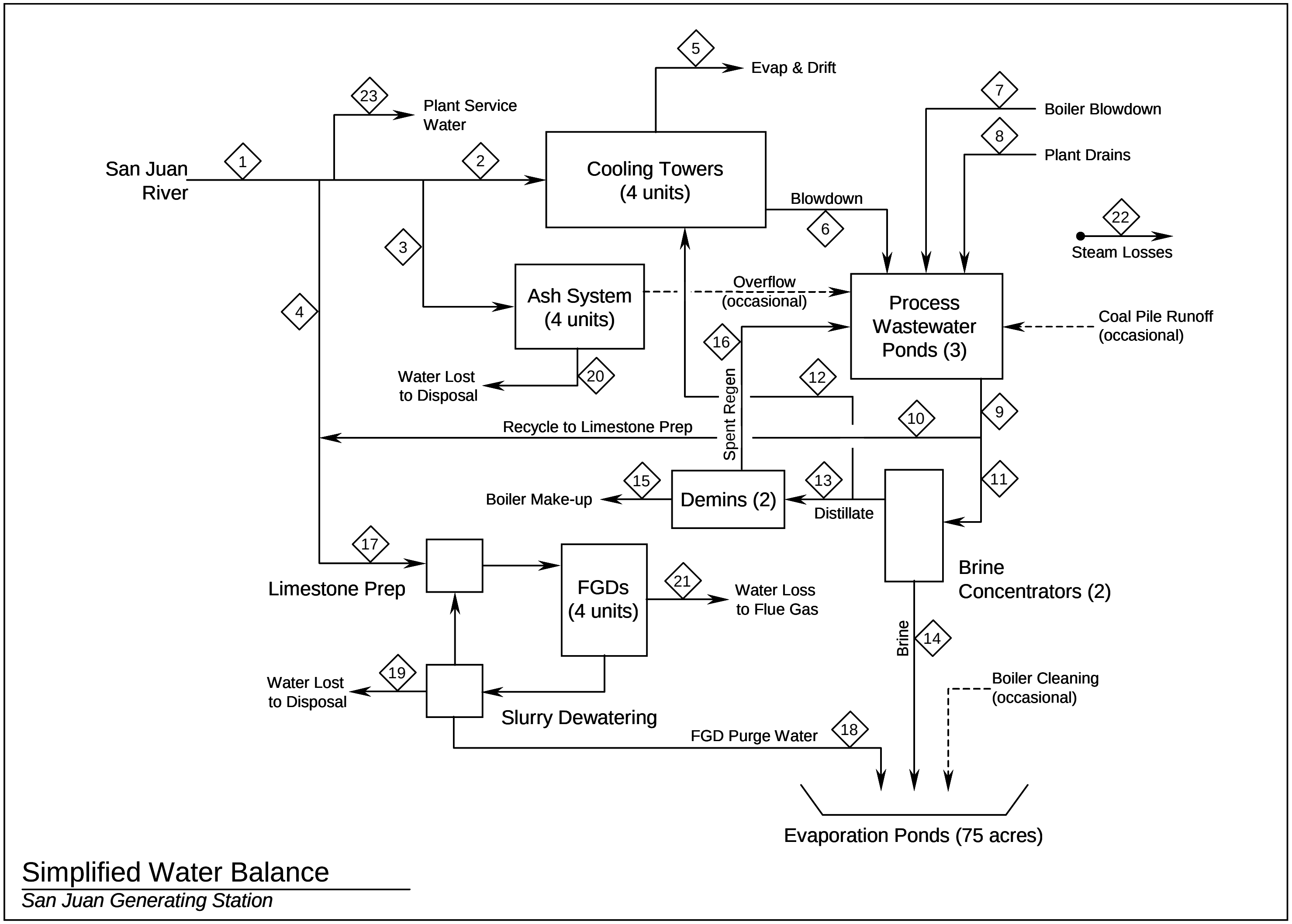

Figure 4-1
Simplified Plant Water System Flowsheet, San Juan Generating Station

Freshwater (FW). San Juan River water is primarily used for cooling tower make-up, absorber[2] make-up, ash system service (bottom ash sluicing, fly ash wetting, seal water, etc.) and plant service water. Some freshwater is required for the absorbers, because recycled wastewater can only supply a portion of their needs. Also, water quality limitations of the absorbers require some freshwater (discussed later).

Recyclable Wastewater (RW). This water is collected in the three Process Wastewater Ponds at SJGS and consists of cooling tower blowdown, boiler blowdown, spent regenerant (from the boiler feedwater demineralizers) and plant drains (primarily service water used for housekeeping and maintenance). A portion of it is used for Absorber make-up and the rest is treated by brine concentrators[3] (BCs) for reuse as boiler feedwater and cooling tower make-up. There is significant flexibility in the wastewater recycle system. Wastewater destined for recycle can be transported from/between any of three Process Wastewater Ponds for reuse or treatment.

Treated Recycled Wastewater (TRW). Approximately half of the water from the Process Wastewater Ponds is treated with BCs. High-quality distillate (TDS < 10 mg/l) from the BCs is further treated by two sets of demineralizers (one for each unit pair) for boiler feedwater. Excess distillate is sent to the cooling towers for reuse.

Non-Recyclable Wastewater (WW). These streams are not useable. They cannot be treated by the BCs (because of water quality limitations) and are sent to the evaporation ponds for final disposal. Of the 13,890 gpm of water used by SJGS on an annual average basis, less than one percent is sent to final disposal in the evaporation ponds (~110 gpm).

Water Lost to Process (Lost). These process streams are not recoverable and consist of cooling tower evaporation and drift loss, absorber water lost to flue gas, steam losses from the power block and waters of moisture and hydration lost to bottom and fly ash and absorber sludge cake. The cooling towers consume the most water (84 percent lost to the atmosphere) followed by the absorbers (12 percent).

4.2.2 Water Quality Constraints in the Process Areas

Major plant process areas – cooling towers, absorbers, ash systems and boilers – all have operating controls and limitations that are related to water quality[4]. Operating constraints for each system are discussed next. Refer to Table 4-1 and Figure 4-1.

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[2] SJGS refers to flue gas de-sulfurizers (FGDs) as absorbers. Another term for this equipment is SO_2 scrubbers.

[3] Brine concentrators are also known as VCEs (vapor compression evaporators) or just evaporators. SJGS refers to this equipment as BCs.

[4] Operating constraints are put into place to prevent corrosion and mineral scale formation, maintain equipment performance and reliability, establish a safe work environment, etc.

Cooling Towers

The allowable cycles of concentration for the cooling towers are controlled by water quality criteria, i.e. levels of calcium (Ca), sulfate (SO_4), silica (SiO_2) and chloride (Cl) among other criteria. The criteria were developed for San Juan River, which is characterized by relatively low levels of TDS. Refer to the following general mineral analysis[5].

San Juan River Water Mineral Analysis

Na	29 mg/l
K	3 mg/l
Ca	54 mg/l
Mg	11 mg/l
HCO_3	125 mg/l
Cl	22 mg/l
SO_4	107 mg/l
SiO_2	12 mg/l
TDS	360 mg/l
pH	8.0

The cooling towers for Units 1, 2 and 4 are operated at 10 cycles of concentration with the limiting factor being calcium sulfate ($CaSO_4$). Therefore, calcium is kept at or below 1,600 mg/l_{CaCO3}. Silica (SiO_2) is kept at or under 150 mg/l. The Unit 3 cooling tower[6] is operated at seven cycles of concentration and its blowdown is sent to the cooling tower at Unit 4 for reuse.

Another area of sensitivity in the cooling system is the metallurgy of the cooling tower hardware. Packing hangers, bolts, etc. are stainless steel and, as such, are susceptible to stress-corrosion cracking at circulating water chloride (Cl) concentrations in excess of 1,000 mg/l. This is not a problem with San Juan River water, but it would be a concern with high-chloride produced water (if it were fed to the cooling towers untreated).

Absorbers

SO_2 is removed from the flue gas in the limestone-based absorbers and converted to gypsum ($CaSO_4 \cdot 2H_2O$). Water is used to slurry and convey limestone to the absorbers (from the

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[5] Average daily chemistry (2002) for the San Juan River provided by SJGS.

[6] The Unit 3 cooling tower is a hybrid design that carries 70 percent of the heat load in an air-cooled dry section and 30 percent in a wet section.

limestone preparation area) and compensate for water lost to the flue gas (by way of evaporation).

Make-up for the absorbers is satisfied with recycled water from the Process Wastewater Ponds and water from the San Juan River. Gypsum sludge is dewatered and the filtrate is recycled back to limestone preparation. A portion of the filtrate – Purge Water – is disposed of to the evaporation ponds. Purge Water is bled from the absorbers to control chloride levels to less than 5,000 mg/l to minimize internal corrosion. Most of the chloride entering the absorbers is organically bound in the fuel, and after combustion, it is released as HCl (hydrochloric acid) in the flue gas. As the HCl is scrubbed in the absorbers, the chloride concentration rises[7].

A secondary concern for the absorbers is magnesium (Mg). Magnesium sulfate ($MgSO_4$) is very soluble and, if magnesium levels are elevated in the absorbers, converted SO_2 would remain in the soluble sulfate form. Purge Water also helps to avoid elevated concentrations of $MgSO_4$. In a recent chemical analysis (data presented later), the Mg concentration in one of the absorbers was 4,200 mg/l [8] (this sample was taken during normal operating conditions). There are currently no operating standards for Mg, however, it is monitored closely by the plant.

For other constituents, the absorbers operate at roughly eight cycles of concentration (based on the water balance around the absorbers). The flow from the Process Wastewater Pond could be increased to the scrubbers (with commensurately less freshwater) if one of the BCs was down for maintenance and the chloride concentration in the absorbers was within limits. Also, if high-chloride produced water is fed to the absorbers untreated, it could exacerbate corrosion and/or require an increase in the Purge Water rate.

Ash System

The ash system requires water for sluicing bottom ash and wetting fly ash. San Juan River water is used for this service. The sluice system educts bottom ash from collection bins under the furnace. Sluice water is also used to seal the ash bins beneath the furnace and to wash the ash from the bin walls. The sluiced ash is sent to decanters and clarifiers where the ash is allowed to separate and settle. After clarification, the sluice water is returned for further service. Bottom ash water occasionally overflows from one of the two sluice system sumps into the plant drain system.

Sulfuric acid is added to the sluice water to maintain a pH of 7 to 9 (otherwise it rises to over 11 and causes significant scaling). The TDS of the sluice water system is 3 to 6 times[9] (1,000 to 2,000 mg/l) that of fresh water as a result of acid addition (for pH control) and evaporative losses

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[7] 85 to 90 percent of the chloride entering the absorbers comes from scrubbed HCl and this is equivalent to 6.6 tons of HCl per day.

[8] 90 to 95 percent of the Mg in the absorbers comes from the limestone. The limestone used at SJGS is dolomitic and is comprised of 95% $CaCO_3$ (limestone) and 2.5% $MgCaCO_3$.

[9] This is based on anecdotal information provided by SJGS plant staff. There is no control limit for TDS – 1,000 to 2,000 mg/l appears to be the operating level for the system as it is operated.

in the furnace bins and ash clarifiers. Sluice system corrosion is monitored and inhibitors are added to minimize corrosion and scale formation. For the purpose of this analysis, sluice water TDS should be maintained at or less than 2,000 mg/l to minimize corrosion in sluicing equipment (uncoated return piping, sluice pumps, seal water piping, etc).

Water is also used to wet fly ash as it unloaded into hauling trucks. This is done manually, and therefore, is not implemented consistently. At times there is excessive over-spraying which flows to the plant drain system.

Bottom ash and fly ash water uncontrolled releases eventually reach the Process Wastewater Ponds. High-TDS produced water used in the ash system could contaminate water to be recycled to other plant areas that are sensitive to high-salt levels.

Boiler Feedwater

Boiler feedwater is supplied to the plant by make-up demineralizers originally designed to treat water from the San Juan River. After reconfiguring the plant to zero liquid discharge, the demineralizers now receive high-quality distillate from the BCs (fed by the Process Wastewater Ponds). Regeneration frequency has been dramatically reduced because the distillate TDS is <10 mg/l (typically around 1 to 2 mg/l).

The BCs “pass” trace levels of boron in the form of boric acid to the distillate. This has been found to cause deposition problems on steam turbine blades even though distillate is further treated with the plant demineralizers. Boron levels should be less than 1 mg/l in the feedwater to the BCs (current levels) to minimize passage to the distillate. Produced water is a concern because it consistently has high levels of boron.

Summary

Compared to San Juan River water, produced water has very high levels of TDS and chloride, high levels of boron and moderate levels of silica. Table 4-2 summarizes water quality constraints for the process areas discussed above. These criteria are general and are meant to be guidelines for assessing produced water use at SJGS. The constraints are used as guidelines in the remainder of this section to evaluate untreated and treated produced water as a supplemental water supply at SJGS.

Table 4-2
Summary of Water Quality Constraints by Process Area, San Juan Generating Station

Process Area	Water Quality Constraint		Notes
Cooling Towers	Ca	1,600 mg/l_{CaCO3}	Circulating water
	SiO_2	150 mg/l	Circulating water
	Cl	1,000 mg/l	Circulating water
Absorbers	Cl	5,000 mg/l	Purge water
Ash System	TDS	2,000 mg/l	Sluice water TDS after pH adjustment
Brine Concentrators	B	<1 mg/l	BC feedwater (to prevent boron carryover)
	Cl	9,000 mg/l	BC recirculation water (Footnote 11)

4.2.3 Recycled Wastewater

SJGS collects the following wastewater streams for reuse (refer back to Figure 4-1):

- Cooling tower blowdown[10] from Units 1, 2 and 4
- Boiler blowdown (four units)
- Plant drains – mostly service water used for housekeeping and maintenance
- Spent regenerant (intermittent flow) from the boiler feedwater demineralizers
- Ash system – overflow from the bottom ash system sumps and spillage from excessive spraying in fly ash unloading area.
- Coal pile run off (occasional flow during the rainy months)

The above streams are sent to the Process Wastewater Ponds for recycle to the absorbers and the BCs. Recycled water constitutes about 40 percent of the absorber water demand. The BCs treat the remainder of the water from the Process Wastewater Ponds. BC distillate is sent to the demineralizers to be further treated for boiler feedwater. Excess distillate is sent to the cooling towers as supplemental make-up. BC brine is sent to final disposal in the evaporation ponds.

A limit of 9,000 mg/l of chloride[11] has been established for BC recirculation water. The wetted stainless steel (316L) components of the BCs experience corrosion above this limit in the form of

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[10] Cooling tower blowdown from Unit 3 is sent to the Unit 4 cooling tower. The Unit 3 cooling tower is less efficient (thermally) and routinely operates at higher circulating water temperatures. Therefore, it is more prone to certain types of scale formation and operates at a lower cycles of concentration (seven rather than ten). At seven cycles of concentration, its blowdown was considered recyclable (at commissioning) and has always been fed to the Unit 4 cooling tower.

pitting. Untreated produced water with high levels of chloride could cause a problem for the BCs.

4.2.4 Final Disposal of Wastewater at SJGS

Final disposal of wastewater at SJGS is to the evaporation ponds. The evaporation ponds consist of three 25-acre cells for a total of 75 acres. The evaporation ponds receive wastewater that cannot be recycled or treated for reuse. Of the 1,530 gpm of wastewater that is generated at SJGS, only 110 gpm (BC brine and absorber purge water) is considered unusable and disposed of in the evaporation ponds.

Plant staff have determined that every acre of pond evaporates the equivalent of 2 gpm of continuous wastewater inflow. Refer to Table 4-3 for a summary of wastewater streams and their volume requirement in the evaporation ponds.

Table 4-3
Wastewater to Evaporation Ponds, San Juan Generating Station

Wastewater Stream	Flow	Reserve Volume
Brine Concentrator Brine (1)	10 to 20 gpm	5 to 10 acres
Absorber Purge Water	100 gpm	50 acres
Boiler Cleanings (2)	Occasional	15 acres
Total		70 to 75 acres
Excess Capacity		0 to 5 acres

Notes:

As a safety margin, the plant assumes a BC brine rate of 10 to 20 gpm to the evaporation ponds.

This volume is reserved for occasional boiler cleanings.

Absorber Purge Water requires 66 percent of the available evaporation pond capacity, i.e. 50 of 75 acres. Because of this, Purge Water is assessed along with produced water when evaluating treatment alternatives (later in this section). Freeing up 50 acres of evaporation ponds would make that volume available for waste streams generated by produced water treatment.

4.3 Produced Water Resources in the Study Area

Three areas of produced water – Close-in, Tri-City and Fairway – are presented in Section 3, *Infrastructure Availability and Transportation Requirements*, Figure 4-4. Collection would be accomplished by gathering produced water from the Tri-City and Fairway areas using the Hart

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[11] This operating constraint was established by the brine concentrator manufacturer, Ionics-RCC. A higher grade of stainless steel (316 LM, 5+% Mo) would be required to operate at higher chloride concentrations for internal circulating water at a design pH of 4.0 to 5.0.

Canyon Line and CO_2 Gas Line, respectively. A new Collection Center would be built in Bloomfield to store and pre-treat the water (oil removal). A 28.5-mile pipeline originating at the Collection Center would be installed to convey the water to the plant. Close-in water from the Kirtland area would be filtered and injected directly into the pipeline just prior to delivery at SJGS. Refer to Figure 4-2 for a schematic of produced water sources, gathering and conveyance.

4.3.1 Produced Water Chemistry and Volume

The produced water chemistry in Table 4-4 is for a number of sources in the Study Area. Table 4-4 presents more chemistry data than found in Section 2, *Produced Water Assessment*, Tables 2-4 and 2-5 include additional Close-in sources and coal bed methane (CBM) wells in the Fairway[12]. The table includes chemistries and flow information for:

- Three Close-in CBM wells – Salty Dog 2/3, Turk's Toast and Taber Locke
- BHP Billiton mine water (primary source of coal for SJGS) – two samples (similar to CBM water) of like concentration were averaged
- Prax Air – cooling tower blowdown from a nitrogen plant in Kirtland
- Tri-City – average values of 30 samples
- Fairway – average values of three CBM well samples

Also an effort was made to calculate maximum probable concentrations of heavy metals by using PQL[13] values (practical quantitation level) as the minimum non-detectable values. This type of analysis is meaningful when conducting blend calculations for constituents that are near their detection levels, because PQLs are used rather than a zero value for a non-detectable concentration. One aspect of this type of analysis, is that it provides higher values for trace-level constituents. Table 4-4 also includes metals concentrations.

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[12] Fairway chemistry was within the reported ranges for the McGrath SWD. McGrath receives water from a variety of sources including the Fairway area.

[13] The practical quantitation level is the minimum concentration value a laboratory is willing to report with confidence for a specific analyte. Concentrations less than their PQL are considered non-detectable.

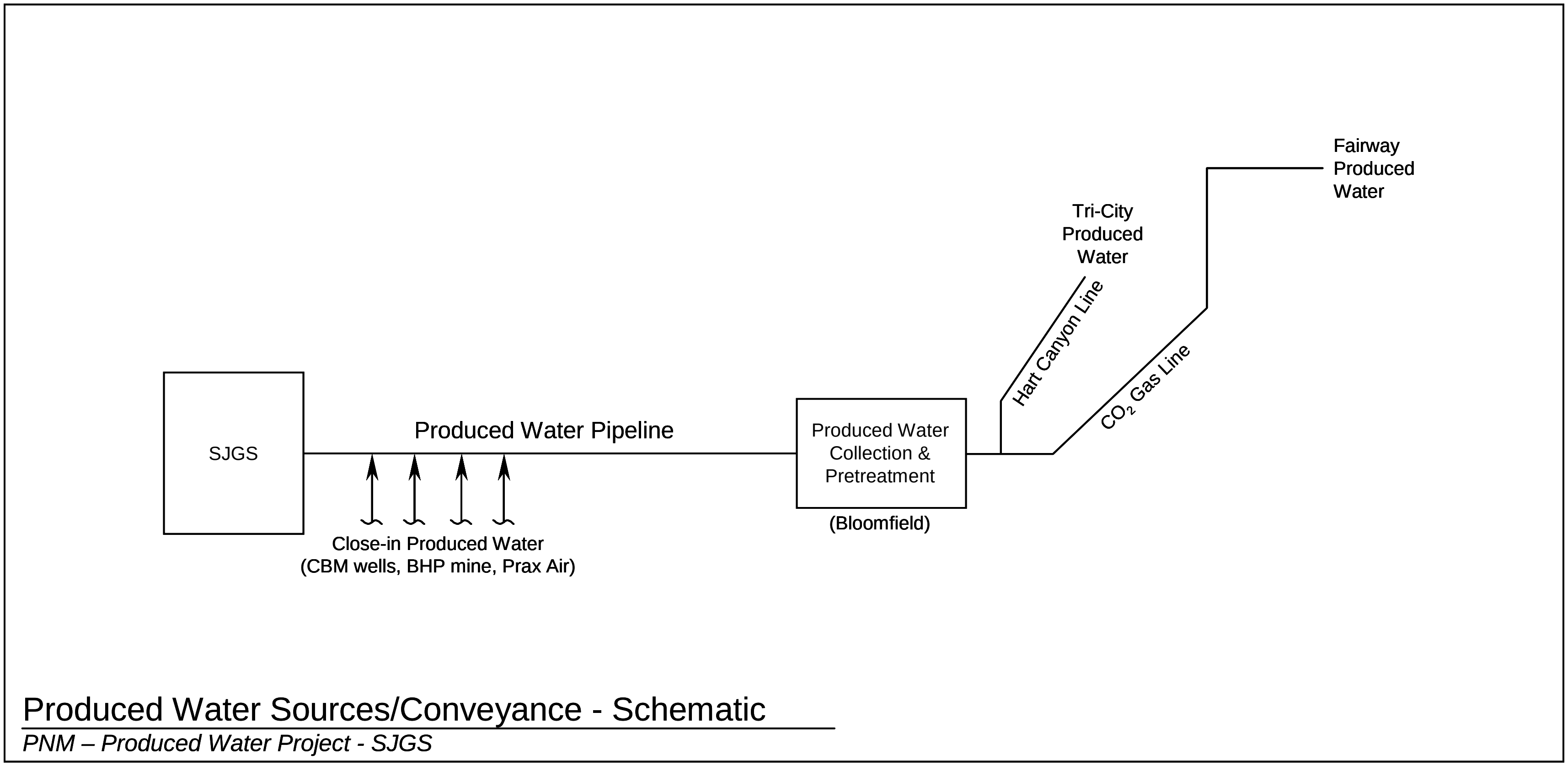

Figure 4-2
Schematic of Produced Water Sources/Conveyance, San Juan Generating Station

Table 4-4 shows daily volume estimates, along with the relative contribution from each site (expressed as a percent of total). It should be stressed that these are volume estimates and are highly dependent on the participation of individual oil and gas producers.

4.3.2 SJGS Reuse Opportunities for Untreated Produced Water

The following analysis shows how much untreated produced water could be used at SJGS, while still meeting all of the water quality constraints outlined previously. Refer to Table 4-4 for an estimate of produced water blend chemistry (all sources delivered to SJGS). Two scenarios were developed to evaluate this concept – one using the cooling towers as receivers of untreated produced water, and the other, the SO_2 absorbers. These areas of the plant were chosen because they are the largest users of water, and therefore, can theoretically accept relatively large quantities of saline produced water before their water quality limits are affected.

No assessment was done for the ash system, because the TDS of produced water chemistry, which exceeds 13,600 mg/l, is significantly greater than the 2,000 mg/l TDS constraint of the ash system. Also the water requirement for the ash system is only 100 gpm.

Produced water delivered to SJGS would contain estimated concentrations for target constituents as follows:

Target Constituent	Delivered Produced Water (1)
Ca	79.0 mg/l
SiO2	18.5 mg/l
Cl	5,043 mg/l
TDS	13,670 mg/l
B	2.51 mg/l

Note:

Extracted from Table 4-4.

The water balance shown in Figure 4-1 was used to develop a flow- and mass-calculation spreadsheet to evaluate produced water addition to the cooling towers and absorbers for a variety of produced water flow rates. Key streams in the spreadsheet could be varied, such as brine flow rate from the BCs and purge water from the absorbers, to keep the processes within their operating constraints. As can be seen in Table 4-5, all the water quality limits set for target constituents could be met, but only if certain wastewater stream flows are increased. Both scenarios are discussed next.

Table 4-4
Produced Water Chemistry – All Sources, San Juan Generating Station

Produced Water Chemistry - All Sources (page 1 of 2)

PNM - Produced Water Project - SJGS

		Salty Dog 2/3	Turk's Toast	Taber Locke	BHP Mine Water	Prax Air Blowdown	Tri-City (McGrath)	Fairway	All Sources
Flow Rate	BPD	5,000	2,500	2,200	1,700	300	10,000	20,000	41,700
	gpm	146	73	64	50	9	292	583	1,216
Flow Fraction		11.99%	6.00%	5.28%	4.08%	0.72%	23.98%	47.96%	100.00%
Na (1)	mg/l	9,563	2,119	6,848	2,936	364	4,201	3,620	4,501
K	mg/l	149	6.45	25.0	18.8	16.8	177	26.5	75.7
Calc'd NH_4	mg/l	12.4	2.16	121	56.7	3.72	16.3	16.1	21.9
Ca	mg/l	128	6.27	66.6	40.3	693	143	31.0	75.1
Mg	mg/l	87.4	4.34	32.1	41.0	105	34.1	15.1	30.3
Ba	mg/l	20.8	1.86	13.6	1.10	0.94	3.08	25.1	16.2
Sr	mg/l	20.6	1.73	18.3	3.61	9.36	19.4	14.6	15.4
Dissolved Fe	mg/l	0.84	ND	ND	ND	3.50	33.1	4.87	10.4
Cu	mg/l	ND	ND	ND	ND	0.200	ND	ND	0.131
Zn	mg/l	0.298	ND	ND	ND	ND	0.230	ND	0.180
As	mg/l	ND	ND	ND	ND	ND	ND	ND	<0.180
Cr	mg/l	ND	0.005	ND	0.009	0.090	ND	0.003	0.004
Pb	mg/l	0.036	ND	ND	ND	1.550	ND	0.040	0.037
Se	mg/l	0.017	ND	ND	ND	ND	ND	0.015	0.013
Hg	mg/l	ND	ND	ND	ND	ND	ND	ND	<0.002
Ag	mg/l	NA	ND	ND	ND	0.250	NA	ND	0.022
U	mg/l	NA	ND	ND	ND	ND	NA	ND	<0.003
TC	mg/l_{CaCO3}	21,697	4,649	15,557	6,827	3,004	9,970	8,117	10,285

Produced Water Chemistry - All Sources (page 2 of 2)

PNM - Produced Water Project - SJGS

		Salty Dog 2/3	Turk's Toast	Taber Locke	BHP Mine Water	Prax Air Blowdown	Tri-City (McGrath)	Fairway (4)	All Sources
HCO_3	mg/l	1,440	1,952	1,050	853	139	764	6,377	3,622
CO_3	mg/l	5.51	34.2	0.68	5.04	0.61	0.64	21.8	13.6
Cl (1)	mg/l	14,518	2,089	10,418	3,536	352	6,219	2,018	5,021
Br	mg/l	15.6	2.74	3.17	8.72	NA	14.5	18.9	15.1
F	mg/l	ND	2.30	1.47	1.04	NA	ND	0.74	0.61
NO_3	mg/l	2.55	ND	ND	0.32	NA	ND	3.49	1.99
NO_2	mg/l	ND	ND	ND	ND	NA	ND	ND	ND
SO_4	mg/l	24.9	37.4	ND	1,082	2,300	544	4.32	198
TA	mg/l_{CaCO3}	21,697	4,649	15,557	6,827	3,004	9,970	8,130	10,291
SiO_2	mg/l	9.7	12.2	32.5	15.9	1.82	18.5	21.44	19.0
Total Fe	mg/l	0.78	4.05	9.08	4.08	NA	41.3	4.58	13.1
Total Alkalinity	mg/l_{CaCO3}	1,180	1,910	1,050	790	119	697	5,398	3,101
Total NH_3	mg/l_N	10.6	1.90	94.0	45.2	3.20	12.8	13.25	17.6
B	mg/l_B	2.87	1.60	2.40	0.81	29.0	2.05	2.31	2.41
O-PO_4	mg/l_P	ND	ND	ND	ND	ND	ND	ND	ND
Total Sulfide	mg/l_S	ND	17.0	NA	NA	NA	ND	4.45	NC
pH		8.23	8.82	7.40	8.37	8.27	7.05	8.00	7.83
EC	µS/cm	40,300	9,160	29,900	13,200	NA	19,880	14,556	19,246
TDS (Calc'd)	mg/l	26,010	6,300	18,660	8,610	4,160	12,210	12,236	13,658
TSS	mg/l	42	16	18	814	NA	108	26	79
TPH	mg/l	ND	17	2.3	75	NA	163	71	77

Notes…..

1. Na and Cl values adjusted (as required) to achieve ionic balance.
2. NA = not analyzed, ND = not detectable, NC = not calculable.
3. PQL = practical quantitation limit.
4. Fairway TPH is an average of three sources - two sources had TPH concentrations typical of CBM production, i.e. at or less than 5 mg/l. One source (or the sample) was likely contaminated.

Table 4-5
Operating Adjustments to Meet Target Constraints, San Juan Generating Station

Mass Balance - Target Constituents		**Current Operation**	**Scenario 1 Produced H_2O to Cooling Towers**	**Scenario 2 Produced H_2O to Absorbers**
Cooling Towers	Ca	1599 mg/l_{CaCO3}	1594 mg/l_{CaCO3}	1594 mg/l_{CaCO3}
	SiO2	140 mg/l	145 mg/l	140 mg/l
	Cl	260 mg/l	1,000 mg/l	260 mg/l
Absorbers	Cl	5,000 mg/l	5,000 mg/l	5,000 mg/l
BC Feedwater	B	0.8 mg/l	1.0 mg/l	0.8 mg/l
BC Brine	Cl	9,000 mg/l	9,000 mg/l	9,000 mg/l

Key Water Balance Stream Adjustments to Meet Target Constraints

Produced Water	0 gpm	155 gpm	100 gpm
BC Brine to Evap ponds	14.8 gpm	58.1 gpm	14.7 gpm
Absorber Purge Water to Evap Ponds	100.0 gpm	171.9 gpm	200.3 gpm
Total Wastewater to Evap Ponds	114.8 gpm	230.0 gpm	215.0 gpm
Additional Wastewater to Evap Ponds	0 gpm	115.2 gpm	100.3 gpm
Net Water Savings, gpm	0 gpm	39.8 gpm	(-0.3 gpm)
Annual Plant Demand	22,400 AF	22,336 AF	22,401 AG
Annual Freshwater Savings	0 AF	64 AF	(-1 AF)

Scenario 1 – Cooling Towers

155 gpm of produced water could be added to the cooling towers without increasing blowdown (operating at the current cycles of concentration). Chloride levels in the cooling tower would rise from 260 mg/l to 1,000 mg/l. Because the chloride concentration in the blowdown to the Process Wastewater Ponds is significantly higher, the waste brine rate from the BCs would have to be increased from 14.8 gpm to 58.1 gpm to maintain the 9,000 mg/l chloride operating limit in the BCs. Likewise, the absorber purge water rate would have to be increased from 100 gpm to 171.9 gpm to maintain the 5,000 mg/l chloride limit. Under these conditions, chemistry constraints would be met for all process systems. However, the total-plant wastewater flow to the evaporation ponds would increase from 114.8 to 230.0 gpm. This would require the addition

of at least 58 acres[14] (equivalent to 116 gpm) of new evaporation ponds to receive the additional wastewater.

The amount of produced water could be increased above 155 gpm, but there would be a commensurate increase in cooling tower blowdown to maintain a chloride content of 1,000 mg/l. The cooling tower blowdown would double from 1,000 to 2,000 gpm, if 356 gpm of produced water is added to the cooling tower stream (maintaining 1,000 mg/l of chlorides in the cooling tower blowdown). Note that produced water has a relatively high chloride concentration, so increased amounts to the cooling tower increase the blowdown. The blowdown would go to the Process Wastewater Ponds where a balance is maintained between the BCs and the absorbers. The BC receives 800 gpm (BC capacity) and the balance is sent to the absorbers. Therefore as the blowdown increases, the recycle flow of wastewater increases to the absorbers. With more cooling tower blowdown (at 1,000 mg/l of chlorides), Process Wastewater Pond water would be higher in chlorides. This in turn would require an increase in the purge water flow to maintain absorber chlorides. If the produced water rate to the cooling towers exceed 466 gpm, the Process Wastewater Ponds would generate more water than is recyclable to the absorbers and BCs. At this point, excess Process Wastewater Pond water would be sent directly to the evaporation ponds.

Scenario 2 - Absorbers

The absorbers have an operating limit of 5,000 mg/l of chlorides. Coincidentally, the produced water blend has a concentration of 5,040 mg/l of chlorides. If produced water were added to the absorbers it would create a wastewater stream of slightly larger magnitude. For example, if 100 gpm of produced water were added to the absorbers, purge water would increase by 100.3 gpm from 100 gpm to 200.3 gpm. Given this water chemistry, there are no direct-addition scenarios that are feasible for untreated produced water.

4.3.3 Summary

The use of untreated produced water is not practical at SJGS. Small amounts of high-TDS produced water (a fraction of what is available on a continuous basis) generate excess wastewater that cannot be handled by the plant.

The next section identifies technologies capable of treating produced water, develops treatment configurations, and preliminarily assesses the economics of each treatment configuration.

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[14] Logistically, SJGS can install an additional 20 to 30 acres of evaporation ponds on relatively flat terrain within the plant proper. Additional ponds would have to be installed at the edge of plant property by Highway 64 (about 3 to 4 miles from the existing ponds).

4.4 Produced Water Treatment

Three sources of produced water – Close-in, Tri-City and Fairway – along with water from the BHP Billiton coal mine and a small amount of industrial wastewater are evaluated for treatment. Treating absorber Purge Water[15] is also assessed, because it would free up 50 acres of evaporation pond capacity to accommodate wastewater generated by produced water treatment.

Off-the-shelf commercially-available technology is evaluated next in this section. Public Service of New Mexico (PNM) is currently looking at supplemental sources of water for SJGS, so proven technology is needed to implement any project in a timely manner. Water treating equipment at the plant is also included in the evaluation.

4.4.1 Water to be Treated

Table 4-4 presents the chemistry for produced water and for BHP Billiton coal mine water and Prax Air cooling tower blowdown. Flow assumptions (provided by the oil and gas producers) are used to calculate an estimated blend chemistry. Refer to Table 4.6 for produced water blend chemistry, Purge Water chemistry and a hypothetical blend of produced water and Purge Water (PW/PW).

If it is feasible, blending purge water with produced water produces two benefits:

- Frees up 50 of the 75 acres of evaporation ponds at the plant. 50 acres are equivalent to $8.6 million in new evaporation pond costs.
- Provides an additional 100 gpm of water that, if treatable, can be reused at the plant.

<u>Produced water sources</u> are characterized as follows:

- Relatively high TDS – 13,700 mg/l – comprised mostly of sodium bicarbonate and sodium chloride salts.
- Ammonia in the form of ammonium (NH_4^{+1}) is moderately high.
- Low level of calcium and magnesium hardness[16] at 325 mg/l_{CaCO3}.
- Iron concentration is typical for produced water[17] and can range as high as 20 to 30 mg/l at times.
- Heavy metals concentrations are low - near the detection limit for most constituents and non-detectable for the remaining.

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[15] Purge Water generates a continuous flow of 100 gpm and uses 50 of the 75 acres of evaporation ponds at SJGS.

[16] Calcium and magnesium hardness is calculated as follows: Ca-Mg Hardness, mg/l_{CaCO3} = Ca, mg/l_{ion} x 2.50 + Mg, mg/l_{ion} x 4.12.

[17] Piping and tankage in oil and gas production are usually bare carbon steel, so iron levels from corrosion are typically high.

- Silica is relatively low. Some produced waters in California have SiO_2 concentrations as high as 180 mg/l.
- Boron levels are high – consistent with oil and gas production.
- pH is slightly above neutral – probably lower (6.8 to 7.2) before release at the wellhead.

Total suspended solids (TSS) and total petroleum hydrocarbons (TPH) presented in Tables 4-4 and 4-6 are not representative of produced water "just out of the ground". Samples taken at McGrath SWD (salt water disposal injection facility in the Tri-City area) were grabbed prior to injection, i.e. pretreated for oil separation and filtration. CBM produced water from Close-in and Fairway production was taken from storage prior to pretreatment[18] (unfiltered). Also, the McGrath SWD receives conventional produced water with high levels of TPH. Close-in produced water is consistently low in TPH but high in TSS (mostly coal fines). Of note is the Fairway TPH concentration in Table 4-4. It is an average of three sources - two sources had TPH concentrations typical of CBM production, i.e. at or less than 5 mg/l. One source (or the sample) was likely contaminated.

Purge water[19] is characterized as follows:

- Higher TDS – 20,500 mg/l – than the produced water blend and mostly comprised of sodium chloride and magnesium sulfate salts.
- Ammonium is twice that of produced water.
- Much higher levels of calcium compared to produced water.
- Significantly high levels of magnesium at 4,160 mg/l – the magnesium comes from the dolomitic limestone used in the SO_2 absorbers.
- Strontium levels are significantly high at 207 mg/l.
- Very low levels of iron – the absorber vessels are lined to minimize corrosion.
- Relatively low concentrations of heavy metals (except for selenium), although they are consistently higher than those for produced water.
- Selenium levels are very high at 6.18 mg/l.
- Fluoride and nitrate concentrations are very high – both likely originate from the coal. Also, NOx is likely being scrubbed as well.
- Sulfate levels are very high at 18,000 mg/l as a result of scrubbing SO_2 from flue gas.

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[18] CBM water pretreatment prior to produced water injection consists of filtration to remove coal fines. Unlike conventionally produced water, CBM does not contain floatable hydrocarbons, and thus does not require oil separation.

[19] It is assumed that the significantly high concentrations of ammonium, strontium, selenium, fluoride and boron in the absorber Purge Water derive from the plant coal. The nitrate concentration is from scrubbed NO_2 in the flue gas.

Produced Water & Purge Water Chemistry

PNM - Produced Water Project - SJGS

		Produced Water	Purge Water	Blend	PQL (3)
Flow Rate	BPD	41,700	3,429	45,129	
	gpm	1,216	100	1,316	
Flow Fraction		92.40%	7.60%	100.00%	
Na (1)	mg/l	4,491	2,785	4,362	
K	mg/l	76.4	178	84.1	
Calc'd NH_4	mg/l	19.7	31.4	20.6	
Ca	mg/l	79.0	493	110	
Mg	mg/l	30.5	4,160	344	
Ba	mg/l	13.8	0.374	12.8	
Sr	mg/l	13.6	207	28.3	
Dissolved Fe	mg/l	11.3	<0.01	10.5	
Cu	mg/l	0.131	0.134	0.131	0.13
Zn	mg/l	0.180	0.667	0.217	0.14
As	mg/l	<0.180	0.042	0.020	0.018
Cr	mg/l	0.004	0.035	0.006	0.003
Pb	mg/l	0.037	<0.005	0.034	0.005
Se	mg/l	0.013	6.180	0.482	0.011
Hg	mg/l	<0.002	<0.002	<0.002	0.002
Ag	mg/l	0.022	<0.02	0.022	0.02
U	mg/l	<0.003	0.078	0.009	0.003
TC	mg/l_{CaCO3}	10,268	24,966	11,384	
HCO_3	mg/l	3,619	87.3	3,351	
CO_3	mg/l	14.3	0.01	13.2	
Cl (1)	mg/l	5,043	3,976	4,962	
Br	mg/l	14.3	12.6	14.1	
F	mg/l	0.93	120	10.0	
NO_3	mg/l	1.99	305	25.0	
NO_2	mg/l	ND	ND	ND	
SO_4	mg/l	198	18,000	1,550	
TA	mg/l_{CaCO3}	10,320	24,966	11,432	
SiO_2	mg/l	18.5	32.5	19.6	
Total Fe	mg/l	14.0	3.01	13.2	
Total Alkalinity	mg/l_{CaCO3}	3,105	110	2,877	
Total NH_3	mg/l_N	15.8	27.0	16.6	
B	mg/l_B	2.51	129	12.1	
O-PO_4	mg/l_P	ND	ND	ND	
Total Sulfide	mg/l_S	NC	NA	NC	
pH		7.84	7.86	7.84	
EC	µS/cm	18,931	24,050	19,320	
TDS (Calc'd)	mg/l	13,666	20,460	14,182	
TSS	mg/l	83	269	97	
TPH	mg/l	99	64	96	

Notes…..

1. Na and Cl values adjusted (as required) to achieve ionic balance.
2. NA = not analyzed, ND = not detectable, NC = not calculable.
3. PQL = practical quantitation limit.

- Silica is moderately low.
- Boron levels are very high at 129 mg/l.
- pH was slightly above neutral in this sample (it usually is controlled between 5.0 and 6.0)[20].

Since there is twelve times as much produced water (92.4 percent of total) as Purge Water (7.6 percent), the impact of Purge Water on PW/PW blend chemistry is generally minimal. Refer again to Table 4-6. Many constituent concentrations hardly change, e.g. sodium, potassium, ammonium, chloride, alkalinity. TDS of the hypothetical PW/PW blend is only 4 percent higher than that of produced water. Of note however, are the concentrations of magnesium and sulfate, which increase dramatically – 11 times and 8 times those of produced water, respectively. Selenium also increases in the PW/PW blend from 0.013 mg/l to 0.482 mg/l.

<u>Volatile organic compounds</u> (VOCs) are present at low levels in conventional produced water. VOCs in produced water commonly consist of benzene, toluene, ethyl benzene and xylene, which are known collectively as BTEX. CBM water has very low levels of BTEX relative to conventional water – 0.1 to 0.2 mg/l BTX in CBM water versus 10 to 20 mg/l in conventional water. VOCs and VOC removal are discussed in more detail later in this section of the report.

The chemistry just discussed is used to evaluate commercially available technologies next. Water collected from Close-in, Tri-City and Fairway production are assessed as well as the PW/PW blend.

4.4.2 Commercially Available Technology

As discussed previously, small amounts of high-TDS produced water fed to the cooling towers or absorbers would generate excessive volumes of wastewater. Produced water as well as PW must be treated before it can be reused in the plant. Table 4-6 compares operating constraints for the cooling towers, absorbers and ash system against San Juan River water (the fresh water supply to the plant), produced water and the PW/PW blend. Chloride and TDS levels in produced water and the PW/PW blend clearly exceed or approach process operating constraints <u>before the concentrating effect in each of the systems</u>.

Specifically, chloride and TDS concentrations must be significantly reduced before produced water or the PW/PW blend can be used as supplemental feedwater for SJGS. From a water treating perspective, there are several off-the-shelf technologies that can be used to <u>economically</u> lower chloride and TDS concentrations:

- Reverse osmosis (RO)
- Evaporative processes (such as the brine concentrators at SJGS)
- Process combinations of the above

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[20] There was likely a release of CO_2 when the sample was taken, and this would have allowed the pH to rise.

Table 4-6
Process Area Constraints versus Various Water Sources

Process Area	Constituent	System Operating Constraint	San Juan River (1)	Produced Water (2)	PW/PW (2)
Cooling Towers	Ca, mg/l$_{CaCO3}$	1,600	135	200	275
	SiO_2, mg/l	150	12	20	20
	Cl, mg/l	1,000	22	5,040	4,960
Absorber	Cl, mg/l	5,000	22	5,040	4,960
Ash System	TDS, mg/l	2,000	360	18,930	19,320

Notes:

SJGS fresh water supply.

Close-in, Tri-City and Fairway produced water blend and PW/PW data from Table 4-4. Concentrations are rounded for simplicity.

A significant amount of pilot testing (government and private funding) of produced water has been done in the recent past to evaluate proven and developmental technologies. Many of the major oil companies in California have either tested or considered treating produced water. A produced water treatment plant is in final design in San Ardo, California (150 miles south of San Francisco and 35 miles inland from the coast). Likewise, similar studies have been conducted in Alberta, Canada. The focus of most of the testing (and investment) has been on membrane configurations – RO with a variety of pretreatment options, and in some cases, post treatment of RO reject (concentrated waste stream). RO has proven to be a robust process for this service as long as pretreatment is applied correctly and rigorously maintained (discussed more later).

There are other commercial technologies – ion exchange and electrodialysis – capable of reducing TDS, but they are untested (pilot testing or otherwise) in this service, and therefore, are not considered in this analysis. Also, at these levels of TDS, ion exchange would be very costly to operate and would generate large volumes of wastewater.

Reverse osmosis is discussed next part of this section. Brine concentration is discussed in the context of SJGS. The plant has two idled brine concentrators (typical of those found in power plants throughout the Southwest) that are capable of treating produced water.

4.4.3 Reverse Osmosis

Osmosis occurs naturally in membrane systems. A simple example is the passage of water through the root cells of a plant. Water in soil has a lower concentration of salts than in root cells. The cell wall (membrane) allows water to enter the root cell in an attempt to equalize the concentration of salts on each side of the membrane.

Experimentally, if equal volumes of water – one saltier than the other – are placed in a u-shaped tube with a permeable membrane separating them, osmosis takes place. Refer to Figure 4-3. Water from the side with the lower salt concentration diffuses through the membrane to the saltier side. At equilibrium, the salt concentrations on both sides of the membrane are equal and the difference in water levels is sustained by osmotic pressure.

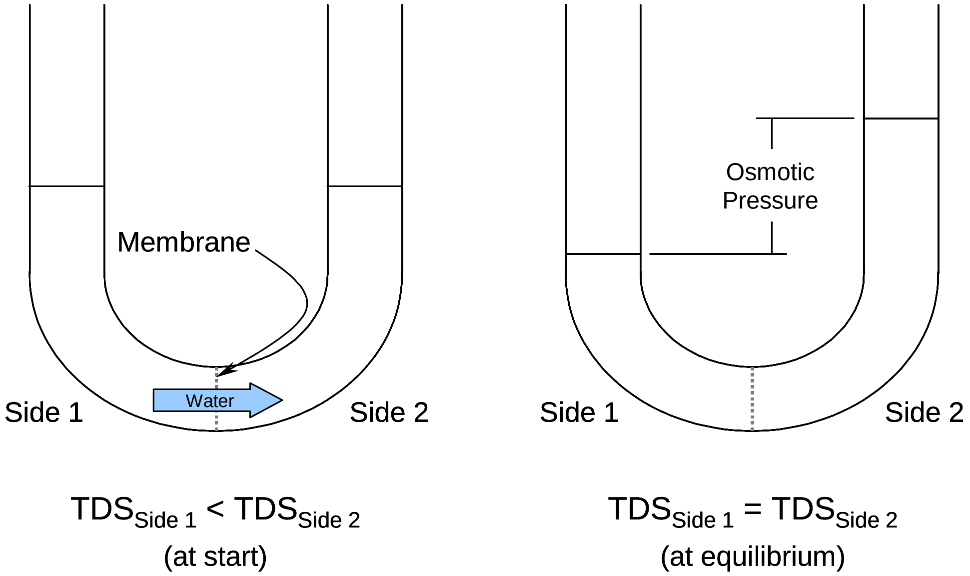

Figure 4-3
Illustration of Osmotic Pressure

Early research in water purification revealed that if pressure is applied to the side of a membrane with the higher salt concentration, water is forced back through the membrane leaving the salts behind. This phenomenon was coined reverse osmosis. In the past 50 years, membrane design has advanced to a high level of sophistication. RO has become a very common process and is used extensively throughout the world to desalinate seawater, purify water for industrial use, and more typical in the past few years, recycle wastewater.

Membrane Configurations

There are two common membrane configurations:

- Spiral wound – the membrane assembly consists of sheets bound to a hollow core. The membrane sheets are wrapped around the core to form a cylinder. Water flows across the surface of the membrane to allow permeate (water that passes through the membrane) to flow to the collection core.
- Hollow fiber – the membrane consists of a bundle of densely-packed hair-thin hollow fibers. The fibers are embedded into an anchor cap at one end and a collection cap at the other. The permeate travels through the hollow passages in the fibers and exits at the collection end of the membrane assembly.

Spiral wound membranes are the most commonly used configuration. Hollow fiber membranes are now used mostly for non-fouling service, because they are very difficult to clean if fouled[21]. Other membrane configurations exist but are not very common, e.g. tubular membranes and plate and frame membranes.

Because spiral wound membranes are more suited to the fouling potential of produced water service, the remainder of this section concentrates on this membrane configuration.

Spiral Wound Membranes

As described previously, layers of membrane sheets are wrapped around a hollow core to form a spiral wound element. Refer to Figure 4-4. Feedwater passes through the membrane face and is channeled inside the membrane envelope to the core for permeate collection. Also, sheet geometry permits turbulent flow to occur across the face of the membrane. Scouring created by turbulence at the plane of the membrane surface makes this configuration better suited for the fouling potential of produced water.

In industrial applications, the elements are either 4 or 8 inches in diameter and typically 40 inches long. The membranes are loaded into a pressure tube – two to six to a tube depending on the size of the system. Water is fed to one end of the pressure tube, and at the other end, permeate and reject (concentrated wastewater) exit in separate lines. Pressure tubes are arrayed such that feedwater is distributed to the tubes equally. In a staged RO system, the first stage receives feedwater and the second stage receives RO reject as feedwater. If there is a third stage, it would receive second-stage reject as feedwater. Refer to Figure 4-5 for a simplified two-stage, three-tube RO system. The number of elements, pressure tubes and stages is dependent on feedwater rate, feedwater quality, recovery goals (how much permeate is desired), etc.

Membrane Types

There are three common types of membrane materials – thin-film polyamide (PA), cellulose acetate (CA) and polysulfone (PS). PA membranes are currently the most widely-used membranes. They have higher flux rates (flow rate per unit of membrane surface area), lower energy requirements (for the same feedwater TDS and flow rate) and better salt rejection than CA membranes. CA was one of the first commercial types of RO membranes, but has lost significant market share to more-efficient PA membranes. CA membranes have better oxidation resistance[22] than PA membranes.

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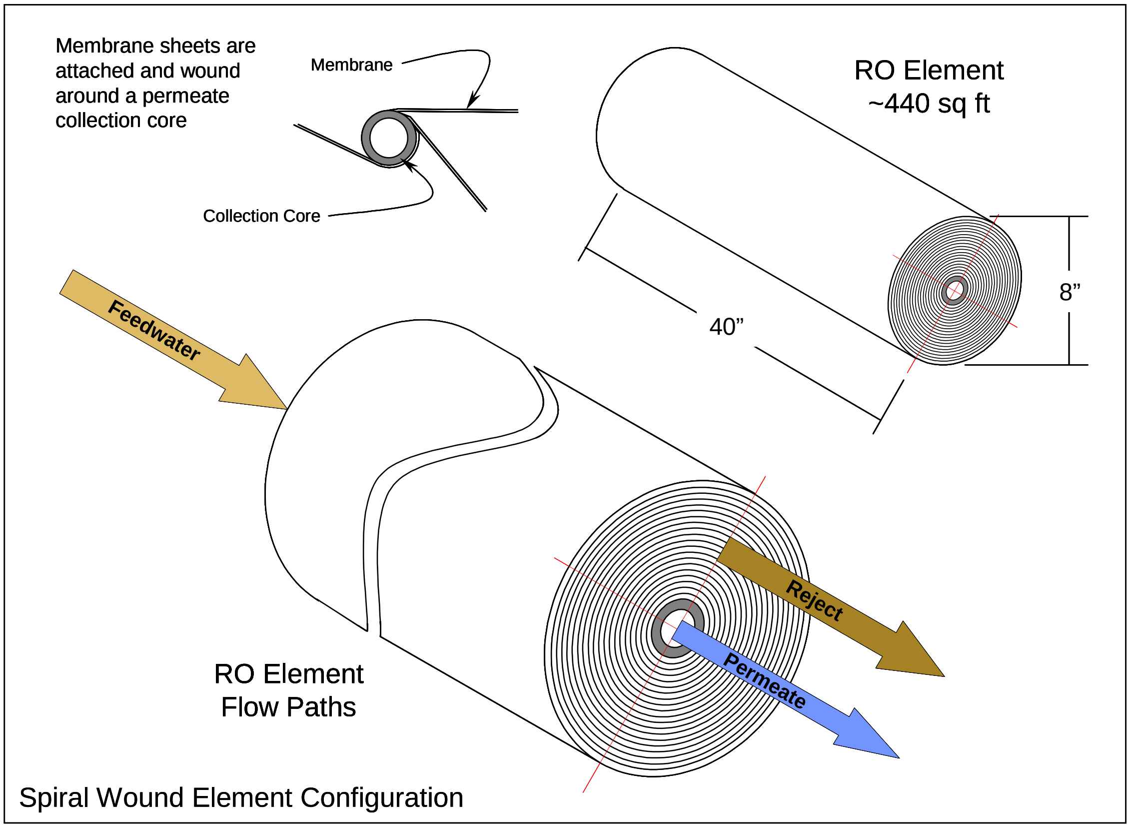
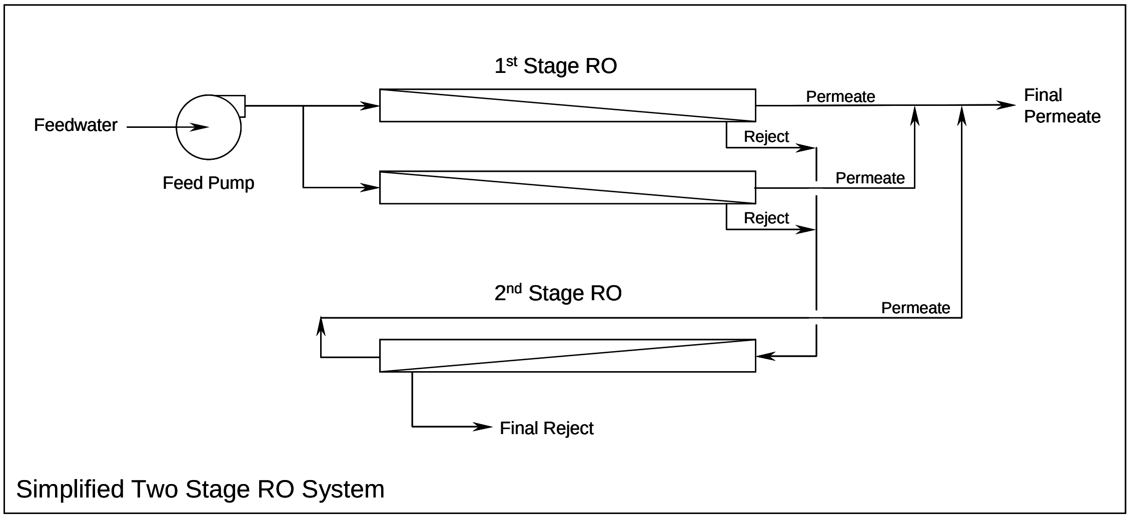
[21] The pores in the membrane surface are believed to be less than 0.001 microns (10^{-9} meters) and are easily fouled without proper pretreatment. Human hair has a diameter of 30 microns.

[22] Disinfecting agents are used to prevent bacteria from growing in RO systems. Bacteria form films which blanket and clog membrane surfaces. Common disinfecting agents (sodium hypochlorite – bleach solution) are used to control biological fouling, however they can destroy membrane functionality.



Deliverable 3/RO Element Spiral DiFilippo 7/6/2004

Figure 4-4
Spiral-Wound Element Configuration



Deliverable 3/2 Stage RO System DiFilippo 7/6/2004

Figure 4-5
Simplified Two-Stage Reverse Osmosis System

Produced Water Service

Spiral wound, thin-film polyamide RO membranes are evaluated for produced water service in this section. Produced water pilot studies have focused on this membrane because:

- The membrane exhibits better resistance to produced water foulants – mineral scale, particulate matter, oil and biological fouling.
- Permeate has lower salt passage characteristics (lower permeate TDS).
- The membranes require less energy, i.e. they operate at lower pressure.

RO fouling remains a real concern for RO systems treating produced water, making pretreatment critical. Pretreatment, which is often more complex than the RO system itself, is developed later in the section.

4.4.4 Brine Concentrator Equipment at SJGS

The brine concentrators at SJGS are typical of evaporation equipment found in power plants in the Southwest. The plant has four brine concentrators – two operating and two idled. SJGS treats and recycles much of its wastewater with two brine PS membranes. They have the best oxidizing resistance so they perform well in waters with biological fouling tendencies; however, PS membrane feedwater must be softened[23] or the membranes lose their salt rejection capability.

In modern membranes, 95 to 99 percent salt rejection is achievable (depending on membrane type and feedwater conditions). Therefore, only 1 to 5 percent of the salts in the feedwater (salt passage) pass through the membrane with the permeate concentrators – capacities of 240 gpm and 560 gpm (800 gpm total) – BC 4 and BC 5, respectively. There are also two idled brine concentrators at the plant – BC 2 and BC 3 – each has a capacity of 580 gpm (1,160 gpm total). The idled brine concentrators are evaluated for treating and reusing produced water at SJGS, since the operating BCs are fully utilized.

The brine concentrators at SJGS operate in the seeded mode, i.e. crystal formation[24] is encouraged in the concentration process to enable higher recoveries of water. Refer to Figure 4-6. Feedwater to the brine concentrators is acidified to between 4.0 and 5.0. Acidification[25] converts all the alkalinity to carbonic acid. Before the feedwater is introduced into the brine concentrator, it is preheated by hot distillate exiting the brine concentrator. A scale inhibitor[26] is

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[23] Softening in this context consists of the removal of divalent (calcium, magnesium, iron, etc.) and trivalent ions (iron, aluminum, etc.).

[24] Crystal formation as calcium sulfate ($CaSO_4$). Calcium sulfate crystals in the presence of scale inhibitors move freely in the recirculating brine and along the heat transfer surfaces. If the BC were not operated in this mode, recovery would be severely reduced to prevent scale formation.

[25] Alkalinity removal via acidification is critical because calcium carbonate tends to foul heat transfer surfaces in the brine concentrator.

[26] Scale inhibitors are proprietary products sold by specialty chemical providers.

also added to the feedwater to modify crystal growth, keep crystals in the bulk fluid and minimize scale adhesion to heat transfer surfaces. After preheating, the acidified water enters a deaerating section where dissolved gases (primarily oxygen and carbon dioxide) are released to minimize corrosion. The acidified and deaerated water is then added to the sump of the brine concentrator.

Water in the sump is pumped to the top of the brine concentrator and allowed to fall (in film-like fashion) in vertical tubes. Each tube has swirl device at the tube sheet to encourage film formation and minimize scale buildup at its entrance. As the film of water moves down the tubes, a small fraction of it evaporates as steam. The steam is collected in the top of the brine concentrator and compressed. After compression, it is admitted to the vapor space on the outside of the vertical tubes where it provides the heat source for evaporation. This type of brine concentrator is known as a vapor compression evaporator, since the compressor provides the thermal input to evaporate the recirculating brine. Brine concentrator distillate is usually of excellent quality – TDS is approximately 1 to 2 mg/l.

The in-service brine concentrators (BC 4 and BC 5) are susceptible to chloride pitting corrosion (at an operating pH of 4.0 to 5.0), because much of the metallurgy (sump, vapor space, recirculation piping, etc.) is 316L stainless steel. For this reason, the chloride concentration in the recirculating brine must be kept under 9,000 mg/l (as recommended by the manufacturer). The vertical tubes (and tube sheet) are titanium, and as such, could withstand much higher levels of chloride.

The brine concentrators are well suited for the wastewater currently being treated at the plant, i.e. low levels of chloride, and a good ratio of calcium and sulfate for seeded-mode operation. Some seeding is required depending on feed chemistry (calcium chloride is added for this purpose). The brine concentrators at SJGS recover over 98 percent of the feedwater (even with their chloride operating constraint). Recirculating brine at the plant can have a dissolved and suspended solids loading exceeding 200,000 mg/l.

The brine concentrator manufacturer was consulted several years ago to determine which of the two idled brine concentrators (BC 2 and BC 3) was best suited for refurbishment, disassembly and reassembly at another PNM power plant (relocation was never implemented). It was later determined that both could be refurbished – BC 3 would require significantly less repair than BC 2. Also, of note is the fact that these brine concentrators have 316 LM metallurgy[27] and are much more resistant to chloride pitting corrosion. If refurbished[28], the chloride operating limit for these brine concentrators would be 50,000 mg/l.

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[27] 316 LM is a high-moly content stainless steel alloy with 5+% of molybdenum.

[28] Several minor components would still require metallurgical upgrades to 316LM to operate at higher chloride levels.

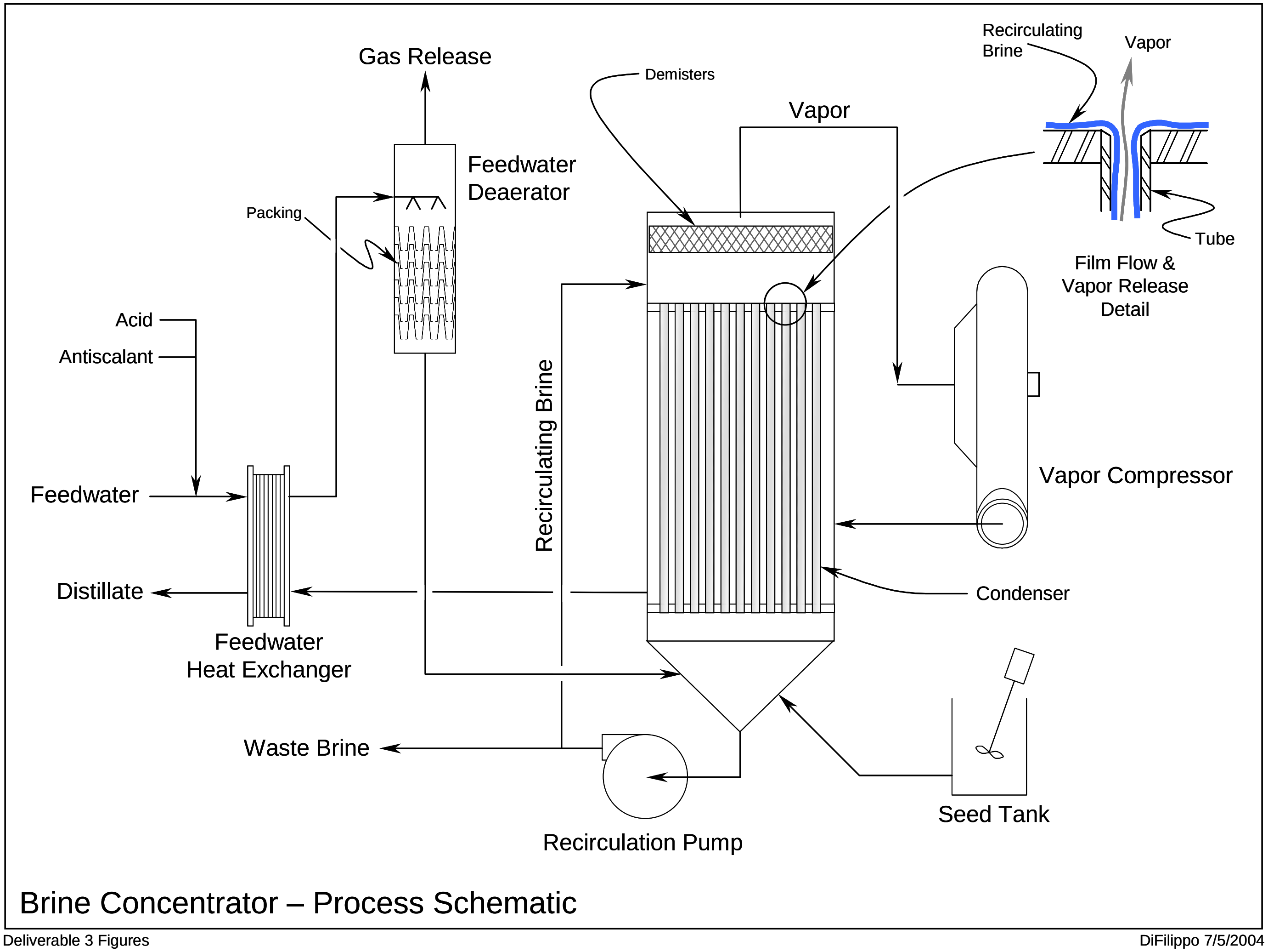

Figure 4-6
Brine Concentrator Process Schematic

4.4.5 Produced Water Treatment Alternatives

Treatment alternatives that incorporate off-the-shelf technology – reverse osmosis and brine concentration – are developed next. Alternatives specifically include spiral-wound thin-film polyamide RO and the BC equipment at SJGS (discussed previously in 4.4.3 and 4.4.4). Refer to Figure 4-7 for a schematic overview of the three basic combinations of RO and BC technologies:

- Treat with only RO.
- Treat only with BC 2 and BC 3.
- Treat with a combination of RO and BC 3 (better of the two BCs).

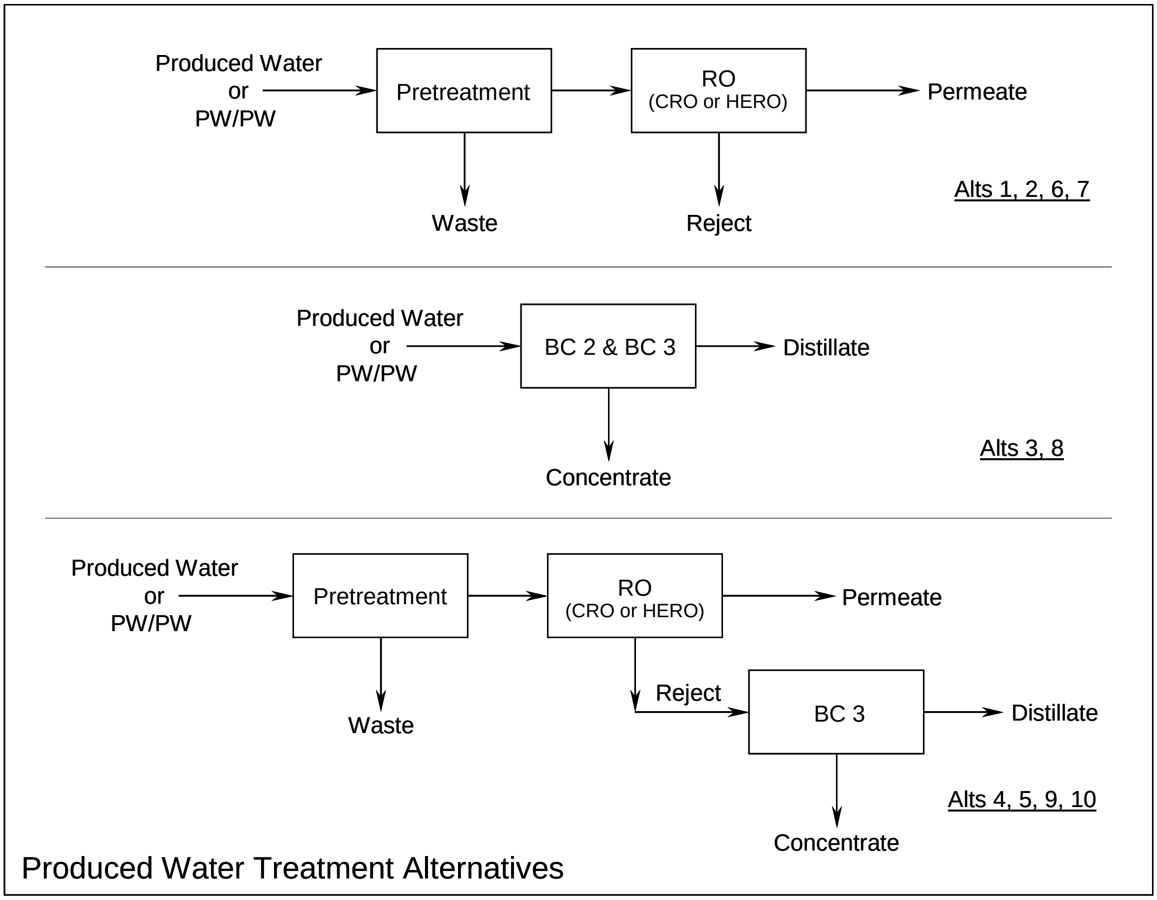

Figure 4-7
Produced Water Treatment Alternatives

This analysis evaluates treatment of produced water delivered to SJGS. Pretreatment at the Bloomfield Collection Center is discussed later in this section.

Two sub-alternatives are investigated for each RO and RO-BC alternative:

- Conventional RO (CRO) – includes standard spiral-wound PA membranes operating at low pH. This is a traditional approach to operating RO systems.

- High-efficiency RO (HERO®)[29] – includes standard spiral-wound PA membranes operating at high pH. This is a relatively new approach with inherent advantages to treating produced water.

Lastly, five treatment alternatives were evaluated for produced water and the same five for the produced water and Purge Water (PW/PW) blend. A total of 10 alternatives are evaluated as described in Table 4-7.

Table 4-7
Produced Water Treatment Alternatives Summary

	CRO	**HERO®**	**BC 2**	**BC 3**	**Produced Water Only**	**PW/PW Blend**
Alternative 1	**X**				**X**	
Alternative 2		**X**			**X**	
Alternative 3			**X**	**X**	**X**	
Alternative 4	**X**			**X**	**X**	
Alternative 5		**X**		**X**	**X**	
Alternative 6	**X**					**X**
Alternative 7		**X**				**X**
Alternative 8			**X**	**X**		**X**
Alternative 9	**X**			**X**		**X**
Alternative 10		**X**		**X**		**X**

RO Pretreatment - General

Pretreatment for RO focuses on the prevention of membrane fouling, which occurs when foreign matter blocks membrane pores. Membrane fouling concerns include:

- Mineral scale occurs when foulant constituents are concentrated in the RO element beyond their saturation levels. Crystals nucleate and attach to membrane surfaces. Examples include calcium carbonate, barium sulfate and silica[30]. Scale can be prevented either by lowering mineral constituent concentrations in RO feedwater or by operating the RO at lower recovery (to avoid over saturation). Precipitation softening is sometimes used to lower metals

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[29] Aquatech International (water treatment OEM) is the sole licensee of HERO® technology for power plant applications.

[30] At concentrations exceeding 150 mg/l and at a pH of less than 10.0, silica polymerizes and forms an amorphous (non-crystalline) deposit that is very difficult to remove.

(calcium, magnesium, etc.) and silica levels[31]. Scale inhibitors are also used to reduce scale formation potential by extending solubility or slowing/modifying crystal growth.

- Particulate matter in the form of inert (non-reactive) particles accumulates on membrane surfaces. Particulate matter can be removed with filtration.
- Biological films, in the form of bacterial colonies, spread over membrane surfaces. In the past, bacteria were a serious concern with CA membranes because they literally metabolized the membrane – PA membranes are not metabolized. Control is usually accomplished by adding oxidizing biocides to RO feedwater followed by a reducing agent[32] to protect the membrane. PA membranes are susceptible to oxidation if the residual chlorine levels continuously exceed 0.1 to 0.2 mg/l during the operating life of the membrane. Once bacteria colonies are established on the membrane, periodic cleaning of the membrane assemblies using non-oxidizing biocides may be required to control their growth[33].
- Non-soluble oil is a common constituent of produced water. Petroleum is mostly comprised of a variety of carbon-hydrogen compounds (light to heavy, straight chain to complex aromatics). Some constituents[34] tenaciously adhere to surfaces. Oil can also form emulsions in water, i.e. small droplets that are difficult to separate from water. Oil can be removed from RO feedwater by a combination of gravity separation, air flotation and filtration. Chemical additives are often used to break emulsions during treatment.
- Colloidal matter in the form of charged clusters of large organic molecules or nucleating precipitants can accumulate on membrane surfaces. Like-charged colloidal clusters repel each other, and as such, tend to spread over the membrane surface. These materials can be neutralized with polymers (carrying an opposite charge, e.g. cationic polymer) and filtered from RO feedwater.

A complicating factor with fouling is that combinations of certain foulants can exacerbate the problem. For example, bacterial foulants form sticky slimes that can embed mineral scale or inert particles. Oil can also provide base material for mineral scales and particulate matter.

Produced water from the San Juan Basin has the potential to generate all of the above fouling problems. Pretreatment is discussed in more detail as the alternatives are developed next.

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[31] Depending on the amount of magnesium removed, water temperature and contact time, precipitation softeners can also be used to remove silica from feedwater. Silica sorbs onto magnesium hydroxide floc.

[32] Reducing agents, e.g. sodium bisulfite, readily react with excess oxidizing agents. They are injected into the feedwater after the oxidizing agent has had sufficient residence time for disinfection.

[33] These compounds, which are sold as formulations by specialty chemical providers, are toxic and must be applied under controlled conditions, i.e. offline during a membrane cleaning cycle.

[34] Two compounds of concern are paraffins and asphaltenes. Paraffins are high-molecular weight hydrocarbons that form paste-like deposits under high pressure (such as the operating pressure in RO membranes). Asphaltenes, which are commonly found in oil, are large charged molecules that readily adhere to surfaces.

Produced Water Treatment Alternatives

By intention, wastewater generation is minimized in all of the alternatives discussed next. Waste streams generated by produced water treatment are recycled to the "front end" of the treatment system. Depending on the treatment alternative, final (non-recyclable) wastewater is either RO reject or BC brine. These streams would be sent to the evaporation ponds for final disposal. No wastes would be sent offsite for disposal. Sludge generated by precipitation softening would be sent to the SO_2 absorbers as supplemental limestone feedstock[35]. Use Table 4-7 as a guideline for the process configuration of each alternative, i.e. combinations of CRO, HERO® and BC. Refer to Table B.1 in Appendix B in for detailed process information (flows, chemicals, power requirements, etc.) for all of the alternatives and Table B.2 for process chemistry.

<u>CRO Only - Alternatives 1 and 6</u>

These alternatives would use conventional reverse osmosis to treat produced water (Alternative 1) and the produced water/purge water (PW/PW) blend (Alternative 6). Refer to Figure 4-8[36] for a process schematic of Alternatives 1 and 6. CRO systems are operated at low pH to reduce the calcium carbonate scaling potential of feedwater[37]. Pretreatment would include softening via lime precipitation (reactor clarifier) to reduce calcium, magnesium, barium, strontium and dissolved iron. Without softening, RO recovery would be quite low, rendering the technology infeasible. Ultrafiltration (UF)[38] would be used upstream of the RO to filter particulate matter and remove organic foulants. Basket strainers would be used to protect the UF from particulate loading generated by the reactor clarifier. Sodium hypochlorite (disinfectant) would be added upstream of the strainers and reduced with sodium bisulfate upstream of the UF and RO to protect membranes. A degasifer would be used to remove CO_2 generated by acid addition (for pH control). Softening and UF are designed to minimize, <u>but not eliminate</u>, scale formation and organic fouling. Additionally, at low pH, RO is more susceptible to oil fouling[39] making UF critical to successful RO operation.

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[35] Sludge generated by precipitation softening would be similar in assay to the dolomitic limestone used by SJGS for SO_2 scrubbing. Depending on the produced water treatment configuration, precipitation softener sludge would replace 2 to 10 percent of the limestone used by the plant. Based on known produced water chemistry, there are no observed or obvious constituents that would interfere with the SO_2 absorption process or SO_2 absorber sludge stability (leaching characteristics). This would require more review if implemented.

[36] Figure 4-8 is a simplified schematic – an actual RO system would have more pressure tubes, inter-stage pumping, etc.

[37] At low pH (typical range is 4.5 to 5.5), carbonate alkalinity in the feedwater is converted from a mixture of carbon dioxide (CO_2), bicarbonate (HCO_3^{-1}) and carbonate (CO_3^{-1}) to mostly CO_2. This reduces the likelihood of calcium carbonate ($CaCO_3$) formation.

[38] Ultrafiltration (or a similar type membrane process, such as micro-filtration), is usually recommended for water with a high fouling potential such as produced water. There are a number of membrane configurations – some similar to RO, e.g. spiral wound and tubular membranes. There are also ceramic media configurations.

[39] At low pH, certain oils become less soluble and present a greater potential for deposition.

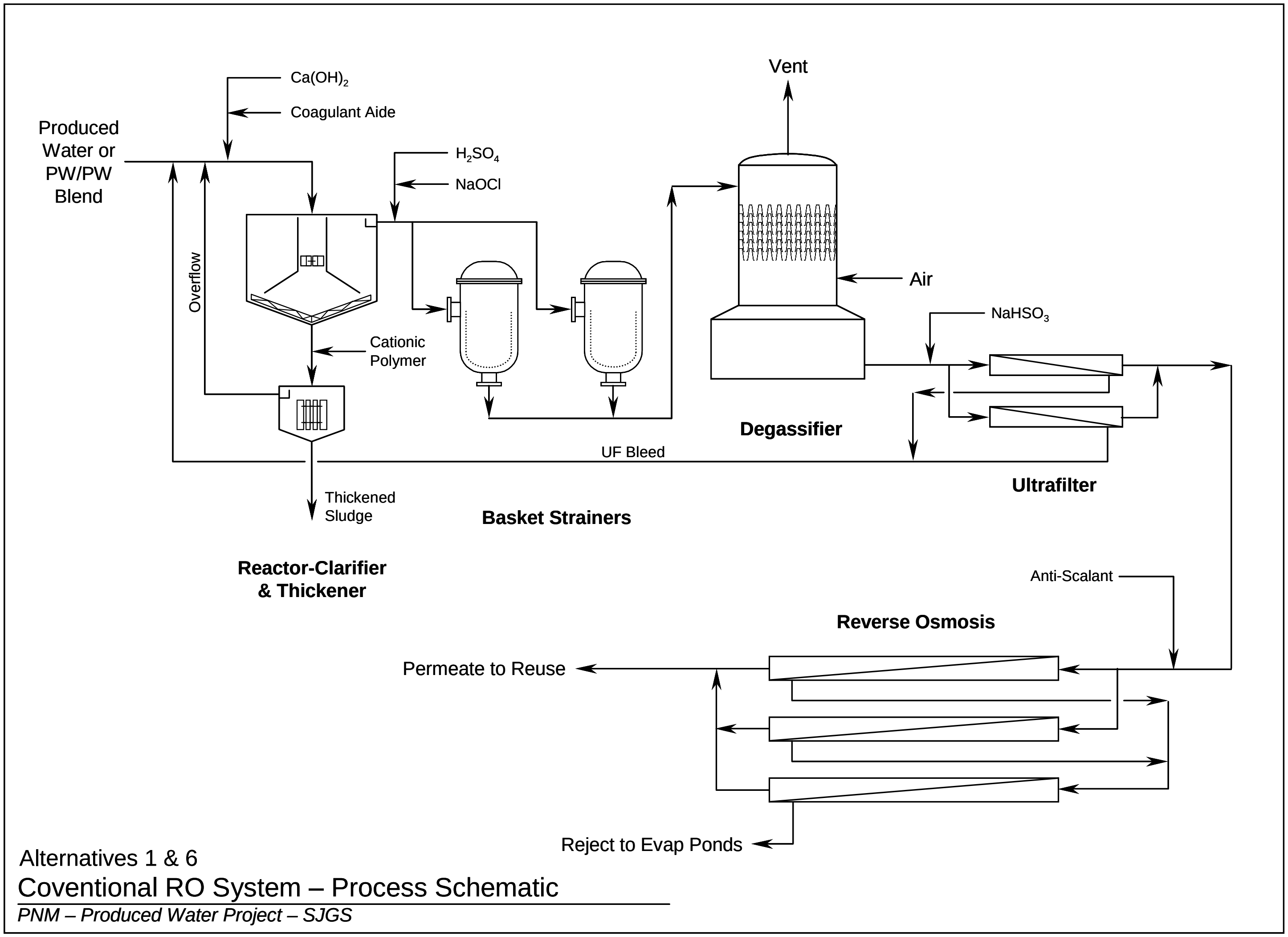

Figure 4-8
Alternatives 1 and 6: Conventional RO System Process Schematic

Softener sludge, which is mostly calcium carbonate (limestone) would be dewatered in a thickener and sent to the limestone preparation for use as feedstock in the absorbers. A coagulant aide and cationic polymer would be used to assist the reactor clarifier and thickener in dewatering sludge.

UF bleed and sludge thickener overflow – would be recycled to the front end of the system. After softening and filtration of the feedwater, the CRO would operate at 77 percent recovery. Calcium sulfate, barium sulfate and strontium sulfate would still form at this recovery level and would have to be controlled with a scale inhibitor/crystal modifier. Even with all the pretreatment precautions, the CRO would likely require cleaning every one to two months to remove mineral scale, organic foulants and biological growth.

Cleaning would be accomplished by isolating one section of the RO, connecting it (via portable hoses) to a clean-in-place (CIP) skid and flushing it with one or more cleaning solutions. Cleanings are planned around RO performance, e.g. they are initiated when RO flux (throughput at a given feed pressure) drops below a certain threshold. Treatment is usually customized for whatever problem is expected. On occasion, one membrane element is removed and cut apart to determine the types of foulants present and the overall condition of the membrane.

HERO® Only - Alternatives 2 and 7

High-efficiency reverse osmosis consists of pre-softening for the complete removal of hardness and the operation of the RO at high pH (9.5 to 10.5). Refer to Figure 4-9 for a process schematic. This mode of operation would significantly minimize common RO operating problems such as hardness (calcium, magnesium, barium and strontium) and silica scaling. Also, at high pH, organic fouling and certain oil constituents would dissolve, minimizing fouling in the RO.

Pretreatment would include softening via lime precipitation (reactor clarifier) to reduce calcium, magnesium, barium, strontium and dissolved iron (same as Alternatives 1 and 6). Media filters (sand and anthracite) would be used to protect weak acid cation (WAC) [40] ion exchangers from particulate fouling. WAC would be used because all of the hardness is eventually returned to and removed by the reactor clarifier. Wastewater generated by the pretreatment system – filter backwash, effluent hardness from the reactor clarifier, thus removing the potential of mineral scale in the RO. A degasifer would be used to remove CO_2 generated by the WACs. Sodium hypochlorite would be added upstream of the degasifier and reduced with sodium bisulfite upstream of RO to protect membranes.

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[40] In this service, WAC ion exchangers would be operated in the hydrogen form, i.e. they would exchange hydrogen ions (H^{+1}) for divalent (calcium, magnesium, barium, strontium, etc.) and trivalent ions (aluminum, iron, etc.) associated with alkalinity. If 100 equivalents of alkalinity were in the feedwater, WAC would exchange 100 equivalents of H^{+1} for 100 equivalents of hardness ions. Produced water and the PW/PW blend have alkalinity concentrations significantly higher than divalent/trivalent metals so all of the hardness would be removed by the WACs.

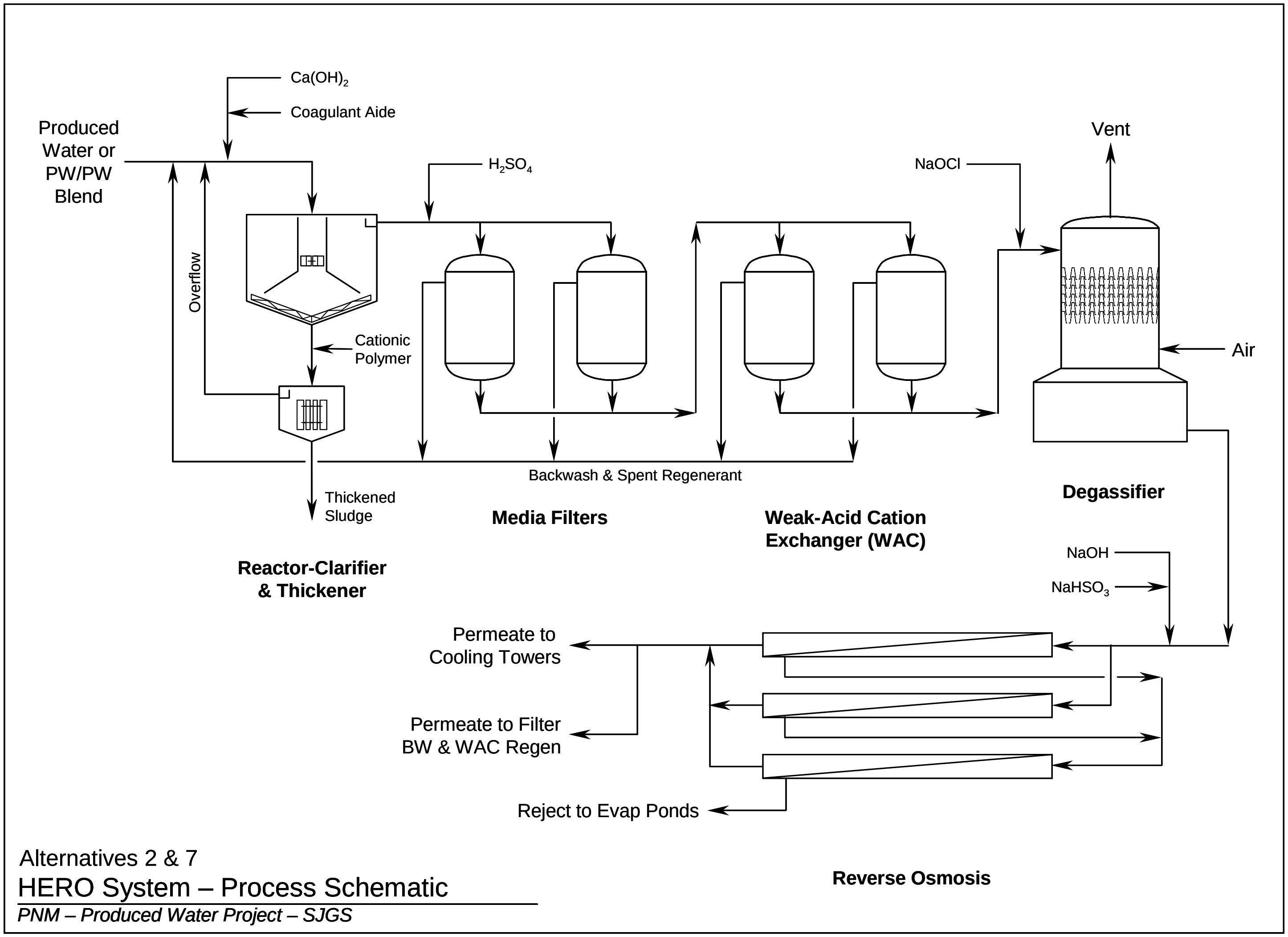

Figure 4-9
Alternatives 2 and 7: HERO System Process Schematic

Softener sludge would be dewatered in a thickener and sent to limestone preparation (similar to Alternates 1 and 6). Alternatives 2 and 7 would generate more solids than WAC spent regenerant and sludge thickener overflow, which would be recycled to the front end of the system.

After softening and filtration, HERO® could operate at 86 percent recovery. Recovery would be higher than CRO, because calcium, barium, strontium, etc. were removed from the feedwater (no concern of mineral scales). At elevated pH (>10), silica[41] dissociates to form silicates and does not form silica scale. Also, in the absence of divalent and trivalent metal ions, silicate scales do not form.

As a precautionary measure, HERO should be cleaned every twelve months to remove trace amounts of mineral scale, organic foulants and biological growth.

SJGS BCs Only - Alternatives 3 and 8

In these alternatives, produced water and PW/PW would be fed to refurbished brine concentrators, BC 2 and BC 3. The combined capacity of the BCs is 1,160 gpm (580 gpm each). This falls just short of the required capacity for Alternative 3 of 1,216 gpm of produced water and Alternative 8 of 1,316 gpm of PW/PW.

The BCs would be operated in the seeded mode, because of the presence of scaling constituents in the feedwater. Refer again to Figure 4-6. In the seeded mode, the pH would be kept between 4.0 and 5.0 to minimize scale deposition in the feedwater heat exchanger and condenser. Low pH also converts feedwater alkalinity to CO_2 (to eliminate $CaCO_3$ scale). CO_2 would be removed in the deaerator (along with oxygen to minimize corrosion potential). Anti-scalant would also be required to minimize scale deposition on BC heat transfer surfaces.

A significant amount of calcium chloride ($CaCl_2$) would be added to the BCs to promote seed formation – 16.0 tpd and 21.6 tpd, respectively for Alternatives 3 and 8. There is not enough calcium relative to sulfate in produced water or the PW/PW blend to promote crystal formation. Insufficient crystal formation would lead to severe scaling problems in the BCs. With sufficient nucleation sites available, crystals circulate with the BC bulk fluid and tend not to deposit.

BC 2 and BC 3 have the requisite 316 LM stainless steel (5+% molybdenum) metallurgy to operate at low pH and very high chloride levels. Some minor metallurgy upgrades would have to be made to both BCs (316LM cladding in certain areas) to operate in this mode. The operating limit for chloride would be 50,000 mg/l in the BCs and would limit recovery to 87 and 86 percent for Alternatives 3 and 8, respectively.

BC 2 and BC 3 would have to be cleaned (hydro-lancing the condenser tubes) every 12 months to maintain operating recovery and throughput.

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[41] Soluble SiO2 is a weak acid in water – H_4SiO_4 – silicic acid. Silicic acid dissociates to $H_3SiO_4^{-1}$ at pH greater than 10, and at higher pH, $H_2SiO_4^{-2}$.

CRO & SJGS BCs - Alternatives 4 and 9

In these alternatives, CRO would be operated as a pre-concentrating device ahead refurbished BC 3 (the better of the two idled BCs). Pretreatment for Alternatives 4 and 9 would be the same as Alternatives 1 and 6 (refer to Table 4-7). The BC would be operated in the seeded mode, because saturated levels of scaling salts in the reject from the CRO would be fed to the BC. Some minor metallurgy upgrades (316LM cladding in certain areas) would have to be made to BC 3 to enable it to operate in this mode.

A significant amount of calcium chloride ($CaCl_2$) would also be added to the BCs to promote seed formation – 15.2 tpd and 15.3 tpd, respectively for Alternatives 4 and 9. BC recovery would be limited to 48 and 53 percent of RO reject for Alternatives 4 and 9, respectively. Chloride levels would limit recovery. The overall recovery (CRO and BC) would be 88 percent for Alternatives 4 and 9.

BC 3 would have to be cleaned (hydro-lancing the condenser tubes) every 12 months to maintain operating recovery and throughput.

HERO® & SJGS BCs - Alternatives 5 and 10

In these alternatives, HERO® would be operated as pre-concentrating device ahead of refurbished BC 3 (the better of the two idled BCs). Pretreatment for Alternatives 5 and 10 would be the same as Alternatives 2 and 7 (refer to Table 4-7). The BC would be operated at high pH in the un-seeded mode. The benefit of high-pH operation (10 to 11)[42] is that the BC could be operated at high levels of chloride without any alloy upgrades to stainless steel wetted parts. The need for acid, anti-scalant and $CaCl_2$ would be eliminated, because there would be no scaling salts in the BC feedwater.

BC recovery would be limited to 73 percent of RO reject for Alternatives 5 and 10. This is significantly higher than seeded-mode operation, because total solids is the limiting factor rather than chloride. Total solids would be controlled between 200,000 mg/l to 250,000 mg/l. The overall recovery (HERO® and BC) for Alternatives 5 and 10 would be 95 and 97 percent, respectively.

As a precaution, BC 3 should be cleaned (hydro-lancing the condenser tubes) every 36 to 48 months.

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[42] The BC must be operated at a pH less than 12.0 to avoid embrittlement of the titanium tubes (Ti2 alloy) in the BC heat exchanger. Titanium embrittlement information was provided by James Grauman of Timet Corp. (April 14, 2004 phone conversation). Timet provides titanium tubes and sheet to BC manufactures.

4.4.6 Preliminary Economic Analysis of Treatment Alternatives

A preliminary cost analysis is presented in this section of the report. The analysis is used to determine which alternative is the most economically feasible to treat and reuse produced water at SJGS. Table 4-8 is a summary the results of the analysis and includes:

- Feedwater flow, overall system percent recovery and recovered water flow
- Recovered water quality (TDS)
- Additional evaporation pond capacity required for produced water treatment
- Additional operating and maintenance staff
- Power requirements
- Capital and operating costs
- Unit operating cost ($/1,000 gallons of recovered water)

A detailed analysis of produced water recovery is presented in Section 7, *Cost/Benefit Analysis.* Produced water flow is predicted over the life of the project for five recovery cases and three production declination scenarios[43] (for a total of 15 evaluations). The flow rate selected here for preliminary economic analysis is within the likely produced water recovery envelop (roughly midway).

Process and cost support information is presented in Appendix B for all of the alternatives. Refer to Table B.1 for process information details, Table B.2 for process chemistry, Table B.3 for capital and operating cost development and Table B.4 for the assumptions used in the cost analysis.

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[43] As oil and gas fields mature, production and produced water generation decline.

Table 4-8
Produced Water Treatment – Preliminary Economic Analysis

Produced Water Treatment - Preliminary Economic Analysis

PNM - Produced Water Project - SJGS

Alternative	Produced Water	FGD Purge	CRO	HERO	BC 2	BC 3	Flow Basis gpm (4)	Overall Recovery	Recov'd Water gpm	Recov'd Water TDS mg/l	Add'l Required Evap Ponds acres (3)	Total Addl Staff	Total Power Reqm't kw	Capital Cost (2) $MM	Total Annual Cost (1) $MM	Unit Op Cost $/kgal Net
1	X		X				1,216	76.9%	935	270	140	5.2	650	$46.09	$6.42	$13.07
2	X			X			1,216	86.1%	1,047	260	87	5.2	600	$31.35	$4.08	$7.41
3	X				X	X	1,160	87.0%	1,009	10	75	5.2	4,830	$29.12	$7.23	$13.64
4	X		X			X	1,216	87.8%	1,068	240	73	5.2	1,718	$32.58	$6.54	$11.65
5	X			X		X	1,216	96.6%	1,174	240	24	5.2	1,774	$18.92	$3.13	$5.07
6	X	X	X				1,316	74.6%	981	260	114	5.2	720	$39.77	$5.68	$11.01
7	X	X		X			1,316	82.3%	1,083	270	67	5.2	660	$26.53	$3.86	$6.78
8	X	X			X	X	1,160	86.1%	999	10	31	5.2	4,780	$17.46	$6.43	$12.24
9	X	X	X			X	1,316	87.8%	1,155	230	27	5.2	1,875	$21.13	$5.37	$8.85
10	X	X		X		X	1,316	95.3%	1,255	230	0	5.2	1,915	$14.12	$2.98	$4.52

Notes…..

1. Includes capital recovery at 7.5% for 20 years.
2. Does not include costs for offsite equipment - Collection Center in Bloomfield and the 28.5-mile pipeline.
3. Alternatives 6 to 10 receive a 50-acre credit for Purge Water capacity.
4. The flow basis is approximate and within the likely produced water recovery range.

Based on the analysis, Alternative 10 is the most economically feasible approach. Alternative 10 consists of treating the PW/PW blend with HERO® and BC 3 (the better of the idled BCs). The alternative has the lowest evaluated capital cost ($14.1 million) and operating cost ($2.98 million per year), would recover the most produced water for reuse (1,255 gpm) and would require no additional evaporation ponds.

Of the alternatives that use combinations of RO and BC (4, 5, 9 and 10), Alternative 10 would use the most power – 1,915 kw (there is a 200 kw spread among these alternatives). Alternatives 3 and 8, which employ both BC 2 and BC 3, clearly would require the most power (4,830 kw) and are limited in total treatment capacity.

SJGS determined that additional operating and maintenance coverage would be the same for all of the alternatives. One additional operator for each shift (8,760 hours per year – equivalent to 4.2 staff) and one shift of maintenance coverage (1 person) would be required.

Produced water feed rate, produced water reclaimed for reuse, and capital and operating costs for Alternative 10 are refined in Section 7, *Cost/Benefit Analysis*. Full project costs (Bloomfield Collection Center, pipeline and treatment at SJGS) are also presented in Section 7. PNM's implementation plan is discussed in Section 8.

4.4.7 Disposition of Treatment Wastes

Heavy Metals

As shown in Table 4-6, heavy metals are present at detectable levels in produced water and purge water. The metals fall into two groups:

- Cations – Cu, Zn, Pb, Hg and Ag
- Anions – AsO_4, CrO_3, SeO_3/SeO_4 and UO_3

In the alternatives, employing precipitation softening, most of the Cu and Zn, almost all of the Pb and Hg, and some of the AsO_4 and SeO_3/SeO_4 would be removed as precipitants in reactor clarifier sludge. The cations would be removed as $Cu(OH)_2$, $Zn(OH)_2$, etc., and the AsO_4 and SeO_3/SeO_4 would be co-precipitated[44]. If some of the precipitated metals resolubilized[45] in the absorber (recall that reactor clarifier sludge would be blended with limestone feedstock), they would be removed:

- By softening the Purge Water.
- Along with waters of moisture in absorber gypsum sludge.
- Via co-precipitation in absorber gypsum sludge.

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[44] AsO_4 and SeO_3/SeO_4 sorb onto nucleating crystals and become entrapped as precipitate forms.

[45] This is not likely for the highly insoluble salts like $Pb(OH)_2$ and $Hg(OH)_2$.

Metals not removed by softening would be in the RO reject stream[46] in alternatives employing membrane processes. Reject would either be sent to the evaporation ponds or to BC 3. All of the heavy metals in BC feedwater would be in the waste brine stream to the evaporation ponds.

Lastly, the amount of heavy metals loading[47] in produced water is equivalent to 2.72 kilograms per day (kg/day) at the assumed feedwater flow rate. Based on the analytical data found in Table 4-4, the Purge Water stream alone generates 3.90 kg/day of metals loading. Even though the combined waste stream would generate a 70 percent increase in metals loading, most of it would be bound in either precipitation softener sludge or SO_2 absorber sludge or lost as moisture in SO_2 absorber sludge. Therefore, if the Purge Water stream were treated (Alternative 10), the resultant metals loading to the evaporation ponds would decrease.

Volatile Organic Compounds

VOCs are commonly found in produced water. They are usually comprised of a group of aromatic compounds collectively known as BTEX – benzene, toluene, ethylbenzene and xylene. BTEX is expressed as the sum of the concentrations of these compounds. The compounds are volatile, i.e. given good water-air contact, BTEX constituents readily vaporize from water into the air. Table 4-9 summarizes BTEX data for conventional and CBM produced water. Conventional water clearly has higher levels of BTEX than CBM produced water. Also note that McGrath SWD receives a mix of conventional and CBM produced water.

Table 4-9
BTEX Concentrations for Conventional and CBM Sources, San Juan Basin

Produced Water Site	Type	BTEX mg/l	Benzene mg/l	Toluene mg/l	Ethyl-Benzene mg/l	Xylene mg/l
McGrath SWD	Conv/CBM	22.610	4.700	11.000	0.510	6.400
McGrath SWD	Conv/CBM	9.960	0.900	0.940	3.200	4.920
Taber Battery	CBM	0.562	0.060	0.150	0.050	0.302
Taber Battery	CBM	0.207	0.069	0.017	0.037	0.084
Turk's Toast	CBM	0.198	0.002	0.012	0.160	0.023
Salty Dog 2 SWD	CBM	0.124	0.036	0.007	0.057	0.024
Middle Mesa SWD	CBM	0.166	0.008	0.047	0.013	0.098
Pump Canyon SWD	CBM	0.288	0.004	0.120	0.011	0.151

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[46] Salt passage for heavy metals is very low – 99.7 to 99.9 percent rejection should be expected at the membrane.

[47] Loading is calculated by summing the heavy metals concentrations found in Table 4-4. If a concentration is less than the PQL (non-detectable), then the PQL is used as its concentration. Mass loading is calculated as follows: $Mass\,Loading, kg/day = 0.00545\,x\,Flow\,Rate, gpm\,x \sum (Me_i, mg/l)$.

Approximately 50 percent of the BTEX in the produced water delivered to the Collection Center in Bloomfield would be removed by one of the oil removal processes – gas flotation (discussed next). The remaining BTEX would be diluted with Close-in produced water, cooling tower blowdown from Prax Air, BHP Billiton mine water and absorber Purge Water. The concentration of BTEX in the produced water blend to be treated at SJGS would likely range from 1 to 4 mg/l (equivalent to 14 to 56 pounds per day of BTEX at 40,000 BPD produced water delivery). Most of it should be removed to atmosphere in the degasifier air stream and the BC deaerator.

Finally, there could be trace levels of BTEX in the treated produced water. Given dilution with fresh water and gas-liquid contact in end-use processes, BTEX should be at non-detectable levels in the SO_2 absorber liquor or cooling tower circulating water, i.e. most of the BTEX should be lost to the flue gas or cooling tower air stream.

4.5 Collection Center in Bloomfield

The Collection Center in Bloomfield would have three functions:

- Provide a collection point for produced water delivered by the CO_2 Gas Line and the Hart Canyon Line.
- Pretreat produced water to remove oil and grit prior to conveyance.
- Equalize the chemistry of the produced water prior to charging the pipeline from Bloomfield to SJGS.

Refer to Figure 4-10 for a process schematic of the Collection Center. The Collection Center would process produced water from conventional oil and gas production and CBM wells (mostly from CBM wells). Two tanks would be used to receive and store produced water delivered to the Collection Center. Three oil-removal technologies would be used to pretreat the water – gravity separation, gas flotation and media filtration.

A gravity-coalescing oil/water separator would be used to remove un-dissolved[48] and floatable oil as well as grit. An emulsion breaker (EB) would be fed to the water prior to the oil/water separator to de-emulsify oil that is finely dispersed. The separator allows oil to float to the surface in the forward compartment, and inclined plates in the rear compartment collect and agglomerate smaller droplets of oil. Grit falls to the bottom of the separator in both compartments. Oil and grit comes from convention oil and gas production. Produced water from CBM wells is almost free of oil byproducts. Skimmed oil from the separator would be stored separately in a tank and sold to the Giant Oil Refinery in Bloomfield for reclamation. Grit would be removed occasionally to a grit tank and disposed offsite at an approved landfill.

Water from the separator would then be fed to a gas flotation unit where additional oil would be removed. Fine bubbles of air are used to float oil droplets to the surface which are removed to side troughs (this foamy mixture is also known as float). Float would also be removed to a grit tank for disposal offsite. The flotation units should also remove up to 50 percent of volatile BTEX constituents found in conventional produced water (discussed previously in 3.4.7).

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[48] Some petroleum-based organic chemicals are water soluble and cannot be removed by physical means, e.g. gravity separation, flotation or filtration. The compounds would eventually be removed by the HERO process at SJGS.

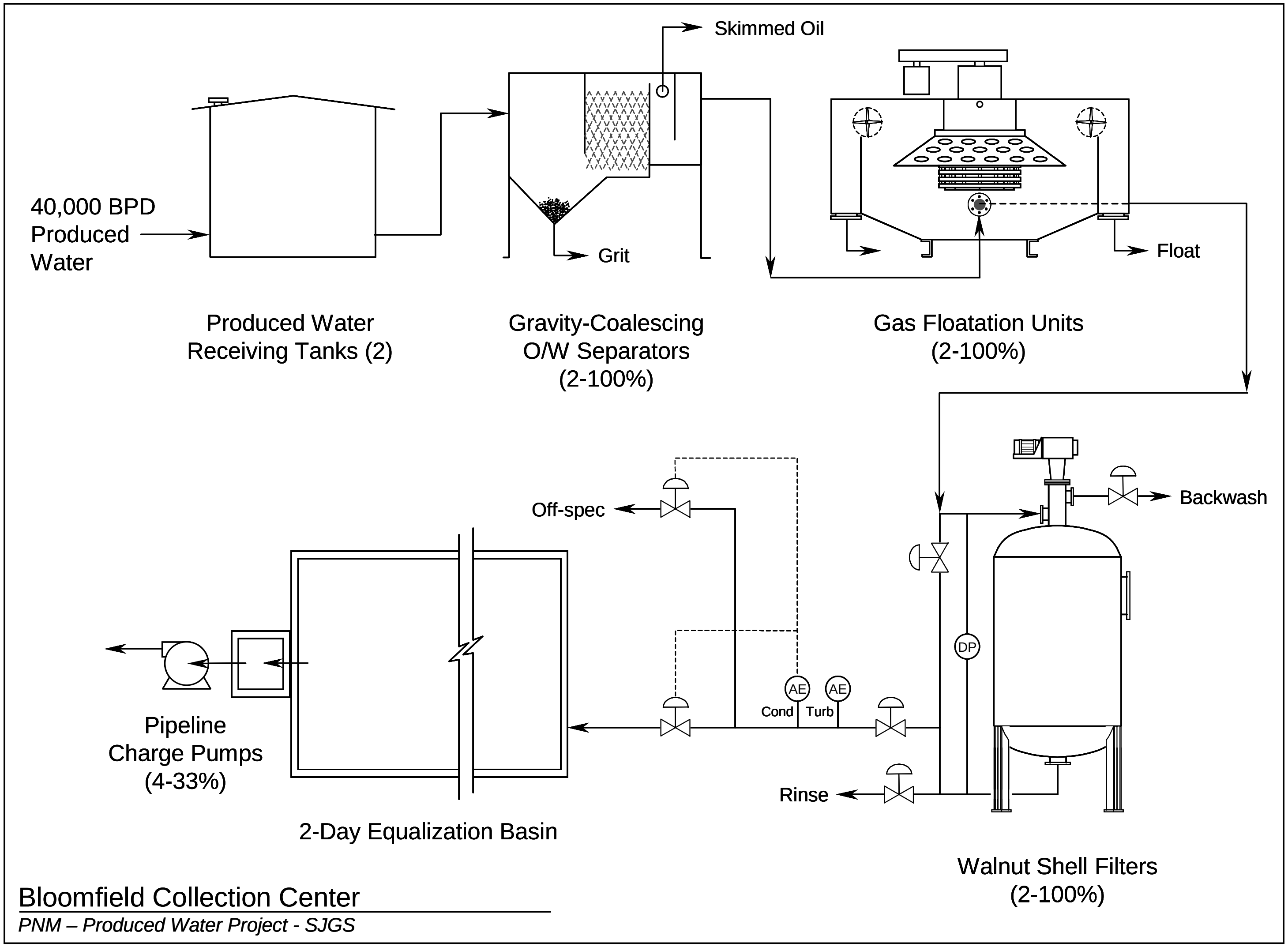

Figure 4-10
Bloomfield Produced-Water Collection Center

Finally, water would be fed to a walnut shell media filter to remove trace levels of oil that escape the oil/water separator. Walnut shell media is used extensively in oil field applications to remove separable oil. Backwash from the walnut filters would be recycled back to the produced water receiving tanks for reprocessing. Rinse from the filters would be recycled back to the walnut filter feed tank. The filter effluent would be monitored for turbidity to initiate the backwash cycle. Filter effluent would also be monitored for conductivity to segregate produced water with very high salinity to an off-spec hold tank. Additional automated testing could be added at this point to identify other off-spec water parameters.

Filtered water would be held in a two-day basin to allow its chemistry to equalize prior to conveyance to San Juan Generating Station. Water from the off-spec tank would be slowly blended into the equalization basin. Water that is significantly off-spec (or that cannot be blended in a timely manner) would be trucked offsite and disposed of via deep well injection at one of two licensed facilities in the Bloomfield area.

Capital and operating costs for the Collection Center in Bloomfield are found in Tables B.5 and B.6 in Appendix B.

4.5.1 Disposition of Volatile Organics

About 50 percent of the BTEX delivered to the Collection Center would be removed by the gas flotation unit to atmosphere. Fine air bubbles, which buoy oil droplets to the surface in the flotation unit, also provide extensive water/air contact to allow BTEX to volatilize into air. About 4 to 12 mg/l of BTEX should be in the produced water delivered to the Collection Center. If half is removed by the gas flotation unit air stream (14 to 56 pounds per day of BTEX), water shipped to SJGS should have a BTEX content of 2 to 6 mg/l. Produced water BTEX levels would be diluted by Close-in CBM water and a number of non-petroleum streams to 1 to 4 mg/l by the time it reaches SJGS.

5

EMERGING WATER TREATMENT TECHNOLOGY TESTING

5.1 Introduction

Public Service of New Mexico (PNM) is evaluating produced water as a supplemental source for the San Juan Generating Station (SJGS). In conjunction with the project, bench-scale testing was conducted by CeraMem Corporation to evaluate ceramic membrane filtration. The process could significantly reduce the level of certain forms of contamination in produced water, i.e. oil and particulate matter. A benefit of this technology is that ceramic membranes could last for a significant period of time, thereby reducing the operating cost of pretreatment. As such, it could be a valuable pretreatment process for reverse osmosis (RO). Testing, which was exploratory in nature and showed promising results, was conducted over a nine-day period at the McGrath Salt Water Disposal (SWD) Facility[1] in July 2005.

5.2 Process Concept

Ceramic filters have been used for filtration for many years, however, produced water in particular has been very difficult to treat, i.e. oil and particulate fouling of membrane surfaces has been especially problematic. In recent years, CeraMem, in conjunction with a large oil company, tested their ceramic membranes on produced water at two locations (Southeast USA and Western Canada). The focus of the testing was to sustain flux over reasonable periods of time while maintaining separation efficiency (oil removal from water). As part of this work, a proprietary ceramic membrane and cleaning technique were developed that was able to maintain process flux for several hundred hours between cleanings.

Crossflow ceramic filters are cylindrical in shape and are comprised of an array of passageways resembling a honeycomb. Refer to Figure 5.1 for a schematic description of a ceramic filter. Filtered water permeates through the ceramic membrane (crossflow to feed flow) and then through the supporting monolith substrate. Filtered water, which is known as permeate, flows toward the periphery of the monolith and is removed through an integral casing at the exterior of the filter. Impurities and a fraction of the feedwater stay behind as retentate (the waste stream) to exit the passageways for disposal (or further treatment). This process configuration allows for continuous operation between membrane cleanings.

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[1] The McGrath SWD is owned and operated by Burlington Resources.

In the manufacturing of the filter elements, a ceramic membrane[2] is uniformly deposited on the surfaces of the passageways of a honeycomb monolith by slip casting[3] coatings of ceramic particles. The coatings are dried and sintered at elevated temperature to bond the particles to the monolith and each other. The resulting coating is porous, with the pore size controlled primarily by the particle size used in the slip. CeraMem membranes can be comprised of two or three layers (depending on the intended service). An initial underlying membrane layer has a pore size of about 0.5 µm and thickness of about 50 µm. Subsequent layers are thinner to minimize flow resistance and contain finer ceramic particles to form finer pore sizes. Figure 5-2 shows a scanning electron micrograph of a cross section of a three-layer membrane. Membranes can also be made of non-ceramic materials, e.g. Teflon®.

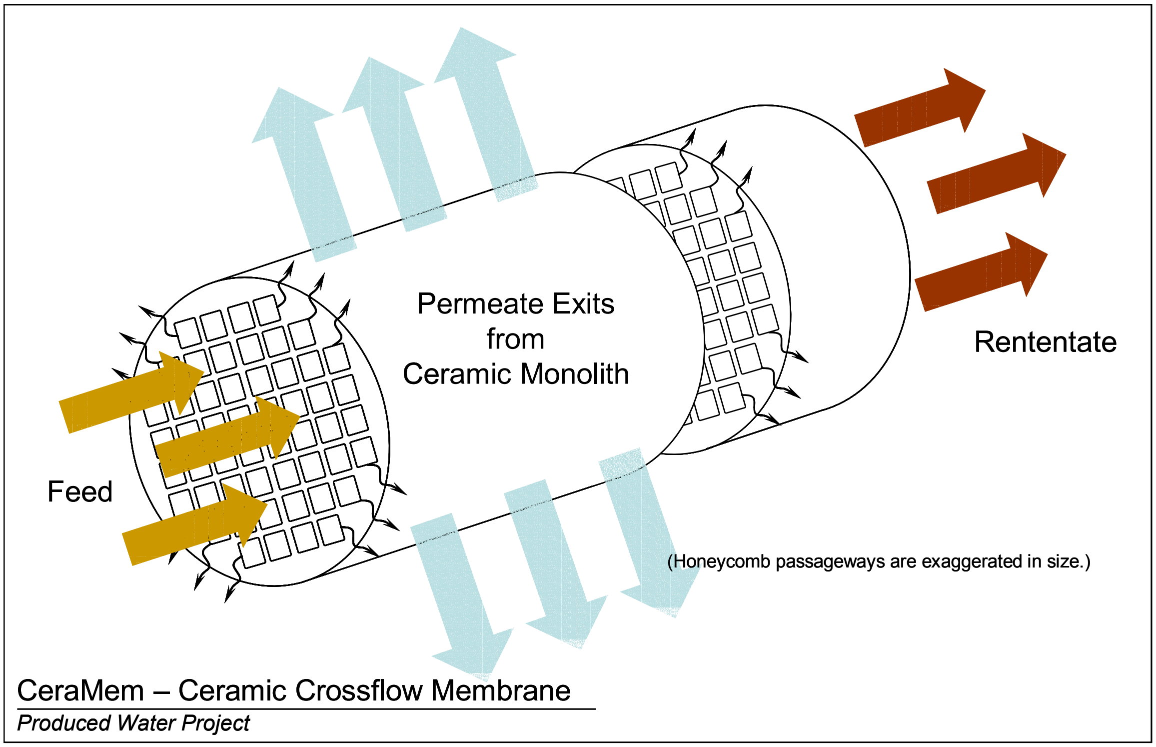

Figure 5-1
CeraMem Ceramic Crossflow Membrane

The potential benefit of this technology is that ceramic membranes could last for a significant period of time, thereby reducing the operating cost of pretreatment. A drawback of this technology is the relatively high first-time capital cost for the membrane assembly and ceramic filters as compared to conventional RO pretreatment.

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[2] The term “membrane” refers to a layer(s) of specifically sized ceramic particles.

[3] Slip is a slurry of specifically sized ceramic particles that are circulated over the surfaces of the passageways. As the process proceeds, a ceramic layer is uniformly deposited.

5.3 Testing at McGrath SWD

Bench-scale testing was conducted at the McGrath SWD. This SWD is an ideal location to test this type of equipment because it receives produced water from a range of sources, and thus water quality varies dramatically. Refer to Figure 5-3 for variations in total suspended solids (TSS) and total petroleum hydrocarbons (TPH).[4]

Water at McGrath SWD is initially stored in receiving tanks and then passed through an API-type oil-water gravity separator to remove grit and floatable oil. The pretreated water is placed in intermediate storage and then passed through two levels of filtration – deep-bed sand filtration followed by 5-micron cartridge filtration. The water is then ready for deep well injection.

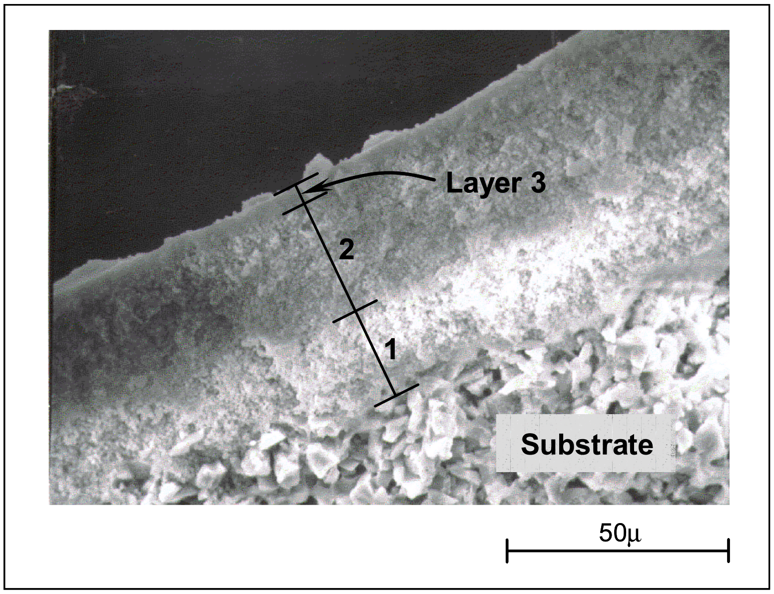

Photo courtesy of CeraMem Corporation

Figure 5-2
Scanning Electron Photomicrograph of Multilayer Ceramic Membrane

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[4] These data were obtained over a thirty day period of sampling at McGrath SWD in 2003.

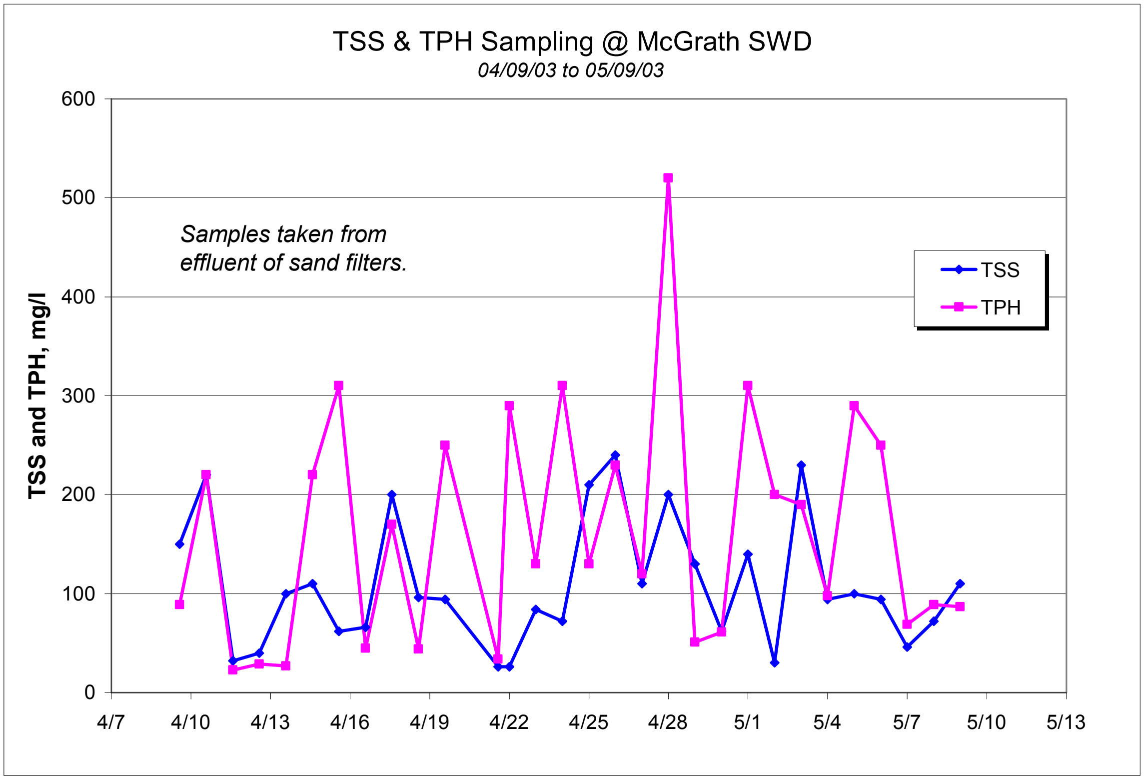

Figure 5-3
Total Suspended Solids and Total PH at McGrath Solid Waste District

It was decided that the CeraMem test equipment should treat effluent from the sand filters. In particular, suspended matter and oil (in the form of stable emulsions) were the target constituents in this testing. Treating pre-filtered produced water with the ceramic membranes (after media filtration) could eliminate the need for ultrafiltration and cartridge filtration for RO. Refer to Section 4, Treatment and Disposal Analysis, for a detailed discussion of RO pretreatment options. The TSS and TPH found in Figure 5-3 were sampled from the effluent of the sand filters.[5] It may also be possible to use this technology to filter produced water at SWDs, since cartridge filter replacement is one of the largest costs associated with deep-well disposal.

5.3.1 CeraMem Test Equipment

Figure 5-4 presents a schematic of the CeraMem test apparatus. The CeraMem Test Report in Appendix C provides a detailed description of the apparatus (along with photos and a more comprehensive schematic). The pilot test skid, which was 3 feet long x 2 feet wide x 3 feet tall, consisted of a feed tank, charge pump and two ceramic membrane modules. The modules could be operated in series or the first module (after the charge pump) could be operated alone. The test apparatus was instrumented with flow and pressure indicators to enable the measurement of

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[5] Note that TSS and TPH levels were high and varied significantly even after sand filtration. It is likely that the sand filters at McGrath were not performing properly. It was assumed that the level of filtration media (sand) was low. Insufficient media depth would show similar results. This was discussed with BR after the data were tabulated.

flux and trans-membrane differential pressure during the test run. Sample taps were used to monitor TSS and TPH in the feed, permeate and retentate.

5.4 Test Results

Ceramem was on site at McGrath for nine days and logged 122 hours of testing. Six runs were conducted during pilot testing using two membrane materials – silica and Teflon®. Runs 1 through 5 were exploratory and were used to test the response of the membranes to produced water under a number of operating conditions. The test runs were complicated by the fact that produced water TSS and TPH varied significantly from run to run. However, the variability was also beneficial, because it allowed CeraMem to more completely identify critical processes that control the filtration steps, i.e. membrane fouling, emulsion-forming mechanisms[6], suspended solids passage, etc. Each test run is explained in detail in the CeraMem test report found in Appendix C.

It was not until the last test run, when surfactant[7] was added to the feed stream, that emulsion and flux stability were dramatically improved. This run demonstrated that flux could be sustained with low TSS (as measured by turbidity, NTU).[8] The membranes performed best – high flux with low permeate NTU – when stable emulsions were formed. It was determined that surfactant dosing was required for continued process performance. Also, Run 6 showed that different membrane materials strongly affect surface chemistry and thus flux and permeate quality. For example, the silica and Teflon® membranes that were tested behaved differently when the emulsion was stable. Membrane material selection is therefore critical for a given set of operating conditions – source water characteristics, desired permeate flux, permeate quality requirements, etc. Lastly, this run demonstrated that permeate quality and flux rate were tightly linked – when one was good, so was the other.

Due to budget and time constraints, additional tests could not be run. So the approach of utilizing surfactants to stabilize emulsion and flux could not be tested further. The results of Run 6 will provide a good starting point for future produced water testing.

5.5 Process Economics

Ceramem evaluated the capital and operating costs of a 53,000 BPD system.[9] Ceramic membranes could be used to pre-filter RO water in lieu of deep-bed media filters or ultrafilters. The RO configuration evaluated for cost analysis was the HERO® system (high efficiency

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[6] Emulsion formation is beneficial, because as emulsions, oil can be retained by the membrane surface. Soluble oil passes through the membrane.

[7] Surfactants alter the characteristics of membrane surfaces as well as emulsion structures.

[8] Turbidity, which is a measure of relative clarity, can be measured by a light scattering and absorption technique. Turbidity is typically recorded as Nephelometric Turbidity Units (NTU).

[9] The size of the produced water system evaluated in Section 7, Cost/Benefit Analysis.

reverse osmosis) which utilizes deep-bed media filters for RO pretreatment. Other RO systems considered in Section 4 for this service incorporate ultrafilters. The costs developed by Ceramem were based on results of Run 6 (one test run only), and as such, should be considered very preliminary.

Ceramic membrane system was estimated to cost $3.9 million installed. The system was based on a flux rate of 120 lmh.[10] This rate was considered conservative, i.e. a higher flux rate (a system utilizing fewer ceramic elements) would reduce the cost of the system. The unit operating cost for the system was estimated to be $0.051 per barrel ($400/AF). This cost includes capital recovery, operator time, chemicals, ceramic membrane replacement, power, etc. Appendix C presents the cost analysis developed by Ceramem.

To put these costs into perspective, the installed cost for the HERO® system was estimated at $11.8 million and its unit operating cost was $400 to $1,000/AF <u>after tax credits and producer cost participation</u>. The total cost for the complete project was $3,000/AF – gathering system, collection center, pipeline and HERO® system. Refer to Section 7 for a detailed discussion of project economics. The Ceramem costs are within the boundaries of the cost estimate developed for a produced water treatment system for SJGS.

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[10] Flux rate is expressed as liters per square meter per hour, lmh.

6

TREATED PRODUCED WATER COMPATIBILITY ASSESSMENT

6.1 Introduction

The compatibility of treated produced water is assessed in this section of the report. Treated produced water is evaluated as a supplement to (or replacement of) freshwater at San Juan Generating Station (SJGS) for the following plant uses:

- Bottom ash sluice water
- Fly ash wetting water
- Cooling tower make-up
- SO_2 absorber make-up

Each area is assessed for flow capacity and chemistry, i.e. constituents of concern, corrosion and deposition potential. Costs associated with the use of treated produced water in each area are assessed and summarized.

6.2 Treated Produced Water Flow and Chemistry

As discussed in Section 4, *Produced Water Use, Treatment, and Disposal Analysis*, produced water must be treated prior to use at SJGS, primarily because of high levels of TDS and chlorides. Public Service of new Mexico (PNM) would treat produced water at SJGS with the HERO® process along with BC 3 – the Alternative 10 treatment process.[1] The produced water feed rate would range from 750 to 1,400 gpm (1,210 to 2,260 AF/yr)[2] over the life of the project with an average flow of 1,105 gpm (1,790 AF/yr)[3]. Refer to Section 7, Cost/Benefit Analysis, for forecasted volumes of produced water. The HERO®/BC 3 process combination would recover 95.3 percent of the produced water and average life-of-project flow rates would be 909

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[1] Additional treatment would be required at the Collection Center in Bloomfield for oil and grit removal.

[2] Produced water volume would include produced water from the Tri-City, Fairway and Close-in areas, cooling tower blowdown from Prax Air, mine water from BHP Billiton and 100 gpm of Purge Water from the SO_2 absorbers.

[3] Based on 75 to 85 percent recovery of the produced water resource in the Tri-City, Fairway and Close-in areas, 6 percent compound declination of the resource and a project life of 20 years. A mid-range recovery of 80 percent was selected for this analysis.

gpm of HERO® permeate and 144 gpm of BC 3 distillate for a total of 1,053 gpm of reusable water.

6.3 Constituents of Concern

Table 6-1 presents the treated produced water chemistry. Treatment chemistry information can be found in Appendix B, Table B.2, Alternative 10. In addition to the blend of the two streams, permeate from the HERO® process and distillate from BC 3 are treated as separate sources of reusable produced water in this analysis. San Juan River water chemistry and differences between permeate and river water and distillate and river water are also shown in Table 6-1.

Table 6-1
Comparison of Chemistry of Treated Produced Water and San Juan River Water

Comparative Chemistry
Treated Produced Water and San Juan River
PNM - Produced Water Project - SJGS

		HERO Permeate	Diff from San Juan River	BC Distillate	Diff from San Juan River	Total Treated Water	San Juan River (1)
Flow Rate, gpm		909		144		1,053	
Na	mg/l	82.4	+53.4	3.94	-25.1	71.7	29
K	mg/l	2.48	-0.5	0.00	-3.0	2.14	3
Ca	mg/l	0.00	-54.0	0.00	-54.0	0.00	54
Mg	mg/l	0.00	-11.0	0.00	-11.0	0.00	11
Ba	mg/l	0.00		0.00		0.00	ATL (2)
Sr	mg/l	0.00		0.00		0.00	ATL
Fe	mg/l	0.00		0.00		0.00	AND (2)
HCO_3	mg/l	1.26	-123.7	0.00	-125.0	1.09	125
CO_3	mg/l	0.02		0.00		0.02	ND
Cl	mg/l	146	+124.4	6.06	-15.9	127	22
Br	mg/l	0.42	+0.4	0.00		0.36	AND
NO_3	mg/l	0.74	+0.7	0.00		0.64	AND
SO_4	mg/l	9.69	-97.3	0.00		8.37	107
Total SiO_2	mg/l	1.01	-11.0	0.00	-12.0	0.87	12
Total NH_3	mg/l_N	14.6	+14.6	0.00		12.6	AND
Total Alk	mg/l_{CaCO3}	1.09	-101.4	0.00	-102.5	0.94	102
B	mg/l_B	0.62	+0.6	Trace (3)		0.54	ATL
o-PO_4	mg/l_P	0.00		0.00		0.00	AND
TDS	mg/l	267	-93.4	10	-350	232	360
pH		8.64		7.00		8.42	8.00

Notes…..

1. Chemistry provided by SJGS.
2. ATL = assumed trace levels. AND = assumed non-detectable levels.
3. Possible trace levels of boron in BC 3 distillate.

Relative to San Juan River water, four constituents in HERO® permeate are at notably higher levels – sodium (Na^{+1}), chloride (Cl^{-1}), ammonia (NH_3) and boron (B) – and five constituents are at lower levels – calcium (Ca^{+2}), magnesium (Mg^{+2}), carbonate alkalinity (primarily $HCO3^{-1}$), sulfate (SO_4^{-2}) and silica (SiO_2).

BC 3 distillate would have a TDS of 10 mg/l (likely 1 to 2 mg/l), but could have trace levels of boron. If BC 3 distillate were used for boiler feedwater, boron deposition could pose problems[4]. BCs 4 and 5 already generate more water than the boilers can use. Therefore boiler feedwater was not considered as a possible use for BC 3 distillate in this analysis.

6.4 HERO® Permeate

Table 6.2 presents a summary of major process water users at SJGS, their freshwater demand and possible constituents of concern found in HERO® permeate. San Juan River water is fed to the ash system for bottom ash sluicing and fly ash wetting, the cooling towers for make-up and the absorbers via limestone preparation for make-up. Refer to Section 5, Figure 5-1. HERO® permeate compatibility is discussed next for each system.

Table 6-2
Process Water Users at SJGS and Potential Reuse Concerns of HERO® Permeate

Process Area	Water Uses	San Juan River Demand	Water Reuse Concerns
Ash Systems	Bottom Ash Sluicing Fly Ash Wetting	100 gpm	TDS – none Chloride – none NH_3 – none
Cooling Towers	Make-up	12,480 gpm	Chloride – none NH_3 – potential stress cracking of condenser tubes
SO_2 Absorbers	Make-up to Limestone Prep	1,210 gpm	Chloride – somewhat higher than San Juan River NH_3 – none

6.4.1 Ash System

In Section 4, it was determined that untreated produced water might cause corrosion problems in the bottom ash system because if its high TDS and chloride content. Also, if used for wetting fly

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[4] SJGS has linked borate deposition on steam turbine blades to trace levels of boron in BC 4 and 5 distillate. Boron in the feedwater to the BCs must be kept below 1 mg/l to minimize this problem. Produced water boron levels in HERO® reject to BC 3 would exceed 60 mg/l.

ash, overspray could flow to the Process Ponds[5] (via plant drains), thereby raising the chloride concentration in the feed to BCs 4 and 5.

The TDS of the permeate is projected to be less than that of San Juan River water, so corrosion from high salt content would likely not be an issue. The chloride content of the permeate, while higher than that of San Juan River water, is more than an order of magnitude less than untreated produced water. Therefore, releases to the Process Ponds should not be a concern.

Ammonia (NH_3) is quite high in the permeate, however, it is compatible with the ferrous metals found in the ash system. Ammonia is also found in the flue gas, and as such, is likely a constituent in ash water. If released to the Process Ponds (from over-spraying fly ash), BCs 4 and 5 would remove it[6].

6.4.2 Cooling Tower – Ammonia

The condenser tubes for all four units at SJGS are admiralty brass, which is especially susceptible to ammonia attack[7]. Prolonged exposure to ammonia at concentrations greater than 2 mg/l_N[8] will cause stress corrosion cracking. The ammonia attacks the metal at the grain boundaries in areas where there is stress[9]. Microscopic cracks form at the surface and propagate into the metal. Eventually, tube failures occur. Presently, ammonia levels in the circulating water at SJGS are usually less than 0.05 mg/l_N.

Refer to Table 6.3 for possible ammonia concentrations in the cooling towers using HERO® permeate. Given the permeate concentration and feed rate, it would appear that cooling tower ammonia levels could rise to 10 mg/l_N. However, cooling tower chlorination using 12 percent sodium hypochlorite (NaOCl)[10] would reduce ammonia concentrations in the circulating water. NaOCl reacts with ammonia to form chloramines – monochloramine (NH_2Cl), dichloramine ($NHCl_2$) and trichloramine (NCl_3). At circulating water pH, NH_2Cl would predominate. Chloramines are used in drinking water supply systems as a biocide, because they have a long-lasting residual in closed (to atmosphere) systems. In cooling towers at SJGS, a significant

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[5] The Process Ponds feed BCs 4 and 5 and the SO_2 absorbers – both systems have strict chloride limits. High levels of chlorides entering the Process Ponds could require increased flows of BC brine and Purge Water.

[6] BCs 4 and 5 are operated at low pH, and as such, ammonia (NH_3) would be converted to ammonium ion (NH_4^{+1}). As an ion, it would be concentrated in the circulating BC brine and sent to the evaporation ponds.

[7] Admiralty brass is susceptible to ammonia stress corrosion cracking. NH_3, NH_4OH (ammonium hydroxide) as well as the ionic form NH_4^{+1} (ammonium) participate in the corrosion mechanism.

[8] Use of Degraded Water Sources as Cooling Water in Power Plants, EPRI and the California Energy Commission, 2003, Technical Report 1005359.

[9] With condensers tubes, stress is usually induced thermally during operation.

[10] 12 percent sodium hypochlorite solution is the same as household bleach, but at twice the concentration, and is the most common biocide used for power plant cooling system disinfection.

fraction of the chloramines would leave the water in the air stream[11]. Therefore, if HERO® permeate were fed to the cooling towers, ammonia levels should be much less than 10.5 mg/l_N. The chloramines that remain in the circulating water would provide disinfection and would theoretically reduce the chlorine demand during disinfection cycles[12]. Lastly, chloramines do not participate in stress corrosion cracking of admiralty brass.

Table 6-3
Possible Ammonia Concentration in Cooling Towers, San Juan Generating Station

	Blend Stream	Flow gpm	Feedwater NH_3 mg/l_N	Cycled (3) NH_3 mg/l_N
HERO® Permeate	**A**	909	14.6	---
San Juan River	**B**	(Note 1)	AND (2)	---
BC 4 & 5 Distillate	**C**	165	ND (2)	---
BC 3 Distillate	**D**	144	ND	---
Blend Streams A + B + C		12,645	1.05	<<10.5
Blend Streams A + B + C + D		12,645	1.05	<<10.5

Notes:

Total cooling tower demand for make-up (4 units) is 12,645 gpm.

AND = assumed non-detectable levels. ND = non-detectable levels.

Cooling towers at SJGS (units 1, 2 and 4) operate at approximately 10 cycles of concentration. Unit 3 operates at seven cycles.

Note that BC 3 distillate would not increase or reduce ammonia concentrations in the cooling tower because, like river water (and BC 4 and 5 distillate), BC 3 distillate would have no detectable levels of ammonia.

Clearly, if permeate is to be used for cooling tower make-up, ammonia must be removed to protect condenser metallurgy. There are several ways to remove ammonia from permeate:

- Use a 2nd Pass RO step to remove ammonia. HERO® permeate pH would be reduced to neutral. At this pH, 99.5 percent of the ammonia would be converted to the ammonium ion (NH_4^{+1}). Refer to Figure 6.1. As an ion, NH_4^{+1} would be easily removed in the 2nd Pass RO step. Reject from the 2nd Pass RO would be sent to BC 3 along with HERO® reject. In this configuration, NH_3 would be stripped in the deaerating section of BC 3, and NH_4^{+1} would leave with the brine, which would be sent to the evaporation ponds. The 2nd Pass RO would

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[11] When NaOCl is diluted in the circulating water, it forms a weak acid, hypochlorous acid (HOCl). HOCl is the byproduct of NaOCl dissolution that disinfects. HOCl is volatile and some of it is also released to the air stream during chlorination.

[12] SJGS continuously chlorinates using 12 percent NaOCl, and maintains a continuous residual in the circulating water system of 0.1 to 0.2 mg/l_{Cl2}.

recover 93+ percent of HERO® permeate and produce 845 gpm of 2nd pass permeate with a TDS of less than 20 mg/l. In this ammonia-removal configuration, BC 3 would have to be operated at a higher flow rate – 235 gpm of HERO® reject and 64 gpm 2nd Pass RO reject for a total of 299 gpm. An additional capital cost of $643,000[13] would be required for the 2nd Pass RO. Annual capital recovery would amount to $63,000 per year[14]. Approximately 12 mg/l of H_2SO_4 would have to be added to reduce the pH to neutral or less. Acid addition for the 2nd Pass RO would cost less than $3,000 per year. Additional power for the 2nd Pass RO operating at 200 psi and increased utilization of BC 3 would amount to $142,000 per year. Annual produced water treatment costs would increase by $208,000. Overall recovery of produced water would be reduced by 1.1 gpm with 2nd Pass RO and increased BC 3 utilization.

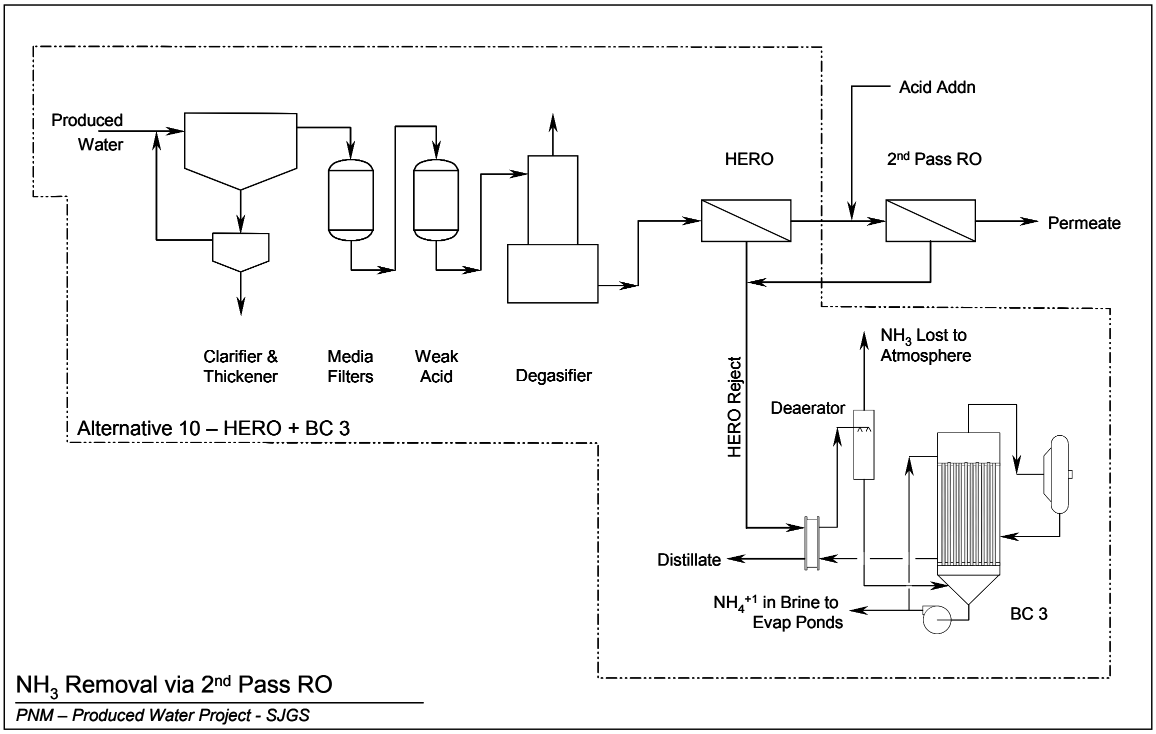

Figure 6-1
Ammonia Removal via 2nd Pass Reverse Osmosis

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[13] Capital cost includes equipment, a 45 percent allowance for installation, 15 percent contingency, 5.5 percent PNM general and administrative costs and 6.125 percent for the New Mexico Gross Receipts Tax.

[14] Capital recovery is based on 7.5 percent interest and paid over a period of 20 years.

- Use breakpoint chlorination to chemically remove the ammonia. To remove ammonia from HERO® permeate, 750 gallons of 12 percent NaOCl solution would be required per day at a cost of $200,000 per year[15]. The chlorine required for biological control[16] in the cooling towers would be reduced because of the sustained presence of chloramines. NaOCl bulk storage, REDOX[17] instrumentation and feed pump equipment for break chlorination would likely cost $50,000. Annual capital recovery would amount to $5,000 per year. The total annual cost of breakpoint chlorination of HERO® permeate to remove ammonia would be $205,000. If a 33 percent credit is applied to the cost of biological control for the cooling towers, the annual cost of breakpoint chlorination would be reduced by $3,500 to $4,500.

6.4.3 Cooling Tower – Chlorides

Chloride levels are a concern because the cooling towers contain 304 stainless steel components – bolts, brackets and other hardware. At concentrations exceeding 1,000 mg/l in the circulating water, chloride can cause stress corrosion cracking of 304 stainless steel components. Stress can be induced at elevated temperature (close to the condenser) or from component loads. Presently, at ten cycles of concentration, the cooling water should not exceed 220 mg/l of chlorides. If 909 gpm of HERO® permeate were added to the cooling tower, chloride levels would rise to 305 mg/l at ten cycles of concentration – well below the 1,000 mg/l threshold.

6.4.4 SO_2 Absorbers

Flue gas contributes a significant amount of chloride and ammonia content to the scrubber liquor in the SO_2 absorbers. In Section 4, it was determined that the absorbers pick up 6.6 tons of HCl per day from the flue gas (85 to 90 percent of the chloride entering the absorbers). The remainder of the chloride intake comes from 1,210 gpm of San Juan River water and 730 gpm of Process Pond Water. Given this intake, if all the permeate were fed to absorbers, the Purge Water stream would have to be increased from 100 to 123 gpm to maintain chloride levels at the control limit of 5,000 mg/l. This additional flow would be treated by the HERO®/BC 3 treatment system and would add an additional 2.1 percent to the operating cost of the system (additional chemicals and power). The cost impact would be $17,000 per year. The treatment system would be designed for a rate of 1,545 gpm to treat produced water during the peak years (also includes 10 percent capacity cushion). Therefore, with a capacity margin of 440 gpm (design minus life-of-project average flow), an additional requirement of 23 gpm would be well within design parameters and would require no additional capital outlays.

The absorbers also pick up ammonia from the flue gas with a scrubber liquor concentration of 27 mg/l_N. Refer to Section 4, Table 4-6. Most of the ammonia is in the NH_4^{+1} form because the

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[15] SJGS pays $0.73 per gallon of 12 percent solution.

[16] With continuous chlorination, the plant uses 40 to 50 gallons per day of 12 percent NaOCl.

[17] REDOX refers to instrumentation that measures oxidation/reduction to determine oxidation residual and control NaOCl feed.

operating pH of the system is less than neutral. There are no apparent corrosion issues involving ammonia in the absorbers so feeding permeate with ammonia should not be a concern.

6.5 BC3 Distillate

BC distillate is characterized by having low TDS – Table 6.1 shows a TDS of 10 mg/l, but in practice, TDS is usually less than 3 mg/l. This water could be used in any of the processes discussed previously – ash system, cooling towers and SO_2 absorbers.

6.6 HERO® Permeate and BC 3 Distillate Blend

The differences in chemistries between HERO® permeate and distillate are significant. Therefore if the streams were blended, the product would resemble permeate at concentrations that were 20 percent lower. However, the same amounts of chloride and ammonia would be entering the cooling towers and absorbers, so similar treatment quantities and associated costs would apply.

Table 6-4
HERO® Permeate Compatibility--Cost Summary

	Improvements Required to Use HERO® Permeate	HERO® Permeate Use, gpm	Additional Capital Improvements	Additional Annual Op Cost (1)
Ash System	None	100	$0	$0
Cooling Towers	2nd Pass RO	909	$643,000	$208,000
	Breakpoint chlorination	909	$50,000	$201,000
SO_2 Absorbers	Increased Purge Water Rate	909	$0	$17,000

Notes:

Includes capital recovery at 7.5 percent for 20 years.

7
COST/BENEFIT ANALYSIS

7.1 Introduction

The costs and benefits of gathering, conveying and treating produced water for use at San Juan Generating Station (SJGS) are presented and assessed in this section of the report. Life-of-project projections are developed for the produced water resource in the Study Area and a number of scenarios are assessed to determine reasonable recoverable volumes of water. A likely range of produced water recovery was established to estimate capital and operating costs for the project. Public Service of New Mexico (PNM) and producer revenue sharing, in the form of reduced produced water disposal costs, is also incorporated into the economic analysis to determine life-of-project water costs.

7.1.1 Project Setting

There is minimal gathering infrastructure in place in the San Juan Basin. Almost all of the gathering is accomplished by transporting produced water by tanker truck from wellhead to a SWD (salt water disposal facility) for disposal via deep well injection[1]. Also, oil and gas production is highly dispersed – one well per 160 to 320 acres. Recently, Bureau of Land Management (BLM) permitted infill drilling to allow one well every 80 acres on BLM land. Production in 2003 generated 45,240 BPD of water in the Study Area[2] which covers 1,500 square miles (38 townships).

A handful of energy companies represent the majority of production in the San Juan Basin. Seven producers (large and small) represent almost 95 percent of produced water generation in the Tri-City, Fairway and Close-in areas. PNM has discussed the produced water project (in varying degrees) with four of these producers.

The San Juan Basin is currently experiencing a period of accelerated development because of increased demands for natural gas, and new well installation in the region is currently limited by

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[1] Based on 2003 OCD (Oil Conservation Division of New Mexico) production data, there were 44 injection wells in the Study area that were used solely for produced water disposal. Of the 44 wells, one took industrial wastewater in addition to produced water.

[2] The Study Area at this point in the report has been reduced from 2,400 square miles as described in Section 2, *Assessment of Produced Water from Oil and Gas Wells*, to 1,500 square miles to focus on high-volume areas of produced water generation. The Study Area is delineated by townships – 32N5W (northeast corner) to 29N14W (southwest corner) and encompasses the Tri-City, Fairway and Close-in areas identified in Section 3, *Produced-Water Infrastructure Availability and Transportation Analysis*, Figure 3-3.

the availability of drilling equipment. In 2003, the San Juan Basin had 19,090 active wells – 8,500 in the Study Area (almost all of which are gas wells). Also, as a result of infill drilling, produced water injection in the Study Area increased 26 percent from 2002 to 2003 and by 34 percent in the Tri-City, Fairway and Close-in areas.

7.1.2 The Need to Work with Producers

The lack of infrastructure and the size of the Study Area make gathering and delivering produced water to SJGS costly. The cost/benefit analysis recognizes this and incorporates producer involvement on the gathering side to reduce PNM's cost exposure to a collection point, pipeline and treatment plant. Producers would provide gathering infrastructure to deliver water to either the Collection Center in Bloomfield or along the pipeline. In doing so, producers would benefit by minimizing their disposal costs. The PNM-producer relationship is structured in this analysis to provide financial benefits to PNM and producers that materially participate[3].

Gathering strategies for the Tri-City and Fairway areas and Close-in production are discussed next.

Gathering Produced Water in the Tri-City and Fairway Areas

The Tri-City and Fairway areas present the greatest challenge to gathering produced water. Burlington Resources (BR) has an extensive production network in the Study Area with existing infrastructure that could be modified for gathering purposes (discussed in Section 3, *Produced-Water Infrastructure Availability and Transportation Analysis*). Produced water gathering would involve BR and PNM and would be segmented into following areas of responsibility:

- BR would build infrastructure by modifying the Hart Canyon Line and CO_2 Gas Line to gather produced water in the Tri-City and Fairway areas.
- BR would deliver the gathered water via an extension of either the Hart Canyon Line or CO_2 Gas Line to the PNM Collection Center in Bloomfield.
- BR would build satellite collection stations along the Hart Canyon Line and CO_2 Gas Line to receive-for-fee produced water from other producers.
- PNM would build a Collection Center at the headworks of the pipeline to receive and pretreat[4] gathered water in the Tri-City and Fairway areas.
- PNM would convey gathered water to SJGS for treatment and use.

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[3] The PNM-producer project relationship presented in this section of the report was developed with three producers.

[4] Pretreatment at the Bloomfield Collection Center would consist of oil and grit removal via gravity separation, flotation and media filtration and is discussed in detail in Section 4, *Produced Water Use, Treatment, and Disposal Analysis.*

The investment in BR gathering infrastructure[5] would be covered by their avoided costs of disposal and fees generated by the receiving water from other producers. BR would share with PNM:

- Avoided costs of disposal of BR produced water
- Fees from other producers for receiving produced water
- BR's avoided costs associated with building new or replacement injection wells and injection well facilities (SWDs).

Close-in Gathering

Close-in producers – Dugan Production Corporation (Dugan) and Richardson Operations Company (Richardson) – would inject filtered produced water directly into the conveyance line. Dugan and Richardson would share with PNM cost savings associated with avoided disposal of produced water (via deep well injection).

The PNM share of BR, Dugan and Richardson avoided costs and fees would be treated as project revenue against the cost of conveyance and treatment of produced water.

7.1.3 Legislative Initiatives

PNM endeavored to address regulatory issues associated with produced water by supporting a bill in the New Mexico legislature that would specifically allow the disposal of produced water at electric generating facilities. The bill consisted of two elements:

- It would allow producers to dispose of produced water at SJGS. This would eliminate a number of regulatory and jurisdictional problems associated with beneficial use of a water resource.
- Acknowledging the high cost of this project, PNM would receive a tax credit from the state to compensate for the cost of conveying and treating produced water. The amount of the proposed tax credit would be $1,000 per acre-foot (AF) of produced water delivered to SJGS. The credit would be limited to $3 million annually. Also, there would be a maximum payable life-of-project cap equal to 50 percent of the capital cost of the project.

The bill was introduced into the January-February 2004 state legislative session and the provision allowing produced water disposal at electric generation facilities such as SJGS was signed into law March 2004. The tax credit was not included in the bill. PNM plans to support tax credit legislation in the next state session in 2005.

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[5] Burlington Resources developed a cost analysis (with PNM) to determine the economic benefits of a gathering system owned and operated by them.

7.1.4 Benefit of the Project to PNM

Power generation is directly proportional to water supply at SJGS, e.g. a five percent reduction in annual water supply would result in a five percent reduction in annual power generation. As discussed in Section 4, Produced Water Use, Treatment, and Disposal Analysis, SJGS treats and recycles a significant amount of its wastewater, and as a result, there are no opportunities for additional water savings to preserve generation capacity. Climate studies conducted by researchers at the University of Arizona (Cavazos et al, 2002) predict that New Mexico is entering a period of extended drought – possibly lasting 60 to 80 years (wet-to-dry-to-wet cycle). For the past two years, water supplies in the Four Corners area have been strained and the plant has guaranteed its supply through one-year purchase agreements with local tribal entities. At some point in the future, these agreements may not be obtained because of dramatically reduced regional water supply.

SJGS has a take-or-pay coal contract, i.e. a fixed amount is paid for fuel annually whether it is used or not. If the plant has to reduce load for significant periods of time because of reduced water supply and if the reduction in load is large enough, PNM must still pay for fuel. Since fuel is the largest expense for SJGS, this is considered a credible worst-case economic scenario given the strong inevitability of drought. PNM has determined that a one-year 30-percent shortage in regional water supply would be significantly more costly in fuel contract penalties and lost generation than the entire capital investment in the produced water project[6].

Produced water from the Tri-City, Fairway and Close-in areas could supply 8.8 to 10.0 percent of SJGS's needs and could prevent SJGS from reaching the take-or-pay coal contract threshold.

7.1.5 Evaluation Basis

Five produced water cases are assessed in this section of the report. The cases represent a range of the produced water recovery in the Tri-City, Fairway and Close-in areas – from 50 to 90 percent (in 10-percent increments) of total produced water generated in the Study Area. Also, as oil and gas fields mature, produced water generation will decline (especially true for CBM production). Three declination scenarios – two, four and six percent – were evaluated for each produced water recovery case for a total of fifteen economic assessments. The assessments are used to evaluate a range of project economics to cover the uncertainty associated with supply.

7.2 Life-of-Project Produced Water Generation

The project would have an operating life of 20 years. The first five years would represent the development of producer-side infrastructure to gather BR water and attract other producers. It is anticipated that BR, Dugan and Richardson would participate at the outset of the project.

The success of the project is highly dependent on the development and utilization of a well-designed produced water gathering system. A system designed to minimize transportation time from wellhead to disposal would attract producers because it could significantly reduce their

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[6] PNM preferred that the fuel penalty contract information not be included in this report.

operating costs. Trucking produced water represents 50 to 80 percent of the disposal costs for many producers. BR would develop and operate the gathering side of the project by extending the Hart Canyon Line and CO_2 Gas Line to a common point and on to the Collection Center in Bloomfield.

BR would install satellite collection stations along both lines at the intersections of heavily traveled disposal truck routes. BR would utilize the stations to reduce their hauling costs and offer the service (for fee) to others as a more cost-effective disposal option. The Hart Canyon Line and CO_2 Gas Line will also be designed to take direct (piped in) deliveries from SWDs. There are four SWDs immediately near the CO_2 Gas Line alignment. Direct disposal to the gathering system would extend SWD injection well life, and in many cases, could eliminate to need to replace wells[7]. Refer to Figure 7.1 for an overall project schematic (showing areas of project responsibility). For additional water, many SWDs could be retrofitted with a well pump to backflow previously-injected produced water.

Dugan and Richardson would inject filtered produced water directly into the 28.5-mile pipeline (just east of SJGS). Prax Air and BHP Billiton would inject cooling tower blowdown and mine water, respectively, into the line in the same vicinity. Eventually, produced water gathering would likely involve seven or more producers (large and small).

The sources of water would include:

- Tri-City and Fairway produced water gathered via the Hart Canyon Line and CO_2 Gas Line using satellite collection stations and accepting direct flow from SWDs.
- Backflow water (from retrofitted SWDs) gathered via the Hart Canyon Line and CO_2 Gas Line.
- Close-in produced water
- Cooling tower blowdown from Prax Air
- BHP Billiton coal mine water
- Purge Water from the SO_2 absorbers at SJGS (collected onsite at the plant)

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[7] SWD injection wells can last from three to as many as ten years. Most last five years and then must be replaced at significant cost. Also, injection well maintenance can be costly if a well experiences problems such as plugging or wall failures.

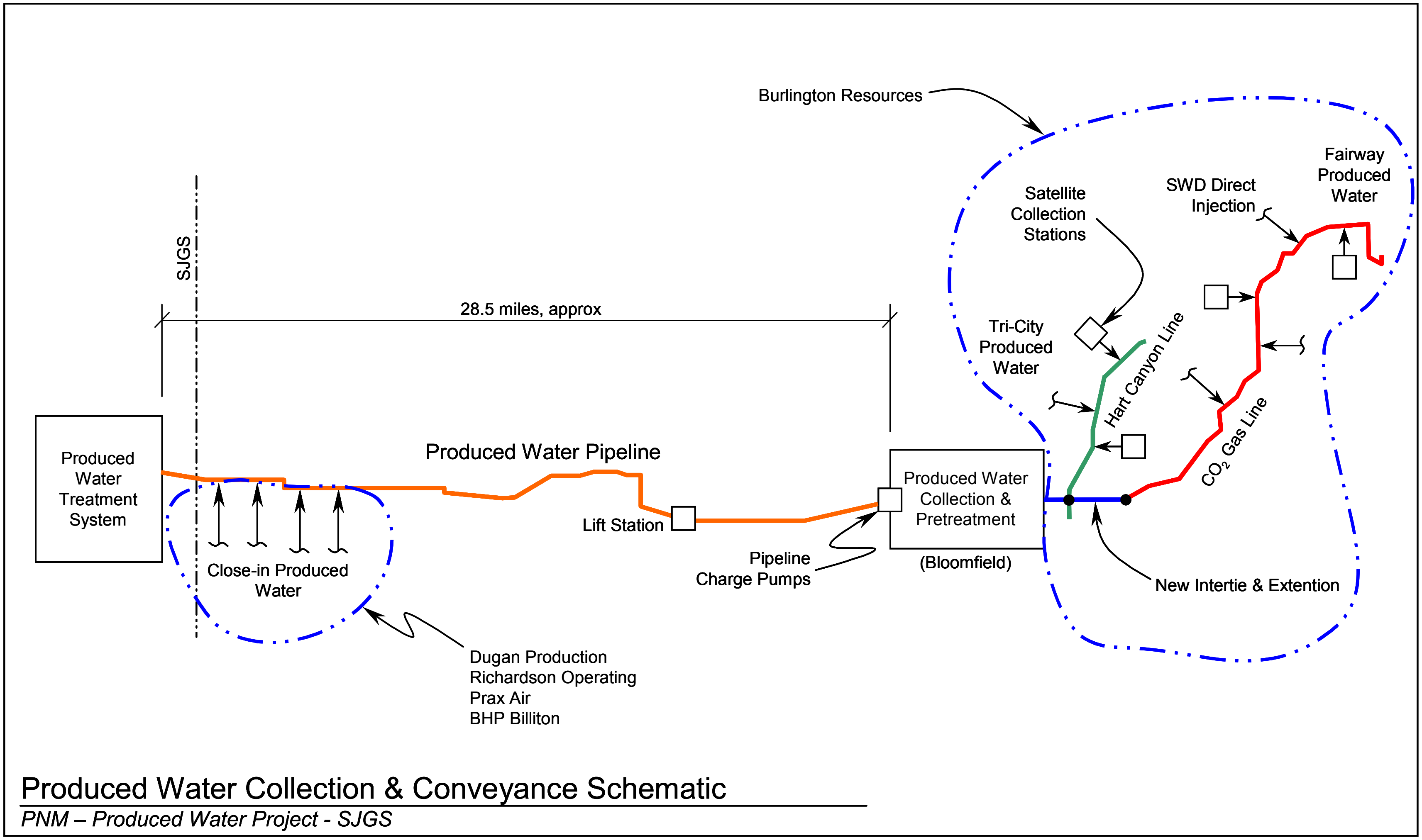

Figure 7-1
Produced Water Collection and Conveyance Schematic

7.2.1 Study Area Resource Estimates

The San Juan Basin is currently experiencing a period of increased development because of market demands for natural gas. It was assumed that the first five years of the project would see growth in gas development (and as well as increased producer involvement in the project). After that, as resources decline, produced water generation would fall. Daily produced water generation for Fairway, Tri-City and Close-in areas is based on 2003 production data (refer to Figure 3-4 in Section 3). These values are escalated by three percent through 2006 to reflect growth in production (new wells). There was actually a 34.7 percent increase of produced water from 2002 to 2003 in the Fairway, Tri-City and Close-in areas. This increase was a result of infill drilling to meet the increased demand for natural gas. The three-percent escalation factor was used because infill drilling will not proceed at this pace in the long term. Also, since there is uncertainty in predicting water production (especially CBM), a three-percent escalation factor was considered a more conservative approach to planning.

For the purposes of this analysis, this would establish a project start date in 2006.

The following table presents assumed produced water generation for each area:

Production Area	Produced Water 2003 BPD	Produced Water 2006 BPD
Fairway	20,680	22,600
Tri-City	2,760	3,020
Close-in	12,520	13,680
Total	35,960	39,300

In addition to the above estimates, it was also assumed that a total of 10,000 BPD could be extracted (backflow) from formations currently or previously used for deep well disposal of produced water. This would likely require the retrofitting of three to four SWD injection wells with pumps.

The total water resource for the Study Area is a combination of produced water from the Fairway, Tri-City, and Close-in production areas, backflow from three to four SWD wells and other non-production sources of water – cooling tower blowdown from Prax Air, BHP Billiton mine water and SO_2 absorber Purge Water. Refer to Table 7-1 for a summary of the possible resource in 2006 (expressed in three different units of measure).

Table 7-1
Total Water Resource - 2006

	BPD	gpm	AF/yr
Fairway	22,600	659	1060
Tri-City	3,020	88	142
Close-in	13,680	399	644
Backflow	10,000	292	470
Total Produced Water	49,300	1,438	2,316
Prax Air – Cooling Tower Blowdown	300	9	14
BHP Billiton – Mine Water	1,700	50	80
Purge Water – SO_2 Absorber Bleed Stream	3,430	100	161
Total Other Water	5,430	159	255
Total Water Resource	54,730	1,597	2,571

Table 7-1 represents all of the water in the resource plus water from other sources previously discussed. Produced water recovery, however, would be limited to effective infrastructure gathering improvements. Other water (non-produced water) can be obtained with much less effort.

Life-of-project recoverable water will be dependent on initial sustained growth as a result of infill well installation followed by a gradual decline in produced water generation as fields mature. In this analysis, it is assumed that growth is sustained at two percent per year until 2008 (five years of growth from expanded production starting in 2004). Three declination scenarios are evaluated – two, four and six percent[8]. Declination is based on a compounding formula as follows:

$$DF = (1 - r)^{n-1}$$

Where: DF = *Declination Factor*

r = *Declination (expressed as percent)*

n = *Year*

Refer to Figure 7.2 for declination rate versus time assumptions for the three scenarios and Figure 7.3 for the total water resource versus time used for each scenario. The water resource

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[8] Declination is difficult to predict because producing formations (including coal seams) have varied water-release characteristics.

increases to 2,700 AF/yr in 2010 and falls to 2,200, 1,900 and 1,600 AF/yr, respectively for Scenarios 1 (2% decline), 2 (4% decline) and 3 (6% decline).

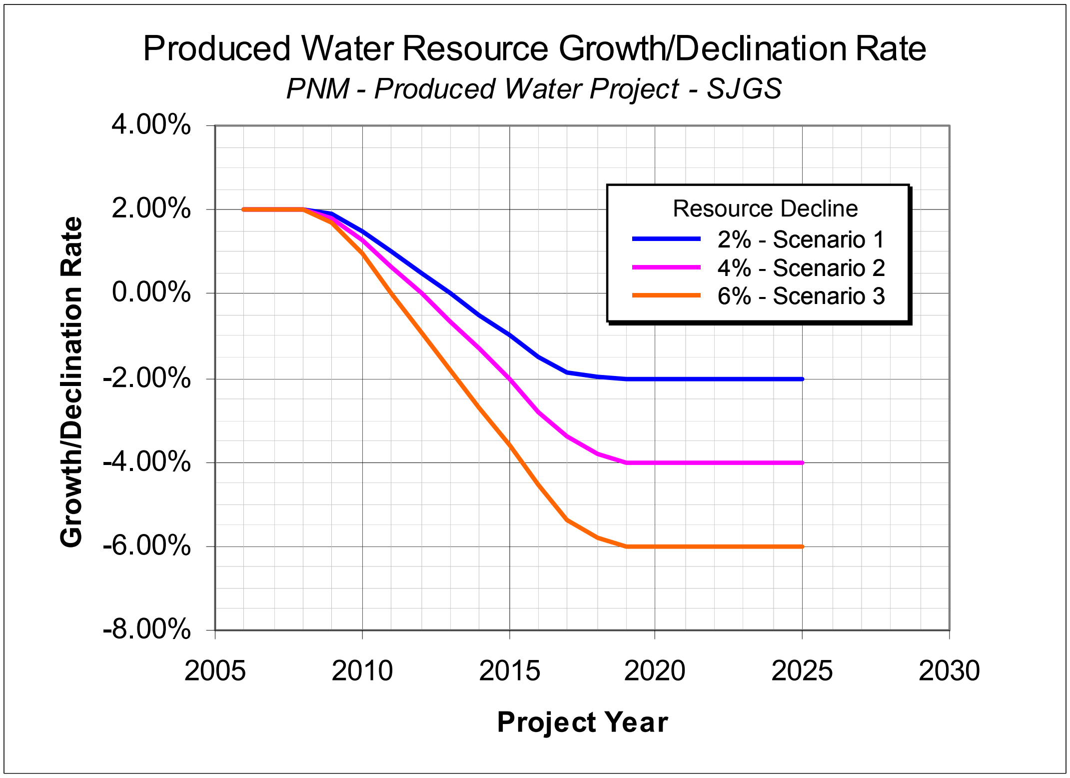

Figure 7-2
Produced Water Resource Growth/Declination Rate

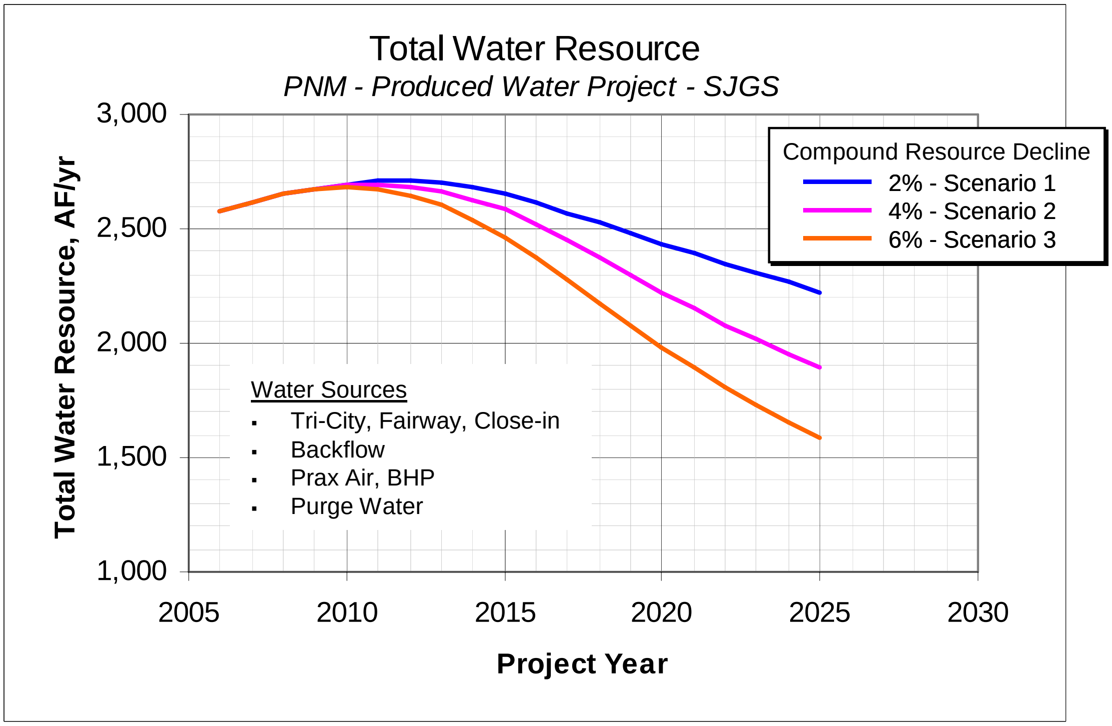

Figure 7-3
Total Produced Water Resource

Depending on the declination scenario, the water resource as defined in Table 7-1 could lose up to 40 percent of its capacity in 25 years. Life-of-project water resource predictions for each declination scenario can also be found in Table D-1 in Appendix D.

Backflow decline (backflow is included in the above resource predictions) was adjusted differently, because previously-injected water is independent of the current infill expansion. Backflow capacity was reduced by two percent per year starting in the fourth project year. Lastly, calculations used to develop resource capacity assume the supply of water from Prax Air, BHP Billiton and SO_2 absorber (Purge Water) remains constant throughout the life of the project.

Again, Figure 7-3 represents all the water in the resource. Recoverable water, which is a function of gathering efficiency, is discussed next.

7.2.2 Recoverable Volume Estimates

At the outset of project implementation, it is assumed that recoverable volume would be limited to produced water provided by BR, Dugan and Richardson. Other producers would likely participate in the project shortly after implementation, and recoverable volume would increase accordingly using the infrastructure developed by BR.

In all cases, the first year starts with half of the potential recoverable daily volume, followed by rapid growth in the second and third years, then slowly peaking at the fifth year. Refer to Figure

7-4 for water recovery cases and declination scenarios[9]. After five years, the fraction of recoverable water levels off to 50, 60, 70, 80 and 90 percent of the total resource, respectively for Cases 1 through 5.

Year-to-year volume recovery for each declination scenario (five recovery cases per declination scenario) can be found in Table D-2 in Appendix D.

7.2.3 Likely Recovery

Recovery of produced water in the Study Area (38 townships north of Highway 64) will be a function of the design of the gathering system and the degree of producer involvement.

First Year of Operation

During the first year of the project, 46 to 49 percent of the produced water in the Study Area – 17,380 to 18,380 BPD – could be recovered. Refer to Table 7-2. This would include BR, Dugan and Richardson produced water as well as 2,000 to 3,000 BPD from a major producer with an SWD next to the CO_2 Gas Line (there are four SWDs immediate to the CO_2 Gas Line). BR would also backflow the McGrath SWD injection well for an additional 3,000 to 5,000 BPD. Prax Air, BHP Billiton and Purge Water from SJGS would generate an additional 5,430 BPD of water. During the first year of operation, 25,810 to 28,810 BPD of water (1,214 to 1,355 AF/yr) would likely be delivered to SJGS. Refer again to Figure 7.4. After treatment at SJGS, 95.3 percent of the recovered water – 1,161 to 1,295 AF/yr – would be made available for reuse at the plant.

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[9] Annual recoverable water in Figure 7-4 represents water delivered to SJGS.

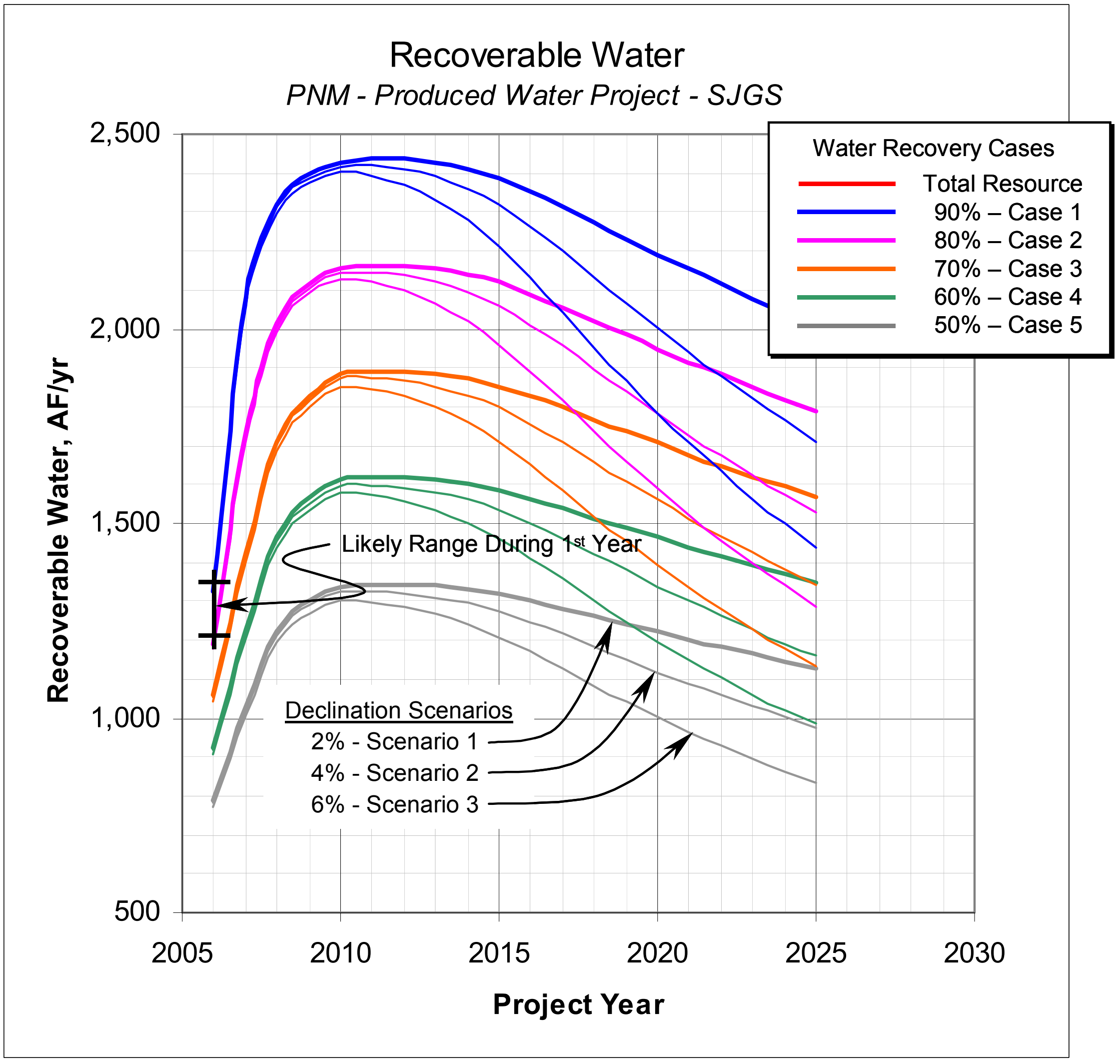

Figure 7-4
Recoverable Produced Water Resource

Long Term Produced Water Recovery

In 2003, seven producers – BR, Dugan, Richardson, BP America, Conoco Phillips Company, Williams Production Company and XTO Energy Inc. – generated:

- 89 percent (40,150 BPD) of the produced water in the Study Area
- 95 percent (34,280 BPD) of the produced water in the Tri-City, Fairway and Close-in areas

Dugan and Richardson generated almost all of the produced water in the Close-in area in 2003.

Table 7-2
Likely Produced Water Recovery During First Year - 2006

Produced Water Resource (1)

Study Area – All Producers (3)	49,450 BPD
Tri-City, Fairway, Close-in – All Producers	43,870 BPD
Tri-City, Fairway, Close-in – Seven Largest Producers	37,470 BPD

Produced Water Project

Tri-City, Fairway, Close-in – BR, Dugan & Richardson	15,380 BPD
Direct Feed from SWD	2,000 – 3,000 BPD
Subtotal	17,380 – 18,380 BPD
Fraction of Resource at Start Up	46.4% – 49.1%
McGrath Backflow	3,000 – 5,000 BPD
Prax Air, BHP Billiton, Purge Water	5,430 BPD
Total Likely Flow at Start Up	25,810 – 28,810 BPD
First Year Delivery to SJGS	1,214 – 1,355 AF/yr
First Year Treated Water for Reuse at SJGS (2)	1,161 – 1,295 AF/yr

Notes:

Produced water volumes are escalated 3% annually from 2003 to 2006.

Treatment at SJGS would recover 95.3% of feed water.

Production from 38 townships north of Highway 64. Refer to Figure 3-3.

One producer will likely direct feed to the CO_2 Gas Line at the outset.

The gathering system would likely generate produced water rapidly during the first year of operation, leveling off after five years, and in the eighth year, volumetric decline would start to occur as the fields mature. The gathering system could accelerate and optimize collection because:

- There are only seven producers that generate most of the production in the area.
- The Hart Canyon Line and CO_2 Gas Line would have six to eight satellite collection stations to accept produced water along heavily traveled transportation routes.
- Four SWDs are located next to the CO_2 Gas Line and could provide a significant portion of the produced water resource.

Given the high density of produced water in the Tri-City, Fairway and Close-in areas among only seven producers, it is reasonable to assume that 75 to 85 percent of the water resource could be recoverable in the Study Area. Recall that a combination of 15 recovery scenarios were evaluated (five recovery cases for each of three declination scenarios). Since the majority of produced water was generated by only seven producers (three were involved in project development), 50 percent recovery seems unrealistically small. Conversely, 90 percent recovery or more seems unrealistically high. A recovery of 75 percent is midway between these two endpoints and would appear to be very likely given volume generation patterns. The 75 to 85 percent range reflects the presumed ability of the gathering system that has been conceptually configured by BR to attract other producers.

For each recovery case, three declination scenarios were evaluated – two, four and six percent. The two-percent declination scenario would result in a 20 percent drop in produced water generation during the life of the project. The four and six percent scenarios would result in a 30 and 40 percent produced water decline, respectively. It was assumed that six-percent declination would be a realistic choice of the three scenarios because:

- Resource decline is the least understood recovery parameter and is difficult to predict. Therefore, a conservative approach was considered essential.
- Currently, the emphasis in the San Juan Basin is on CBM production. CBM wells typically generate water in high volumes early in their life and then drop off more quickly than conventional wells.

Using the above assumptions, Figure 7-5 presents a likely produced water recovery range (delivered to SJGS) over the life of the project.

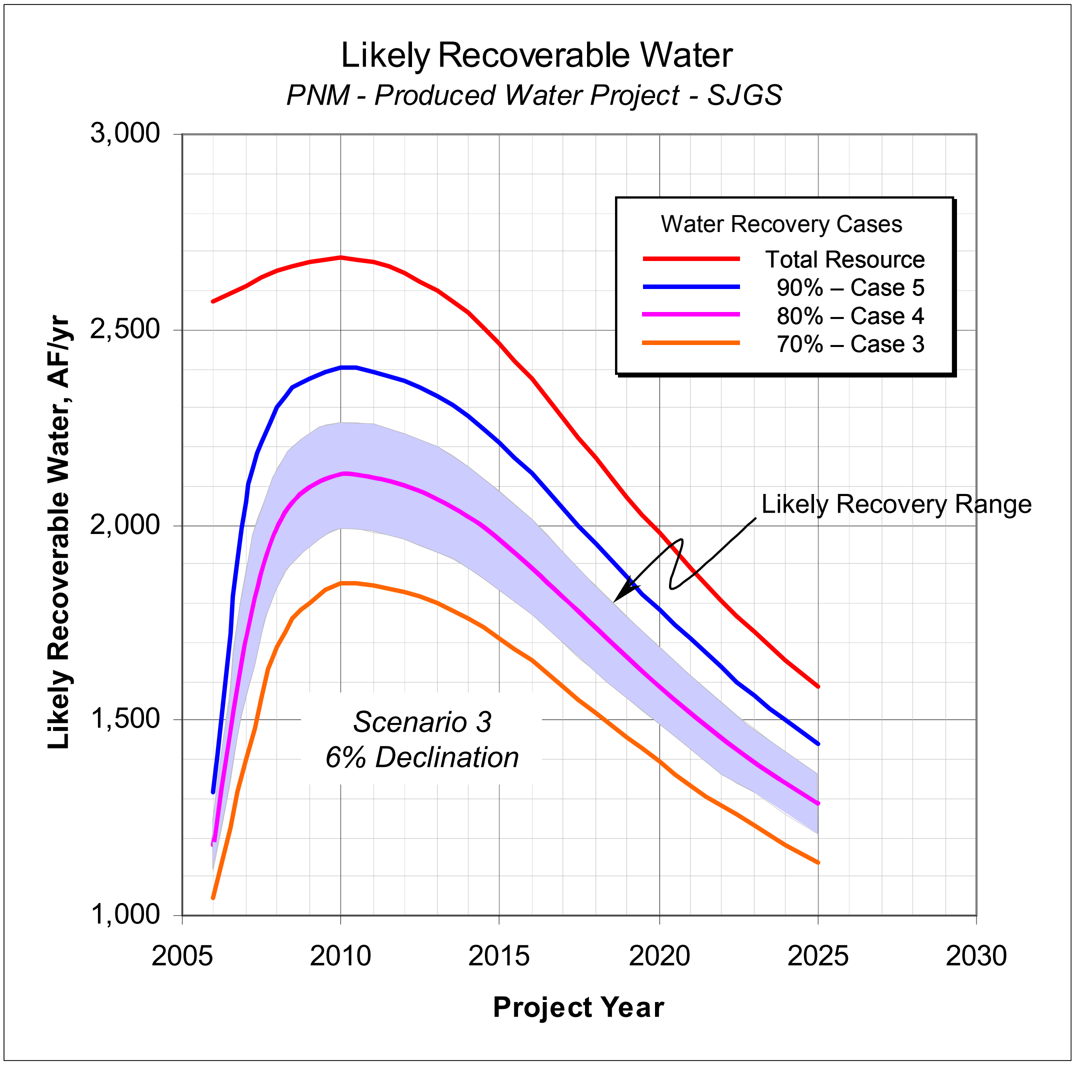

Figure 7-5
Likely Recovery of Produced Water

7.2.4 Project Volume and Revenue

Produced water generated by the project would provide revenue to defray costs associated with gathering, conveying and treating produced water. Project revenue would be realized by reducing the disposal costs of BR, Dugan and Richardson and fees from receiving produced water from other producers. The tax credit (if passed) would also be tied to total produced water recovery[10].

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[10] The tax credit would not include water collected from Prax Air, BHP Billiton or SO2 absorber Purge Water.

Produced water gathering can be grouped into two categories:

- Produced water delivered to the Collection Center in Bloomfield by the gathering system designed, owned and operated by BR.
- Filtered produced water injected directly into the 28.5-mile pipeline by Dugan and Richardson.

The following produced water streams would determine project revenue for each gathering category:

BR Gathering System to Collection Center

BR	Produced water
Other Producers (delivery for fee)	Produced water delivered via satellite collection station
	Produced water fed directly to the Hart Canyon Line or CO_2 Gas Line

28.5-mile Pipeline

Dugan	Filtered produced water fed directly to the pipeline
Richardson	Filtered produced water fed directly to the pipeline

Refer to Figure D-1 in Appendix D for Collection Center volume assumptions and calculations. Also, year-to-year volumes for revenue streams are provided in Table D-3 for each of the five recovery cases and three declination scenarios[11]. Project revenue is discussed later in this section.

7.2.5 Disposition of Off-Spec Produced Water

Occasionally the Collection Center will receive water that cannot be treated, e.g. produced water with very high levels of salinity. There are provisions for holding off-spec water and blending it back into the water leaving the Collection Center, however, there will be occasions when blending is not feasible. Off-spec water will be disposed of at a licensed disposal well in the Bloomfield area. Off-spec water was assumed to be one percent of the volume received at the Collection Center for the first year of operation and tapering off to 0.2 percent by the fourth year as off-spec sources are identified and kept out of the system.

Off-spec water should not to be a problem for Close-in production, because CBM water quality is somewhat constant and there are only trace levels of petroleum byproducts.

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[11] Even though an operating range and declination scenario has been assumed for resource recovery, Table D.3 presents all the cases and scenarios evaluated in this section of the report.

7.3 Capital Cost

There are three categories of capital spending involved in the project:

- Costs incurred by BR to build the gathering system
- Costs incurred by Dugan and Richardson to connect to the 28.5-mile pipeline
- Costs incurred by PNM to build the Collection Center, 28.5-mile pipeline and treatment system at SJGS

7.3.1 Producer Costs

BR estimated that it would cost $5 million to develop the gathering system. This would include:

- Recommissioning the Hart Canyon Line and CO_2 Gas Line including integrity checks and necessary repairs.
- Building six to eight satellite collection stations – each consisting of a receiving tank, transfer pumps, cartridge filters (to remove oil and grit), instruments, valves and piping to either the Hart Canyon Line or CO_2 Gas Line. Each satellite station would use electronic card readers to permit access to a receiving tank for disposal of produced water. This tracking system will also allow BR to identify off-spec sources of water over time.
- Connecting the Hart Canyon Line and CO_2 Gas Line and building an extension to the Collection Center.
- Retrofitting two existing BR SWD injection wells for backflow conversions.
- Providing flange connections and isolation valves for direct-feed of produced water into the Hart Canyon Line and CO_2 Gas Line.

Capital cost details for the gathering system are not included at the request of BR.

Costs for Dugan and Richardson would be minimal and are not estimated[12]. The 28.5-mile pipeline passes both of their operations. They would use existing tanks, filters and pumps and would only have to install several hundred feet of pipeline to intercept the conveyance line.

7.3.2 PNM Costs

PNM capital expenditure would include the Collection Center in Bloomfield, the 28.5-mile pipeline and the produced water treatment system. Refer to Table 7-3 for a summary of capital equipment costs. Produced water would be treated at SJGS using Alternative 10 – the HERO® process and refurbished BC 3. All of the recovered water could be used as supplemental make-up to the cooling towers, SO_2 absorbers and ash system.

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[12] Given the distance to the pipeline and the simplicity of the tie-in, connection costs for Dugan and Richardson are likely less than $100,000 each.

Refer to Section 3 for pipeline details and costs and Section 4 for the Collection Center and the treatment system at SJGS.

Table 7-3
Capital Costs Incurred by PNM

Capital Costs Incurred by PNM

		Collection Center	14-inch Pipeline	HERO + BC 3	Total Project
Capacity, BPD		34,000	60,000	53,000	
Peak Conditions, BPD		30,670	44,710	48,130	
Equipment & Installation		$5,200,000	$12,900,000	$11,800,000	$29,900,000
Contingency	15%	$780,000	$1,940,000	$1,770,000	$4,490,000
NMGRT (1)	6.125%	$320,000	$790,000	$720,000	$1,830,000
PNM G&A (2)	5.5%	$290,000	$710,000	$650,000	$1,650,000
Total Project		$6,590,000	$16,340,000	$14,940,000	$37,870,000

Notes…..
1. NMGRT is the New Mexico Gross Receipts Tax.
2. G&A is a "general and admistrative" charge applied to all PNM projects.

Equipment capacity is based on the maximum treatment throughput that would be experienced by each equipment element during the life of the project (project years four through seven) based on the 75 to 85 percent produced water recovery operating range. The high end of the range – 85 percent – was selected for equipment sizing. Refer to Figure 7-6. The Collection Center and treatment system at SJGS are sized at 34,000 and 53,000 BPD, respectively. During peak recovery periods, this equipment would be operated at 90 percent of rated capacity. Refer to Figure 7-7. The pipeline is sized at 60,000 BPD, and at peak conditions, would be operated at 75 percent of its capacity. The pipeline is considerably oversized to deal with unexpected growth in produced water recovery. With a smaller line and unexpected growth, an additional pipeline (at significant expense) would be required to handle additional flow. Unexpected growth is not a problem with the Collection Center or the treatment system at SJGS, since additional equipment could be added for greater capacity. Lastly, equipment capacity is not optimized and capacity could vary (up or down) after closer analysis.

7.3.3 Total Project Capital Costs

The capital costs associated with the entire project including BR, Dugan, Richardson and PNM are found in Table 7-4. Capital investment would be assumed by each participant in their designated area. Costs include new equipment, upgrades or improvements to existing equipment, one-time right-of-way or land costs, erection and start-up costs.

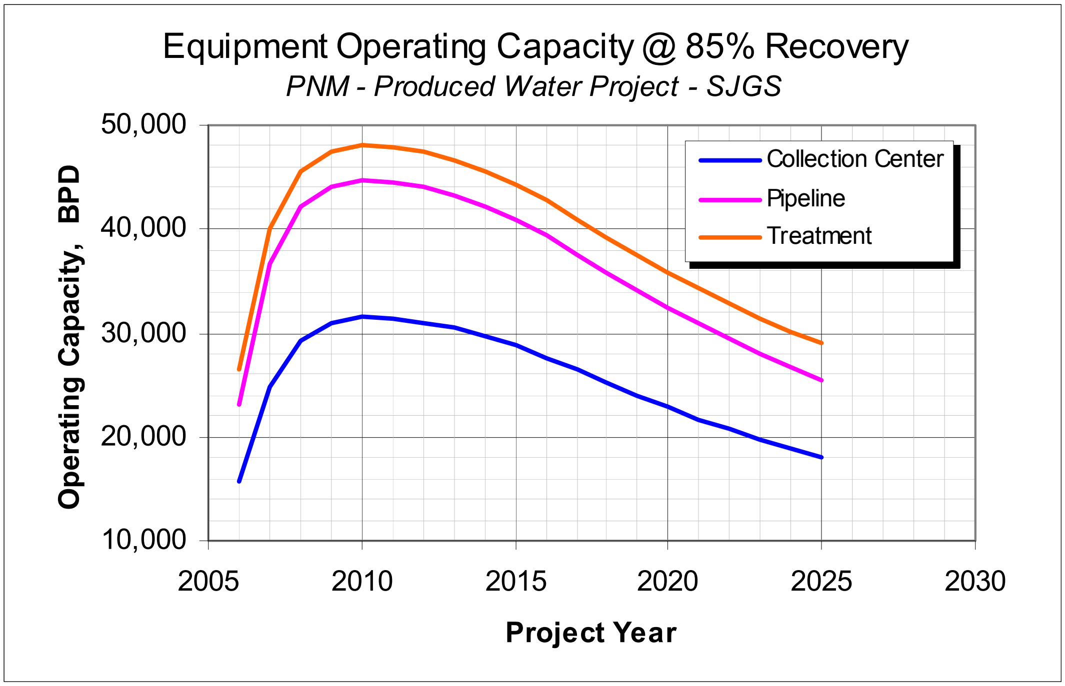

Figure 7-6
Equipment Operating Capacity at 85% Recovery (BPD)

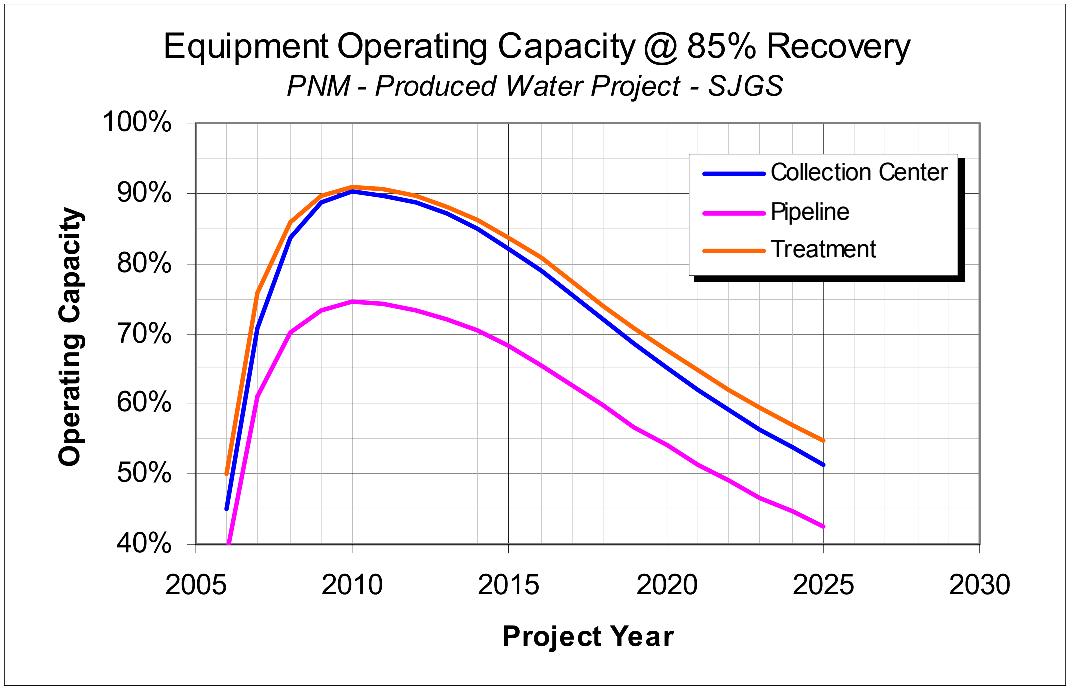

Figure 7-7
Equipment Operating Capacity at 85% Recovery as Percentage of Design Capacity

Table 7-4
Total Project Capital Costs

BR	Gathering system to Collection Center	$5,000,000
Dugan	Inject into pipeline	$100,000
Richardson	Inject into pipeline	$100,000
PNM	Collection Center, pipeline & treatment	$37,900,000
Total Project		$43,100,000

Notes:

Installation costs for Dugan and Richardson are most likely high.

7.4 Operating Costs and Revenues

This section addresses operating costs and revenue. The topics addressed include (1) costs for PNM to operate the Collection Center in Bloomfield, convey produced water to SJGS and treat water for reuse at SJGS; (2) tax credits, project revenue and revenue sharing; (3) project operating costs and revenue adjustments used to determine the life-of-project cost of water for a range of produced water recoveries; (4) BR, and Dugan and Richardson returns on investment for the produced water project as inferred by the analysis[13].

7.4.1 PNM Operating Costs

PNM's operating costs include:

- Chemicals such as sulfuric acid, lime, emulsion breakers, coagulant aids, RO cleaning chemicals, etc.
- Materials include filter media, RO membranes, BC condenser tube inserts, degasifier packing, etc.
- Maintenance – materials and labor for planned and unplanned repairs and contract services such as BC cleaning
- Power to operate equipment
- Off-spec produced water disposal costs – transportation and disposal
- Labor includes PNM operators, maintenance personnel and technicians
- Backflow fees charged by BR and other producers to extract previously injected water from retrofitted SWD disposal wells
- Capital recovery (annual amortization charge assessed to the project to pay for capital equipment).
- Annual rights-of-way payments.

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[13] BR requested that their financial information not be included in this report.

Refer to Table D-1 in Appendix D for volume assumptions and Tables B.4 and B.6 in Appendix B for unit costs for consumables, labor and maintenance assumptions, etc.

Some oil would be recovered at the Collection Center in the gravity separator and sold to the Giant Refinery in Bloomfield. The amount of recoverable oil would be a function of the number of conventional wells that utilize the gathering system for disposal. Based on OCD data, most of the production is CBM (and that fraction is growing). CBM produced water contains trace levels of non-recoverable volatile petroleum byproducts such as benzene, toluene, etc. In time, less and less oil would be recovered as the number of CBM wells increases and the number of conventional wells declines. Therefore, no credit was taken for recovered oil in the operating analysis because of the uncertainty associated with predicting its volume.

Depreciation charges are not included in this analysis. Depreciation is an adjustment to gross income and is used by corporations when determining tax liability. It is designed to encourage new investment by speeding up the recovery of capital invested in a project. Many companies incorporate depreciation into financial analyses, because it reduces corporate expenses in the form of reduced taxes. No attempt was made to include depreciation in this analysis, given the number of entities in the project and the many ways depreciation can be applied. Following the same reasoning, a tax analysis was not performed either, because of the intricacies of tax law and how it can be applied by all parties.

Refer to the Figure 7-8 for calculating "year *n*" life-of-project escalated costs for 75 to 85 percent produced water recovery and six percent declination. Materials and services were escalated annually by 1.93 percent[14] and labor by 2.71 percent[15]. Capital recovery is based on a 7.5 percent interest rate for a payout period of 20 years. Capital recovery is not escalated, rather it is a fixed charge applied annually to the project throughout its life (20 equal payments). Annual capital recovery is calculated as follows:

$$ACR = CI\frac{(1+i_C)^n i_C}{(1+i_C)^n - 1}$$

Where:

ACR = *Annual capital recovery (n equal payments)*

CI = *Capital investment*

N = *Investment payback period, n years*

i_c = *Interest borrowed capital*

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[14] Average annual growth of producer prices of industrial chemicals from 1982 to April 2004, *Chemical Engineering Magazine*, August 2004, Vol. 111, No. 8, page 72.

[15] Average annual growth of hourly earnings in the chemical and oil-related industries from 1992 to April 2004, *Chemical Engineering Magazine*, August 2004, Vol. 111, No. 8, page 72.

Operating costs are calculated using the following relationship:

$$OC_n = (M+S)_n (1+i_{MS})^{n-1} + L(1+i_L)^{n-1} + ACR$$

Where:

OC_n = *Escalated operating cost in year n*
$(M + S)_n$ = *Costs (2006 basis) for materials and services in year n*
$(1 + i_{MS})^{n-1}$ = *Escalation factor for materials and services in year n*
L = *Labor costs (2006 basis)*
$(1 + i_L)^{n-1}$ = *Escalation factor for labor in year n*
ACR = *Annual capital recovery (n equal payments)*

Backflow charges were not included in this portion of the analysis; instead they were deducted from producer revenue (discussed later). Because of escalation, project costs appear to level off in the later years even though volume is significantly reduced. Escalated life-of-project operating costs can also be found in Table D-4 in Appendix D.

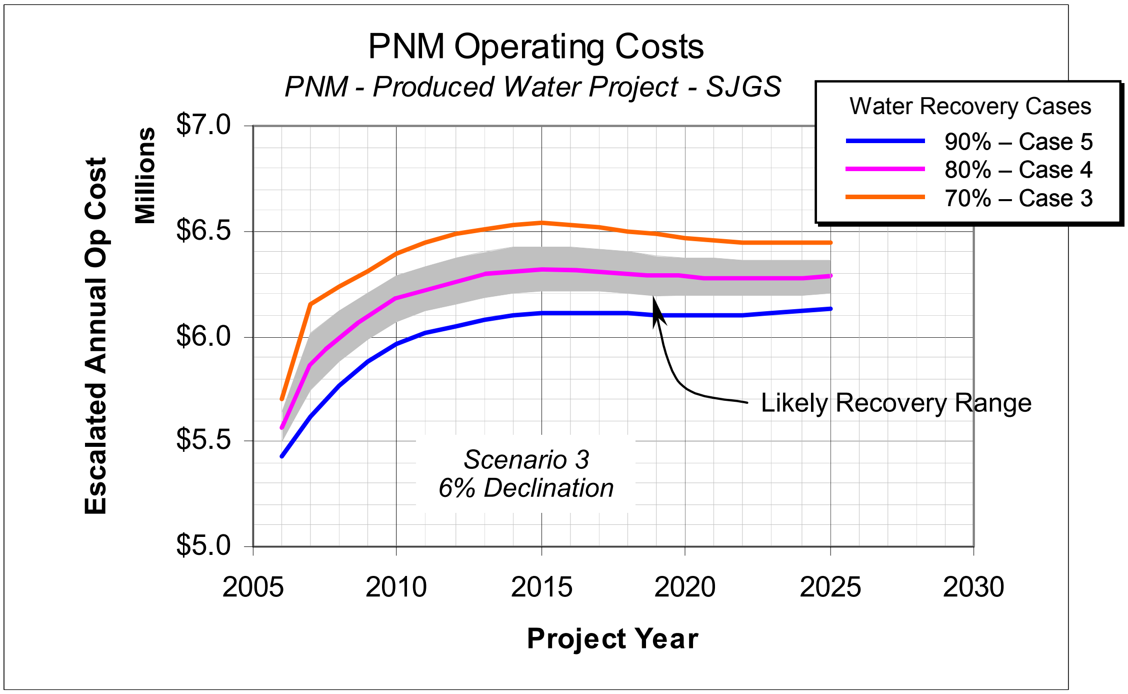

Figure 7-8
Estimated Annual PNM Operating Costs

7.4.2 Tax Credit

If the tax credit were enacted in the 2005 legislative session (in the form proposed in the 2004 session), the following would apply:

- A credit of $1,000/AF of produced water delivered to SJGS
- Credits cannot exceed $3 million annually
- A life-of-the-project cap equal to 50 percent of the capital cost of the project.

Since the capital budget for PNM would be $37,900,000, the life-of-the-project cap would be equal to $18,950,000 (50 percent of the capital budget). Refer to the Figure 7.9 for life-of-project payout of the tax credit for 75 to 85 percent produced water recovery and six percent declination. Lastly, note that the life-of-project tax credit cap would be achieved and would expire in 2017 to 2019 – six to eight years before the end of the project. Life-of-project tax credits can be found in Table D-5 in Appendix D.

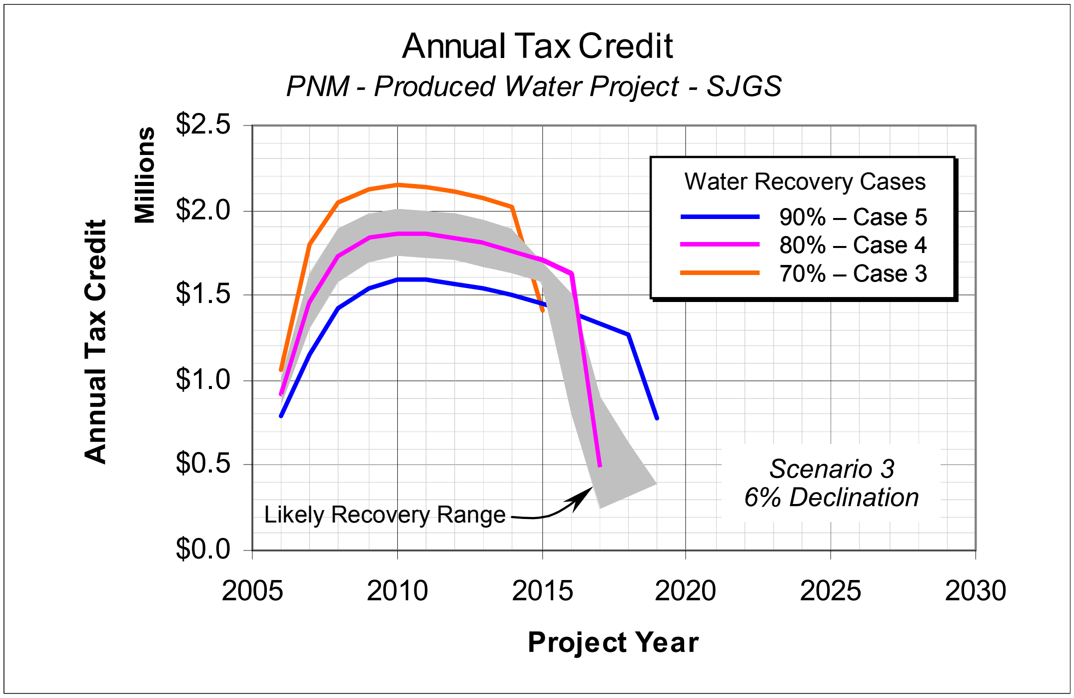

Figure 7-9
Annual Tax Credit

7.4.3 Project Revenues

As discussed previously, project revenues would be generated by BR via their gathering system and Close-in producers Dugan and Richardson. Revenue would come in the form of reduced operating costs, avoided injection well replacement costs and fees from accepting produced water from other producers. Refer to Table 7-5 for the revenue schedule of fees used to calculate project revenue[16].

Year-to-year volumes for revenue streams are provided in Table A-3 for each of the five recovery cases and three declination scenarios.

By sending its produced water to the Collection Center, BR determined it would save $1.2 million per year on injection well replacements. A typical injection well costs between $2 to $3 million to install. One in four wells fail at start-up and have to be abandoned. Depending on the receiving formation and injection rates, wells can last 3 to 10 years (some longer). BR suggested a four-year grace period (to reflect the life cycle of a typical well) before this revenue stream would be implemented. Given the fact that there were 44 injection wells in the Basin in 2003, a significant amount of capital is spent annually by producers to replace (and repair/work over) wells.

Table 7-5
Revenue and Fee Schedule for BR Gathering System and Pipeline

BR Gathering System to Collection Center – Revenue Schedule of Fees

BR	Produced water	$0.55/bbl
	Deferred injection well replacement (starting in year 5)	$1,200,000/year
Other Producers	Produced water delivered via satellite collection station	$0.95/bbl
	Produced water fed directly to the Hart Canyon Line or CO_2 Gas Line	$1.25/bbl

28.5-mile Pipeline – Revenue Schedule of Fees

Dugan	Filtered produced water fed directly to the pipeline	$0.25/bbl
Richardson	Filtered produced water fed directly to the pipeline	$0.25/bbl

As stated previously and for the purposes of this analysis, fees to backflow produced water from retrofitted injection wells are charged against the revenue stream. PNM would be charged a fee of $0.15 per barrel for backflow from BR and $0.25 per barrel for backflow from other producers. Ten cents per barrel was added to the BR unit charge for other producers because they may have to provide more infrastructure than BR to deliver the backflow.

The following relationship is used to calculate year-to-year escalated revenue. Note that the escalation factor for materials and services is used since the basis of revenue is from deferred operating costs and fees for disposal.

$$ER_n = (R_n - BF_n)(1 + i_{MS})^{n-1}$$

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[16] The revenue schedule of fees was developed with BR, Dugan and Richardson.

Where: ER_n = *Escalated revenue in year n*
R_n = *Revenue (2006 basis) in year n*
BF_n = *Backflow charge (2006 basis) in year n*
$(1 + i_{MS})^{n-1}$ = *Escalation factor for materials and services in year n*

Refer to Figure 7-10 for year-to-year escalated total project revenue for 75 to 85 percent produced water recovery and six percent declination. Escalated life-of-project revenue can be found in Table D-6 in Appendix D.

7.4.4 Revenue Sharing and Cost of Recovered Water

Recovering produced water for power generation will benefit PNM by ensuring power generation and avoiding fuel penalties. It will benefit BR, Dugan and Richardson by reducing their costs of operation. The project could generate \$87 to \$99 million in revenue (2006 dollars) over a period of 20 years (assuming 75 to 85 percent produced water recovery, respectively, and six percent resource declination).

The relationship among the PNM/producer group is mutually dependent, i.e. without PNM there would be no project and without the producers there would be no water. Therefore, since the project benefits all parties, all parties should share the revenue. Two levels of revenue sharing are examined:

- 50:50 Split – PNM and the producers would split the revenue evenly.
- 75:25 Split – PNM would receive the greater share.

The 50:50 split is a logical allocation of revenue in a mutually-dependent business relationship, i.e. all parties need each other to generate this revenue.

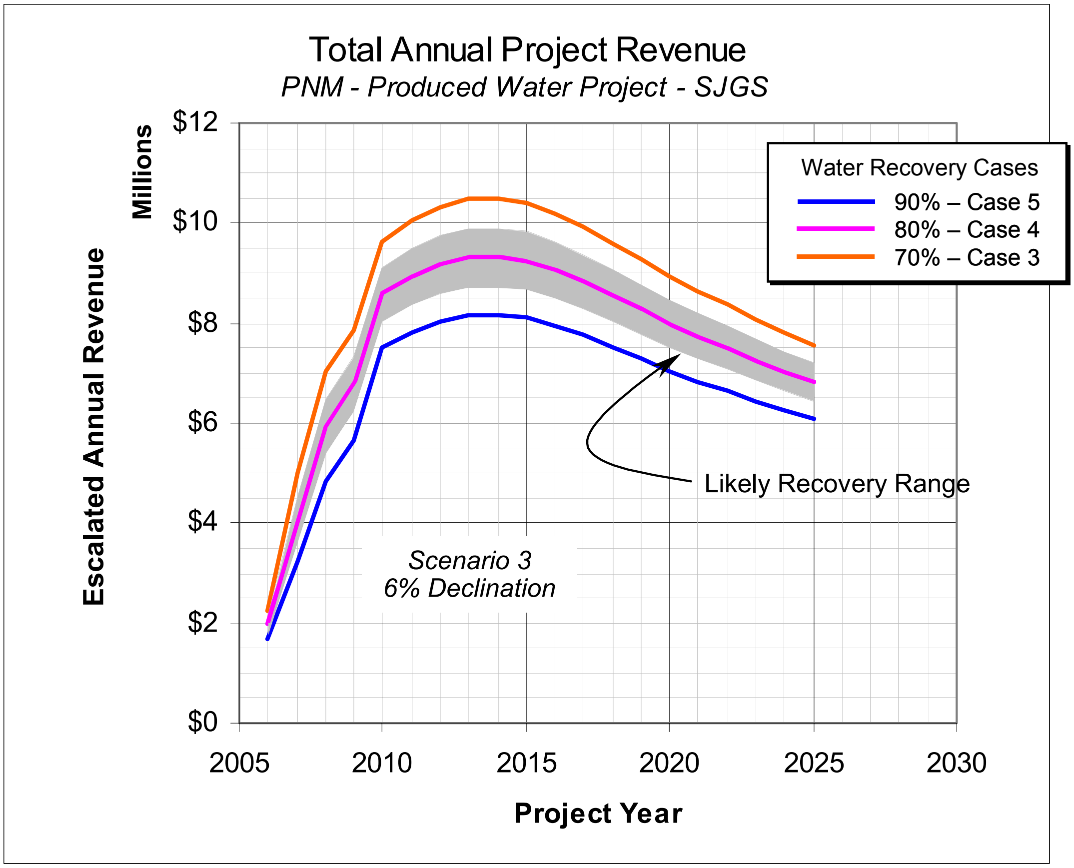

Figure 7-10
Estimated Total Annual Revenue

Several sound arguments support a 75:25 split where PNM receives the greater share:

- PNM's needs are greater because their business could be harmed financially if the project does not occur. During a prolonged drought, PNM might have to reduce load and pay significant fuel penalties. *In this scenario, the producers would continue to operate with no effect on their business.*
- By sending produced water to SJGS, the producers reduce their environmental liability. *PNM would take long-term responsibility and environmental liability for the water.*
- *PNM would be taking the largest financial risk by investing $37.9 million* in the project, significantly more than any of the other participants. Even with a lesser share, the *producers would payout their investment quickly* (discussed later).
- PNM would also assume some financial risk in potential damage to their equipment or unanticipated O&M costs associated with the treatment, use and disposal of this water.

Regardless of pro/con arguments, producers must meet their internal rate of return (IRR) for this and any capital investment. In the two revenue sharing cases discussed here, producer IRR was met.

Using the two revenue sharing splits, an analysis was performed to determine the life-of-project net cost of water. Operating savings, deferred well installations and revenue for accepting water from other producers (as discussed previously) would form the basis of the revenue stream. The revenue would be split and PNM's share would defray project operating expenses. The relationships described below are used to calculate PNM's cost of collecting, conveying and treating water over the life of the project.

The following relationship is used to calculate time-corrected costs (base year 2006) for each year of the project.

PNM costs for chemicals, power, materials, services, labor, etc.

PNM's share of producer revenue

$$\left[\begin{matrix}\textit{Time Corrected}\\ \textit{Project Costs}\end{matrix} = \frac{\textit{Escalated Costs} - \textit{Escalated Revenue} + \textit{Annual Capital Recovery} - \textit{Tax Credit}}{\textit{Escalation Factor}}\right]_{\textit{Year ``n''}}$$

Tax credit based on recovered water

Corrected annual project costs are then summed and divided by the life-of-project net acre-feet (AF_{Net}) of water reclaimed for use at SJGS. Recall that 95.3 percent of the recovered water would be reclaimed via treatment. This calculation yields the life-of-project cost to PMN[17] for collecting, conveying and treating produced water based on 2006 dollars and expressed as dollars per net acre-foot of reclaimed water. Refer to the next relationship for calculation details.

$$CRW_{Net} = \frac{\sum_{1}^{n} \dfrac{(M+S)_n(1+i_{MS})^{n-1} + L_n(1+i_L)^{n-1} - (R_n - BF_n)(1+i_{MS})^{n-1} + ACR - TC_n}{(1+i_{MS})^{n-1}}}{\sum_{1}^{n} Net\ RV_n}$$

Where:

$\mathbf{CRW_{Net}}$ = *Cost (2006 basis) of recovered water, \$/AF_{Net} (net volume after treatment)*

$(M + S)_n$ = *Costs (2006 basis) for materials and services in year n*

$(1 + i_{MS})^{n-1}$ = *Escalation factor for materials and services in year n (also used to re-adjust escalated costs to 2006 basis year)*

L = *Labor costs (2006 basis)*

$(1 + i_L)^{n-1}$ = *Escalation factor for labor in year n*

R_n = *Revenue (2006 basis) in year n*

BF_n = *Backflow charge (2006 basis) in year n*

ACR = *Annual capital recovery (n equal payments)*

TC_n = *Tax Credit earned in year n*

$Net\ RV_n$ = *Net Recovered Volume (95.3% of water delivered to SJGS) of water in year n, AF_{NET}*

---

[17] Producer costs and revenue are not included in this analysis.

In this analysis, the escalation factor used for materials and services (approximately 2 percent per year) is also used to adjust costs back to 2006 dollars. Capital recovery is not escalated since this cost consists of twenty equal payments paid annually throughout the life of the project. Tax credits would be earned based on the volume of recovered water for a given year.

Refer to Figure 7-11 for the life-of-project net cost of produced water for all scenarios and cases. Figure 7-11 also includes a wider range of revenue sharing possibilities – 0, 25, 50 and 75 percent PNM share of producer savings. The importance of revenue sharing and produced water recovery is evident. Without revenue sharing and under low-recovery circumstances, the life-of-project net cost of water could approach $\$4,500/AF_{Net}$.

Refer also to Figure 7-12 for a more focused analysis of the life-of-project net cost of produced water *with and without the tax credit.* The analysis is based on a 75 to 85 percent recovery range, 6% compound declination and two revenue splits – 50:50 and 75:25 (PNM to producer).

With a 50:50 share of revenues, the life-of-project net cost of water would vary between \$620 and $\$1,000/AF_{Net}$ with the tax credit and \$1,200 and $\$1,520/AF_{Net}$ without the tax credit. With a 75:25 share (PNM to producer), the cost of water would vary from -\$300 to $-\$30/AF_{Net}$ with the tax credit (indicating possible net revenue under these circumstances) and \$200 to $\$500/AF_{Net}$ without the tax credit. Clearly, both revenue sharing and the tax credit have a significant effect on the life-of-project net cost of water with an overall range of -\$300 to $\$1,520/AF_{Net}$ (a cost spread of $\$1,820/AF_{Net}$) to collect, convey and treat produced water for reuse at SJGS.[18]

7.4.5 Impact on the Cost of Water at SJGS

At present, PNM has rights to divert 24,200 AF/yr of water from the San Juan River via two contracts:

United States Bureau of Reclamation (USBR) provides rights for 16,200 AF/yr at a contract rate of $9/AF. The water right is granted from the Navajo Reservoir, which is upstream of SJGS.

The San Juan coal contract (through BHP Billiton) provides run-of-the-river rights for 8,000 AF/yr at no cost to PNM as long as the fuel contract is in effect. This type of water right is the most susceptible to being reduced dramatically during long-term shortages.

USBR water from Navajo Reservoir will be provided through a tribal entity in 2006. PNM expects the cost of this water to increase from \$9 to \$70/AF.[19] Therefore, the weighted cost of water[20] from the San Juan River to SJGS will rise from its current cost of \$6.50 to \$47/AF. On

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[18] Even though some cost scenarios yield net revenue for PNM, there is still a significant amount of uncertainty associated with any financial projections.

[19] The water will likely cost between \$60/AF to \$70/AF. The higher cost, \$70/AF, was used in the analysis.

[20] Weighting is based on 66.9 percent of water rights from USBR (Navajo Reservoir) and 33.1 percent from run-of-the-river through BHP Billiton.

this basis, the annual cost of freshwater at SJGS will increase from $146,000 to $1,053,000 in 2006.

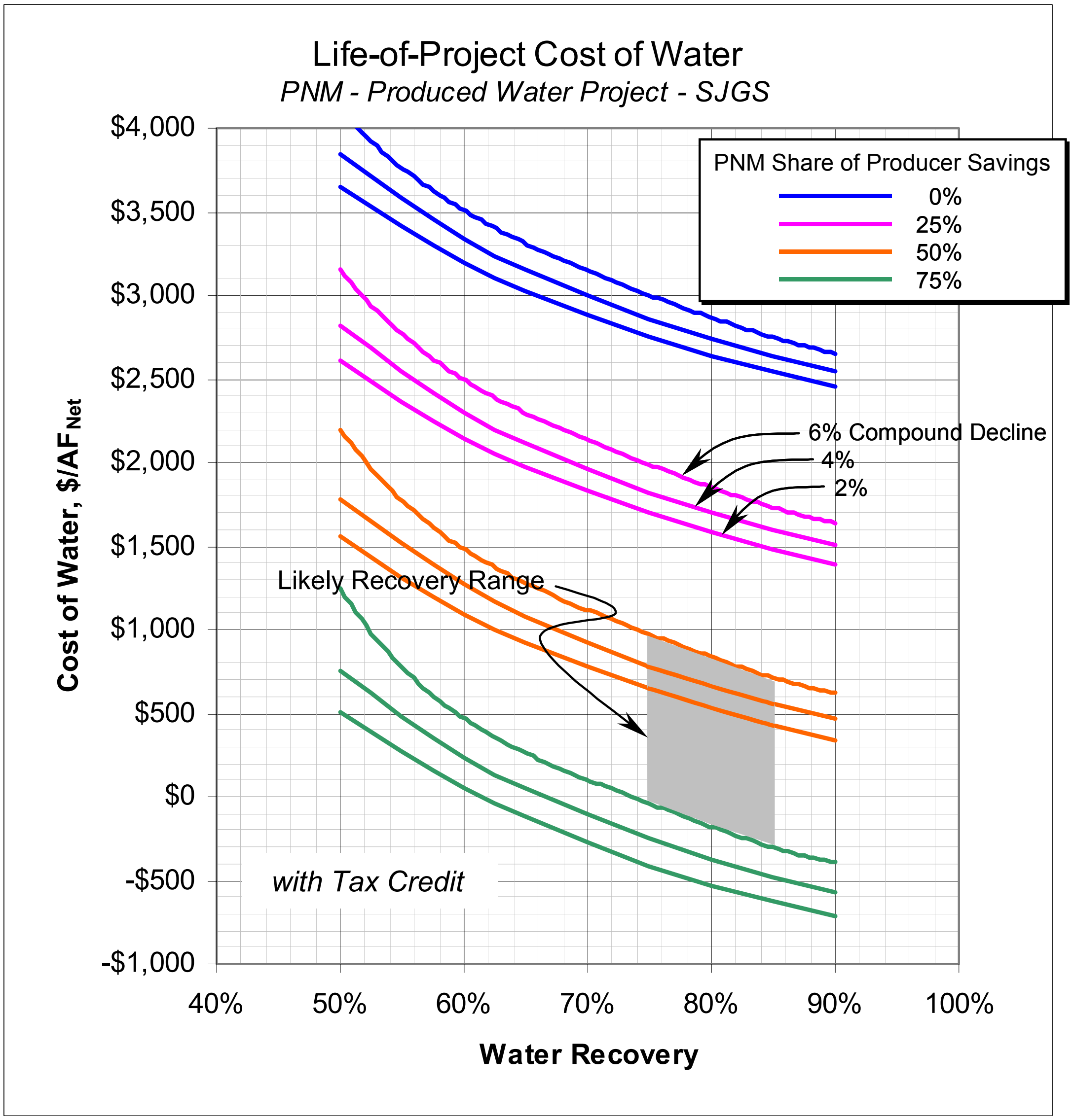

Figure 7-11
Life-of-Project Cost of Water Versus PNM Share of Producer Savings

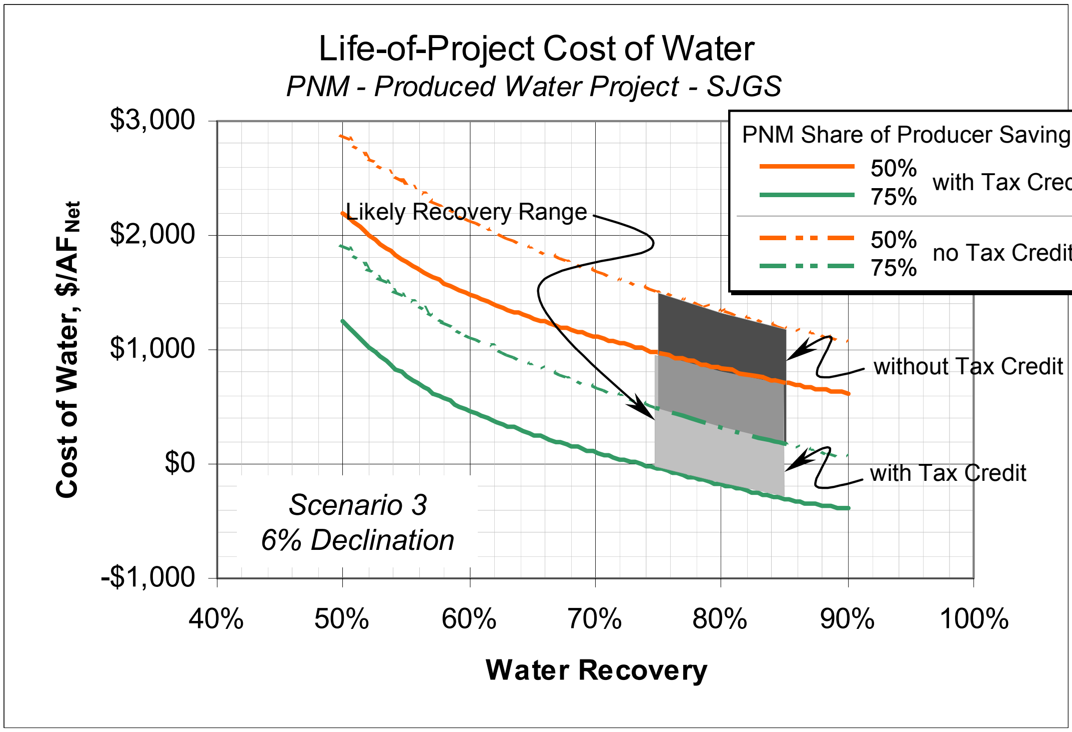

Figure 7-12
Life-of-Project Cost of Water With and Without Tax Credit

To put the produced water project in perspective, refer to Tables 7-6 and 7-7 for an analysis of the blended unit cost and annual cost of water to SJGS, respectively. Note that life-of-project water costs include capital recovery, operating expenses and shared revenue associated with the produced water project. The overall cost impact of the produced water project on the cost of water at SJGS is relatively small. Depending on circumstances such as achievable recovery, passage of the tax credit and the PNM-producer revenue share, the unit cost of blended San Juan River water and produced water would range from –\$9 to \$155/AF_{Net}.

The life-of-project cost of water (2006 basis) *could be reduced by $1.2 million per year* at SJGS with 85 percent recovery of the produced water resource (6 percent decline), a 75:25 (PNM to producer) revenue sharing agreement and the tax credit. Under these circumstances, the produced water project would generate net revenue. Conversely, the cost of water at the plant *could be increased by $2.5 million per year* with 75 percent recovery (6 percent decline), a 50:50 revenue sharing agreement and no tax credit.

As a point of comparison, in California where water resources have been strained for decades, retail water costs range from $200 to $1,200/AF[21]. These are rates that recently-built combined cycle power plants pay for water. Most plants, which must compete with residential and agricultural demands, are paying between $400/AF to $500/AF for water. Reclaimed water is

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[21] These costs were taken from a survey done in support of an as yet unpublished study for the California Energy Commission on the use and cost of water in power plants.

mandated[22] if the plant is reasonably close to a large source of treated municipal effluent. Reclaimed water is typically priced at 90 percent of freshwater costs, i.e. $350 to $450/AF.

Table 7-6
Blended Life-of-Project Unit Cost of Water, San Juan River and Treated Produced Water

		75% Recovery 6% Declination		85% Recovery 6% Declination	
PNM Revenue Share	**Tax Credit Yes/No**	**Blended Cost $/AF**	**Additional Cost $/AF (1)**	**Blended Cost $/AF**	**Additional Cost $/AF (1)**
50%	Yes	$91	$44	$79	$32
75%	Yes	$14	-$33	-$9	-$56
50%	No	$155	$108	$144	$97
75%	No	$81	$34	$60	$13

Notes:

The cost basis for San Juan River Water will be $47/AF in 2006.

Table 7-7
Blended Annual Cost of Water (1), San Juan River and Treated Produced Water

PNM Revenue Share	Tax Credit Yes/No	75% Recovery 6% Declination	85% Recovery 6% Declination
50%	Yes	$2,042,000	$1,774,000
75%	Yes	$315,000	-$196,000
50%	No	$3,469,000	$3,236,000
75%	No	$1,806,000	$1,341,000

Notes:

The cost basis for San Juan River Water will be $1,053,000 per year in 2006.

Relatively speaking, the unit cost (per acre-foot) of blended San Juan River and treated produced water would be significantly less than that paid by power plants in other areas where the market value of water is high. On the other hand, the $37.9 million capital cost of the produced water

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[22] There are no legal requirements for using treated municipal effluent for power plant cooling, but state water policy prioritizes this resource. Power plant developers must consider this resource when proposing a project to the California Energy Commission.

project, which would supply 8 to 10 percent of the SJGS's water during peak recovery years, is quite prohibitive.

Finally, the project investment would have to be weighed against the possibility of a one-year 30-percent shortage in regional water supply. PNM determined that it would be significantly more costly in fuel contract penalties and lost generation than the entire capital investment in the produced water project[23].

7.4.6 Producers Return on Investment

Burlington Resources would invest about $5 million to develop a gathering system for the project. Revenue in the form of reduced operating costs, avoided well replacement costs and fees from receiving produced water from other producers would be generated at the outset of the project. Refer to Figure 7-13. Depending on how revenues are shared with PNM and the extent of produced water recovery, BR could recoup their total investment in gathering system development in 2.8 to 5.0 years. Given the revenue projections for Dugan and Richardson, their investment should payout in less than 4 to 6 months.

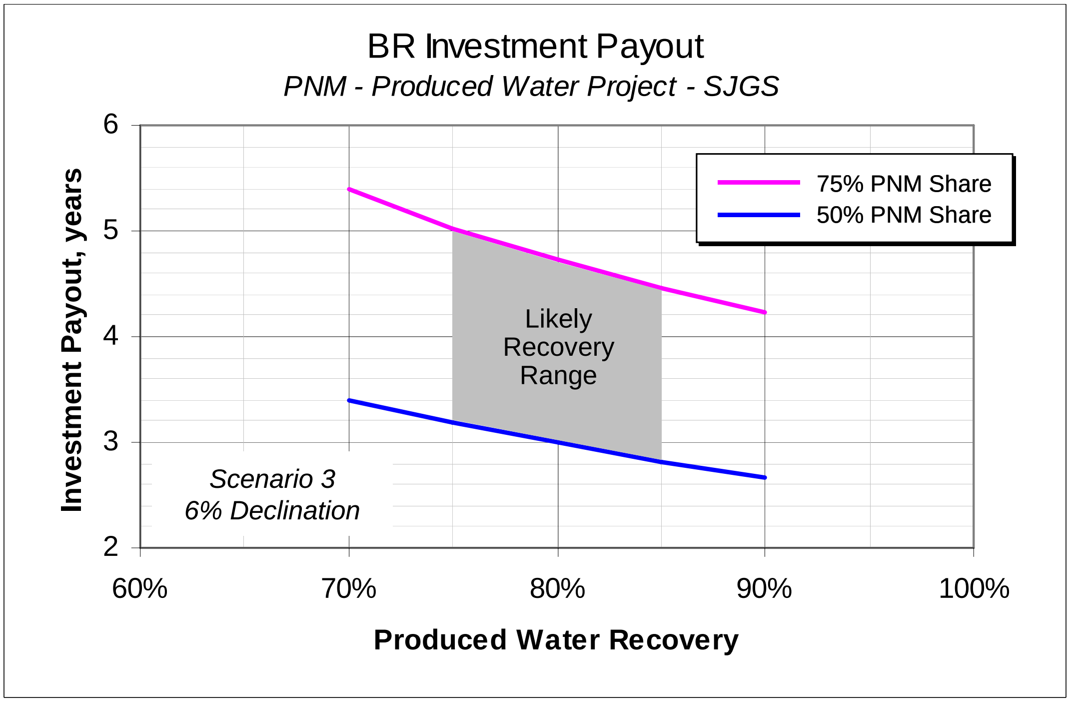

Figure 7-13
BR Investment Payout Time

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[23] As noted previously, PNM preferred that the fuel penalty contract information not be included in this report.

8
IMPLEMENTATION REQUIREMENTS

8.1 Introduction

Public Service of New Mexico (PNM) is evaluating the development of the produced water project in two phases. The first phase would consist of a pipeline to convey water from Close-in producers to a new water treating facility located at San Juan Generating Station (SJGS).[1] In Phase 2, the Collection Center in Bloomfield would be built and the pipeline would be extended to its full length. Burlington Resources (BR) would install satellite collection stations and tie/extend the Hart Canyon Line and the CO_2 Gas Line to the Collection Center. The produced water treatment system at SJGS would be expanded to handle the additional flow.

Legislation enacted in early 2004 removed regulatory barriers that would have required beneficial use assessments for each source of produced water. Produced water can now be disposed of at electric generating stations in New Mexico for treatment and reuse.

There are a number of regulatory agencies that must be engaged and permits that must be obtained to build and operate the produced water gathering, conveyance and treatment system. Project components that must be addressed by PNM are the pipeline, the treatment plant at SJGS and the Collection Center in Bloomfield.

8.2 Two-Phased Implementation Approach

PNM is evaluating the development of the produced water project in two phases to spread capital expenditure over a period of 3 to 5 years. Refer to Figures 8-1 and 8-2 for an overview of the Study Area and a schematic of both phases.

Phase 1 would consist of the following elements:

- Build the first leg of the pipeline (10.8 miles) to convey Close-in produced water to SJGS
- Connect coal bed methane (CBM) producers in the Kirtland area to the pipeline
- Collect mine water from BHP Billiton and cooling tower blowdown from Prax Air
- Install produced water receiving, storage and transfer equipment at SJGS

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[1] Refer to Section 3, *Produced-Water Infrastructure Availability and Transportation Analysis*, for a description of produced water gathering and conveyance. Refer to Section 4, *Produced Water Use, Treatment, and Disposal Analysis*, for a description of the produced water treatment system – Alternative 10 – HERO® + BC 3.

- Install the HERO® system to treat gathered produced water and absorber Purge Water for reuse at SJGS
- Install a 10-acre evaporation pond to handle excess wastewater generated in the Phase 1 portion of the project.

The total recovered water after treatment for Phase 1 would be 534 AF/yr.

Phase 2 consists of the following elements:

- Install satellite collection stations (BR's scope of work) to gather water North of Aztec via their Hart Canyon Line and CO_2 Gas Line
- Build the Collection Center in Bloomfield and pretreat water to remove oil and grit
- Extend the pipeline from the Kirtland area to Bloomfield to a total length of 28.5 miles
- Expand the HERO® system by adding additional media filter, WAC and RO capacity
- Refurbish BC 3 to treat the increased wastewater flow from the HERO® system.

The average life-of-project recovered water after treatment for Phases 1 and 2 would be 1,700 AF/yr.[2]

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[2] The Phase 2 HERO®/BC 3 treatment system would recover 95.3 percent of incoming produced water and Purge Water.

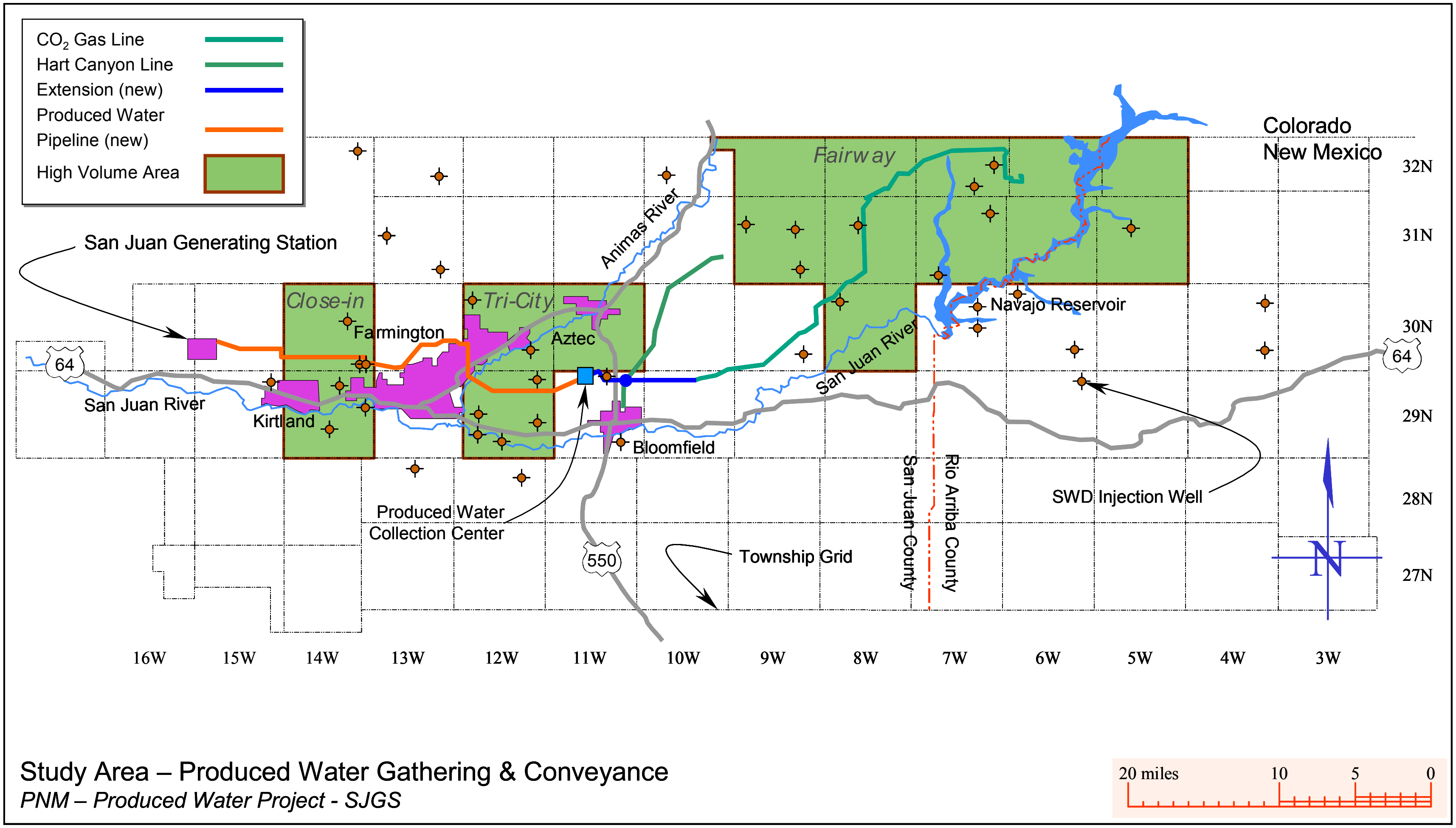

Figure 8-1
Produced Water Gathering and Conveyance System in Study Area

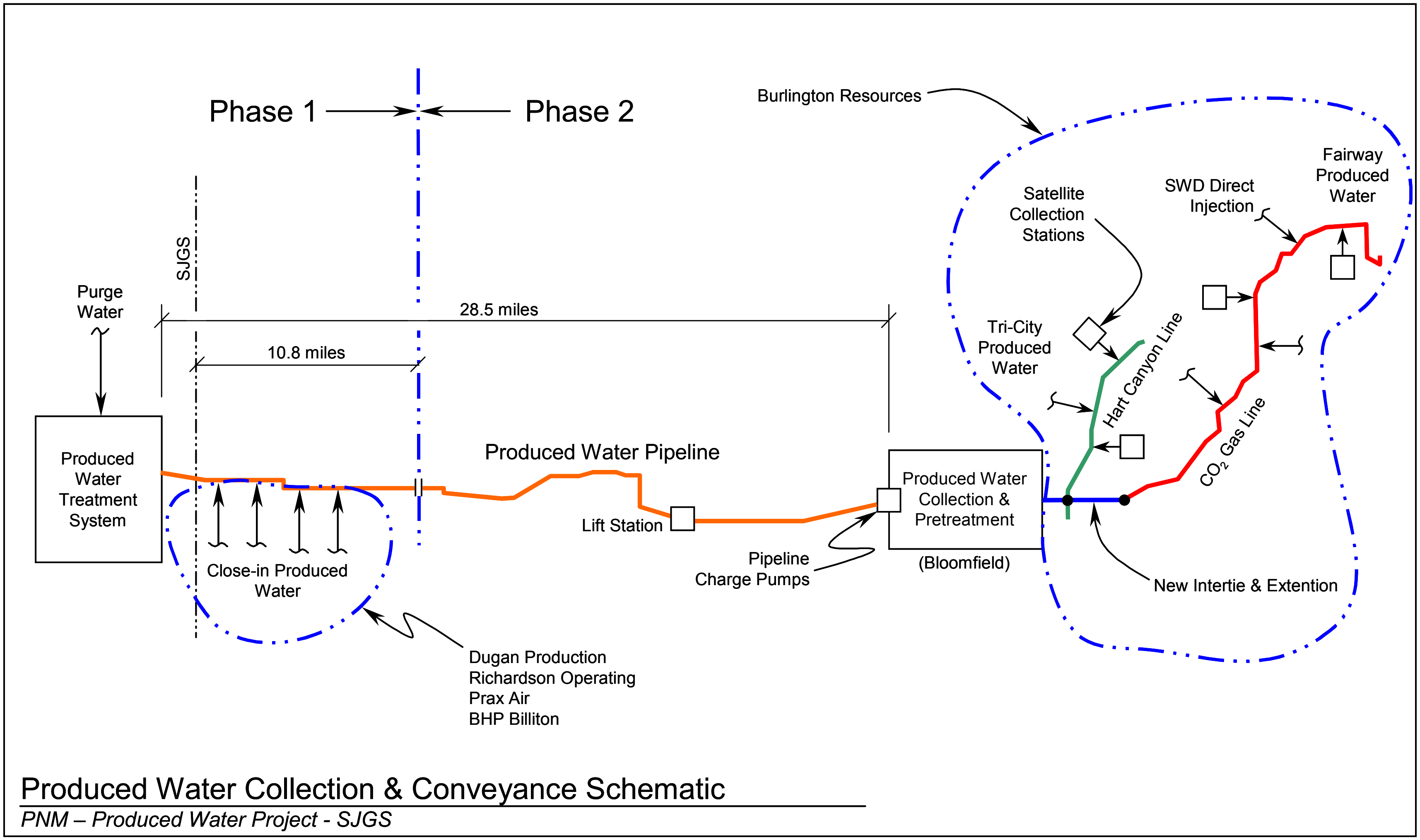

Figure 8-2
Produced Water Collection and Conveyance Schematic

8.3 Regulatory Barriers

The Oil Conservation Division (OCD) regulates all oil and gas production in New Mexico, and as such, produced water is designated a waste byproduct of production. There have been several attempts to utilize produced water (e.g. for dust suppression or road construction) rather than dispose of it via injection. In New Mexico, this action is defined as a beneficial use of the state waters and is regulated by the Office of the State Engineer (OSE). Under this designation, a right to use the water must be obtained and its use must comply with all applicable environmental regulations. The regulatory and environmental protection afforded by the OCD (designating the water as a byproduct of oil and gas production) would be lost with beneficial use.

PNM endeavored to address this regulatory issue by supporting a bill in the New Mexico legislature in January of 2004 that would specifically allow the "disposal" of produced water at electric generating facilities. This would allow produced water reuse as an alternate method of disposal. Therefore, a beneficial use would not be created and the regulatory jurisdiction of the OSE would not be invoked.

The bill was introduced into the January-February 2004 state legislative session and was signed into law March 2004 with the support of both the OCD and OSE. As a result, SJGS could treat and utilize the water for cooling tower make-up, scrubber make-up, ash wetting, etc. Most of the water would be consumed through evaporative losses or waters of moisture in scrubber sludge or ash. Any residual produced water (wastes from treatment) would require disposal to the evaporation ponds at SJGS.[3] OCD jurisdiction of produced water would end at SJGS and would include the treatment system. However, air and wastewater emissions from the treatment plant would be regulated by the New Mexico Environment Department (NMED).

8.4 Phase 1 Implementation

The first phase of the produced water project would consist of the installation of a 14-inch diameter, 10.8-mile pipeline to convey water from Close-in producers in the Kirtland area to SJGS. Refer to Figure 8-3 for a process schematic of Phase 1 and to Figure 8-4 for an overview of Phase 1 and 2 treatment equipment. The pipeline would be the first leg of the 28.5-mile pipeline that would originate in Bloomfield at the Collection Center.

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[3] SJGS would have to obtain an amendment to their existing wastewater disposal permit for this new waste stream.

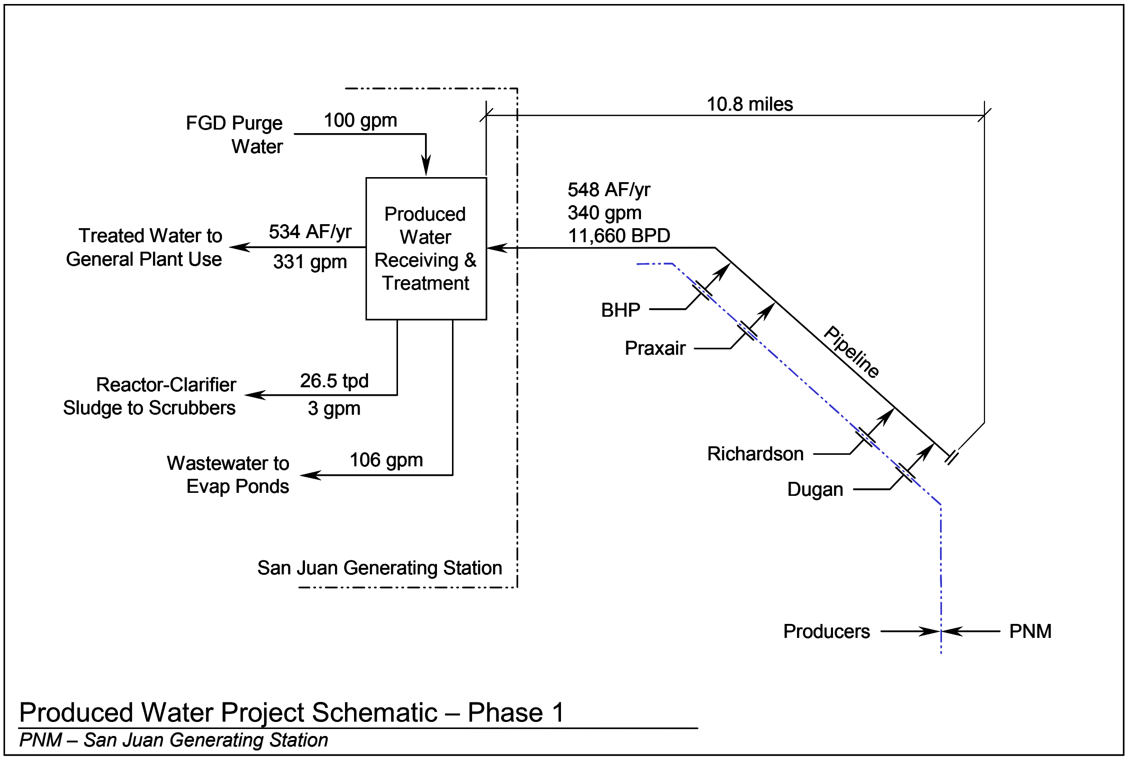

Figure 8-3
Produced Water Project Schematic - Phase 1

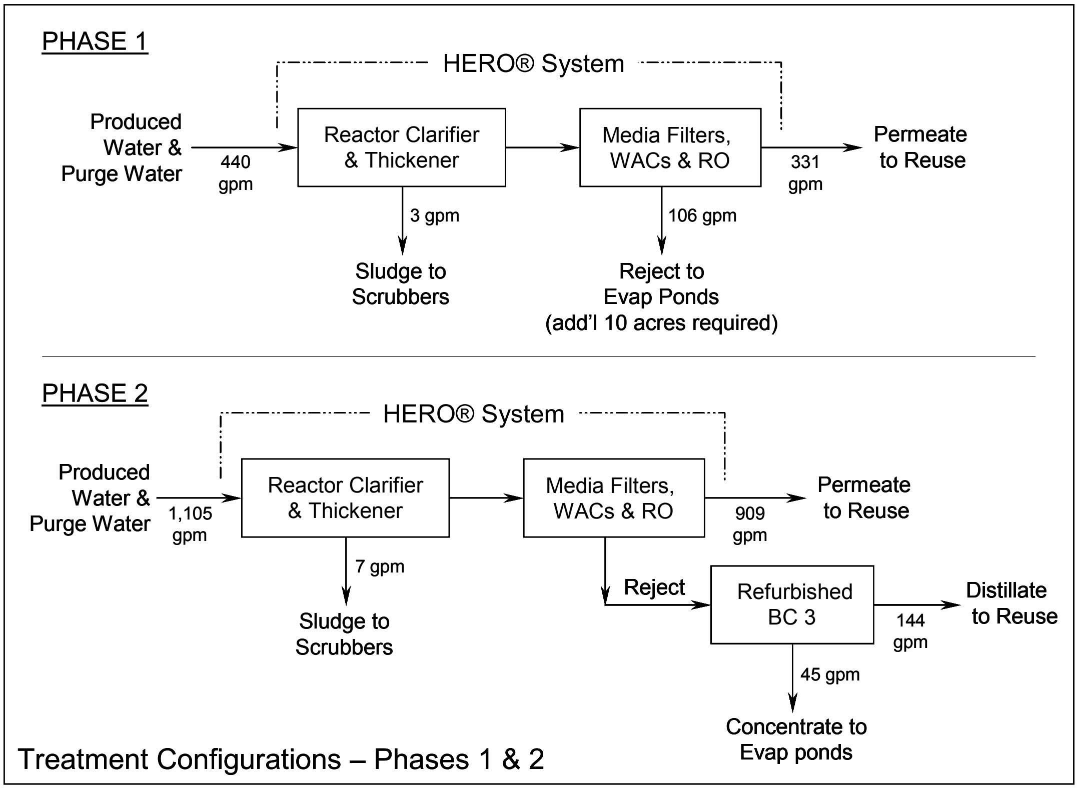

Figure 8-4
Produced Water Treatment Configurations - Phases 1 and 2

In this section of the pipeline, the slope from Close-in production to SJGS is downward, so the first phase would not require charge pumps or booster pumps. Dugan Production Corporation (Dugan) and Richardson Operating Company (Richardson) would supply pressurized water to the pipeline to deliver water to SJGS. BHP Billiton and Prax Air would also supply mine water and cooling tower blowdown, respectively. The total delivered flow in Phase 1 would be 548 AF/yr (340 gpm).

Water would be received and stored at SJGS in a three-day basin. The basin would be sized for Phase 2 flow and would provide two functions. First, it would equalize variations in received produced water chemistry. Second, it would provide water if produced water delivery were interrupted. From the basin, produced water would be pumped to the HERO® system for treatment. Recall that the water delivered from Close-in producers would be exclusively CBM water, and as such, would be essentially free of oil and grit. Dugan and Richardson would filter their water before it is placed into the pipeline for conveyance to SJGS. Purge Water (100 gpm) from the SO_2 absorbers would be blended with produced water for a total flow of 440 gpm (710 AF/yr) before it is fed to the HERO® system.

The HERO® system would recover 75.2 percent of the blend of produced water and Purge Water for a total of 534 AF/yr (331 gpm). Reactor clarifier sludge (26.5 tons per day, wet basis) would be used as supplemental limestone feed to the SO_2 absorbers. The reactor clarifier and thickener would be sized to treat the Phase 2 flow. Two options were evaluated in sizing the reactor clarifier:

- Install a 440 gpm reactor clarifier and thickener in Phase 1 and install another 1,100 gpm reactor clarifier and thickener in Phase 2 for a total capacity of 1,540 gpm (53,000 BPD).[4]
- Install a 1,540 gpm reactor clarifier and thickener in Phase 1.

The reactor clarifier is the most difficult piece of treatment equipment to operate (relative to other HERO® equipment) and two reactor clarifiers would unnecessarily complicate the operation. Also, it would be less costly in the long term if only one reactor clarifier and thicker set were installed.

The HERO® system would generate 106 gpm of wastewater. The evaporation ponds currently have enough capacity to handle 100 gpm of wastewater in this configuration. Therefore, to avoid the capital investment of refurbishing BC 3 in Phase 1, PNM has elected to install ten additional acres of evaporation ponds to handle the 6 gpm of excess water[5].

The treated water could be used for SO_2 absorber, ash system or cooling tower make-up (preferably in this order of use). As discussed in Section 6, *Treated Produced Water*

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[4] Refer to Section 7, *Cost/Benefit Analysis*, for equipment sizing criteria. Also, refer to Footnotes 8 and 10.

[5] Based on the expected evaporation rate, only three acres would be required for additional evaporation pond capacity. The plant determined that a 10-acre pond is the minimum size for a cost effective design. Also, the pond would provide capacity for occasional/unplanned plant wastewater.

Compatibility Assessment, using the treated water for absorber make-up requires minimal expense[6].

8.5 Phase 2 Implementation

In Phase 2, the 14-inch pipeline[7] would be extended to its full length of 28.5 miles to treat an average life-of-project flow of 1,790 AF/yr (1,105 gpm)[8] of Close-in, Tri-City and Fairway produced water, water from Prax Air and BHP Billiton, and absorber Purge Water. Refer again to Figure 8-2 for a schematic of the entire pipeline and gathering system.

BR would install satellite collection stations along the Hart Canyon Line and CO_2 Gas Line. PNM would build the Collection Center in Bloomfield to receive and treat[9] produced water delivered by BR. The pipeline extension would meet the Phase 1 line in the Kirtland area. The pipeline extension would have charge pumps and a booster station to handle an increase in elevation and line losses. Gathered water would be blended with 100 gpm of absorber Purge Water at SJGS.

The HERO® system would be expanded from 440 gpm to 1,550 gpm[10] to accommodate the additional flow from the Tri-City and Fairway areas. This would be accomplished by adding additional media filter vessels, WAC vessels and additional RO modules. The increased wastewater stream generated by the HERO® system would require the refurbishment of BC 3. The HERO®/BC 3 configuration would recover 95.3 percent of the water treated for an average life-of-project total of 1,706 AF/yr (1,053 gpm). BC 3 would generate 45 gpm of wastewater, however, no additional evaporation ponds would be required. Reactor clarifier sludge (78.9 tons per day, wet basis) would be used as supplemental limestone feed to the SO_2 absorbers.

As in Phase 1, the treated water could be used for SO_2 absorber, ash system or cooling tower make-up.

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[6] Using the treated water for the ash system does not require any additional expense, however, the ash system has a relatively small water demand.

[7] The pipeline would be capable of carrying 60,000 BPD (1,800 gpm) of water. In the event that more produced water became available, PNM wanted to be able to have additional pipeline capacity.

[8] Based on 75 to 85 percent recovery of the produced water resource, 6 percent compound declination of the resource and a project life of 20 years. A mid-range recovery of 80 percent was selected for this analysis.

[9] Treatment at the Collection Center in Bloomfield would consist of gravity separation, gas flotation and walnut shell filtration to remove oil and grit. Refer to Section 4.5 and Figure 4-10, Treatment and Disposal Analysis, for more detail.

[10] The HERO® system is sized to treat 53,000 BPD (1,550 gpm), which is the maximum predicted flow (plus a 10 percent capacity cushion) of the produced water resource assuming 85 percent recovery at 6 percent compound declination.

8.6 Environmental Issues

There are a number of regulatory agencies that must be engaged and permits that must be obtained to build and operate the produced water gathering, conveyance and treatment system. This part of the section addresses the major components of the project and the environmental permits and activities required to implement each phase. Project components that must be addressed by PNM are the pipeline, the treatment plant at SJGS and the Collection Center in Bloomfield. The discussion is purposefully general because of the complexity of the permitting effort and the uncertainties associated with eventual permitting strategies. Environmental issues that must be addressed by the participating oil and gas producers are not included in this analysis.

8.6.1 Phase 1 Environmental Issues

Pipeline

An environmental assessment (EA) must be conducted to determine if any impacts would be created by building and operating the Phase 1 portion of pipeline (10.8 miles), e.g. disturbed habitat during construction or operation of the pipeline. Also, because of the possibility of finding Native-American artifacts, archeological surveys and mitigation plans must be included in the pipeline design. The Bureau of Land Management (BLM) would likely be the lead agency in this effort because a significant portion of the pipeline passes over federal lands. OCD would review the pipeline design, require integrity testing before start-up, and require operating and spill contingency plans.

Phase 1 Treatment Plant

The produced water treatment plant at SJGS would be treated like a storage/disposal facility by ODC[11] and a permit would have to be obtained to build and operate it. As part of the permit application, PNM would have to provide site topographic, geologic and hydrologic information, plant design information, plans for waste handling and spills, etc.

BTEX[12] would be released to the air (likely <0.1 pound per day) from the HERO® system and reuse in the SO_2 absorbers, ash system or cooling towers. BTEX emissions would be low, because produced water in Phase 1 would be generated entirely by CBM production. NMED would be notified of the emissions at the outset of the project, however at these levels, it likely would not require a modification to the plant air permit.

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[11] OCD would likely consider the treatment plant similar to a salt water disposal (SWD) facility, where water is stored, treated and disposed of.

[12] BTEX is the sum of the concentrations of benzene, toluene, ethyl benzene and xylene. BTEX is commonly found in conventionally produced water and at trace levels in CBM water. Refer to Section 4.4.7, Disposition of Treatment Wastes for more details.

The plant wastewater permit would have to be modified. Additional wastewater would be generated from HERO® system reject and would be sent to the existing and new plant evaporation ponds for disposal.

Phase 1 environmental permit activity could take up to six months to complete.

8.6.2 Phase 2 Environmental Issues

Pipeline Completion

An EA would be conducted to determine if any environmental impacts would be created by completing the pipeline. Archeological surveys would again be included in the pipeline design. BLM could be the lead agency, however, this leg of the pipeline passes over much more private property and city and state lands than the Phase 1 segment. OCD would review the pipeline completion design, require integrity testing of the extension before start-up and require updates to the operating and spill contingency plans.

Collection Center in Bloomfield

The Collection Center in Bloomfield would be treated like a storage/disposal facility by ODC and a permit would be obtained to build and operate the center. Like the produced water treatment plant at SJGS, PNM would have to provide site and design information and operating plans for the center. An air permit would have to be obtained from NMED for potential BTEX emissions – up to 25 to 30 pounds per day of BTEX could be generated at the center. No wastewater would be generated at the Collection Center. Unrecoverable produced water and waste products (e.g., grit) would be transported to existing licensed disposal operations. Recovered oil would be reclaimed at the Giant Refinery in Bloomfield.

Treatment Plant Expansion

The permit for the produced water treatment plant would be modified to reflect its increased capacity (OCD lead). Both air and wastewater permits would have to be modified to include emissions from produced water treatment (NMED lead). Air emissions could be significantly different with higher levels of BTEX in the delivered produced water (up to 25 to 50 pounds per day). BTEX might also meet the threshold limits to require reporting in the annual Toxics Reporting Inventory (TRI) for SJGS. Wastewater, which would consist of brine from BC 3, would be sent to the evaporation ponds.

Phase 2 environmental permit activity could take six to nine months to complete.

8.7 Capital Expenditure

By developing the project in two phases, PNM could spread capital investment over a period of 3 to 5 years. However, as discussed previously, phasing the project would require a 10-acre evaporation pond to handle excess wastewater in Phase 1. PNM also has decided to use 25 percent contingency for the first phase of the project to cover uncertainties that might arise in a novel reuse project. The evaporation pond and additional contingency would increase the total cost of the project by $3,010,000. Refer to Table 8-1 for a summary of costs by phase.

Table 8-1
Capital Expenditure by Project Phase

PNM – Produced Water Project – SJGS

Equipment Description	**Phase 1**	**Phase 2**	**Total**
Bloomfield Collection Center	$0	$5,200,000	$5,200,000
14-inch Pipeline	$2,940,000	$9,960,000	$12,900,000
Receiving & Transfer Equipment (1)	$1,080,000	$420,000	$1,500,000
HERO System	$3,500,000	$3,760,000	$7,260,000
Refurbish BC 3	$0	$2,970,000	$2,970,000
10-Acre Evaporation Pond	$1,710,000	$0	$1,710,000
Subtotal	$9,230,000	$22,310,000	$31,540,000
Contingency (2)	$2,310,000	$3,350,000	$5,660,000
NMGRT (3)	$570,000	$1,370,000	$1,940,000
PNM G&A (4)	$510,000	$1,230,000	$1,740,000
Total Phased Project Cost	$12,620,000	$28,260,000	$40,880,000
Non-Phased Project Capital Cost (5)			$37,870,000
Additional Project Expenditure			$3,010,000

Notes:

1. Includes Receiving Basin, produced water transfer pumps and treated water tank, transfer pumps and transfer line.
2. PNM elected to use 25 percent contingency for Phase 1 of the project to cover uncertainties. 15 percent contingency is used for Phase 2.
3. New Mexico gross receipts tax assessed at 6.125%.
4. PNM general and administrative expenses assessed at 5.5 percent.
5. Refer to Section 6.3 of the Cost/Benefit Analysis.

9
APPLICABILITY TO OTHER REGIONS IN THE U.S.

9.1 Introduction

Produced water is generated nationally as a byproduct of oil and gas production. Seven states generate 90 percent of the produced water in the continental US. About 37 percent of the sources[1] documented in the US Geological Survey's (USGS) Produced Waters Database have a TDS of less than 30,000 mg/l. This is significant because produced water treatment for reuse in power plants was found to be very costly above 30,000 mg/l TDS. For this report, produced water treatment was assessed using the technologies evaluated for the San Juan Generating Station (SJGS) in Section 4, *Produced Water Use, Treatment, and Disposal Analysis.* Also, a methodology was developed to readily estimate capital and operating costs for produced water treatment. Two examples are presented to show how the cost estimating methodology can be used to evaluate the cost of treatment of produced water at power plants close to oil and gas production.

9.2 Produced Water Generation Nationally

Produced water is a byproduct of oil and gas production, and depending on the site, a significant amount can be generated relative to the actual volume of production. This section outlines how produced water is formed and brought to the surface, where it is produced in the US, and its basic chemistry.

9.2.1 How Produced Water is Generated

Produced water is brought to the surface when oil and gas are extracted from bearing formations. Oil and gas deposits form in ancient sediments of organic matter, e.g. in prehistoric ocean bottoms. In time, oil, gas and water co-mingle in the pores of sediment, and when oil and gas are brought to the surface, water is also lifted. Generally, for every barrel of oil, nine barrels of water are brought to the surface. Over time, the amount of water brought to the surface usually increases relative to oil and gas production.

In coal bed methane (CBM) production, gas is extracted directly from coal seams. To allow the gas to separate from the coal, water above and surrounding the coal must be extracted to reduce hydrostatic pressure to allow methane release (with the water). The amount of water brought to the surface (relative to methane gas) is highly variable and depends on site-specific geologic and

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[1] This threshold value is based on a numeric sort of datasets and is not weighted by produced water volume.

hydrogeologic conditions. In CBM production, water generation is high at the outset and falls off over time.

9.2.2 Where Produced Water is Generated in the US

Refer to Table 9.1 for a summary of produced water generation in the continental US. The table, which was extracted from a report prepared by Argonne National Laboratory[2], identifies produced water generation in 31 states for the years of 1985, 1995 and 2002. For many of the states, produced water generation was estimated by using historic water-to-product ratios. Nationally, produced water volume is dropping along with reduced conventional oil and gas production.

The annual volumes prepared by Veil 2003 also include produced water that is treated and reused for water floods or steam floods in enhanced oil and gas production; therefore, this water is not available for downstream reuse.

Table 9-1 can be sorted into three tiers (refer to the summary below). The first tier of states generated 90.1 percent of the produced water in 2002 (volume greater than 813 MBPY[3]) – Alaska, California, Kansas, Louisiana, Oklahoma, Texas and Wyoming. Texas alone generated 35.5 percent of the produced water in the US in 2002. The next tier (78 MBPY to 813 MBPY) – Alaska, Arkansas, Colorado, Illinois, Mississippi, Montana, New Mexico, North Dakota and Utah – generated 8.5 percent of the produced water. The last tier (15 states) generated 1.4 percent.

	Tier Criteria MBPY	Daily Produced Water Volume BPD	Number of States	Fraction of Total Volume
Tier 1	>813	34,965,000	7	90.1%
Tier 2	78 to 813	3,294,000	9	8.5%
Tier 3	<78	537,000	15	1.4%
Total	----	38,796,000	31	100.0%

Clearly, opportunities for produced water reuse should be focused in Tier 1 states and secondarily in Tier 2 states. The treatment and reuse of produced water at SJGS is a good example of a Tier 2 opportunity.

Current market pressures to increase CBM development and production are accelerating produced water generation in many states. New CBM development should dampen the decline

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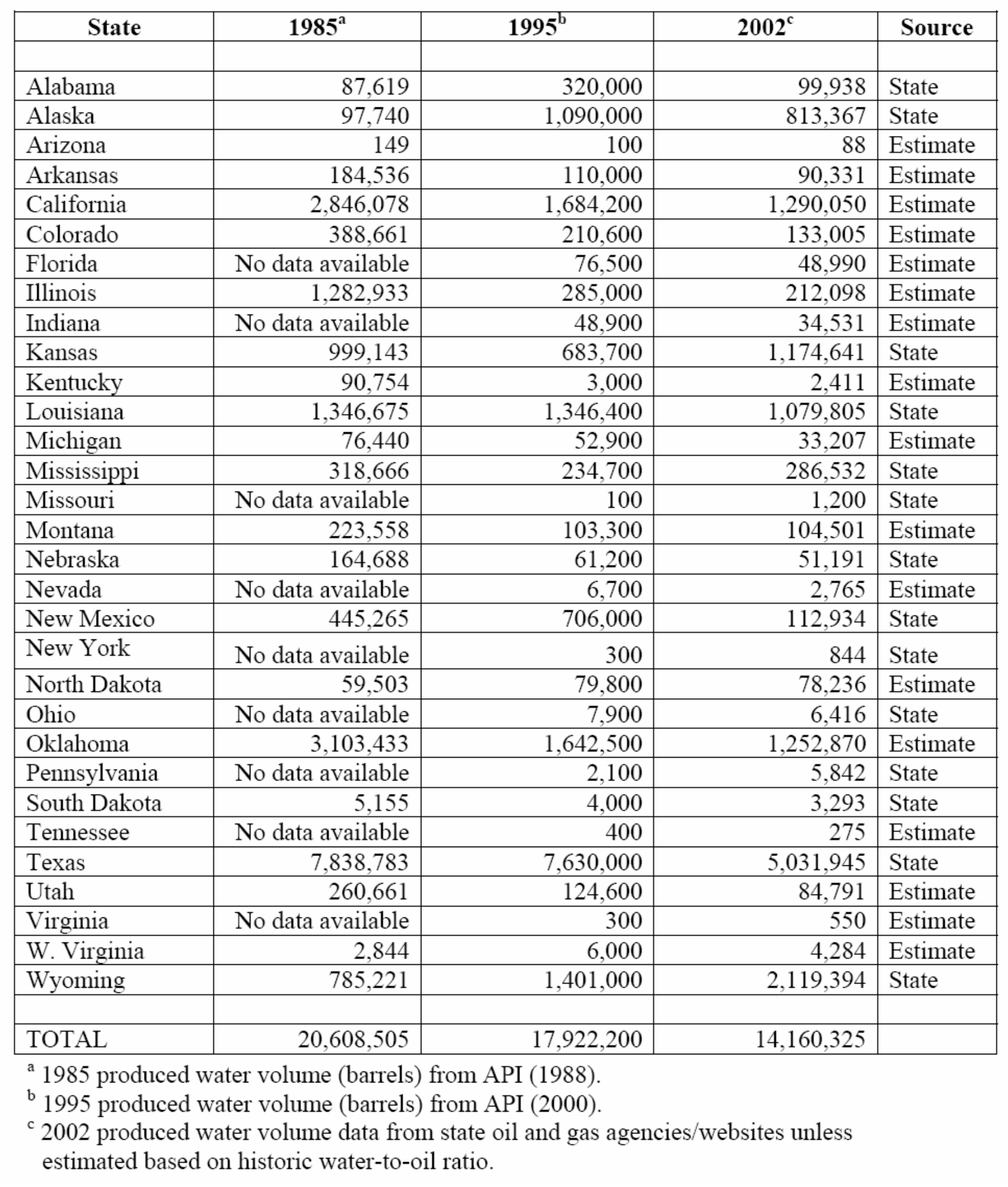
[2] J.A. Veil, M.G. Puder, D. Elcock and R.J. Redweik, Jr., "A White Paper Describing Produced Water From Production of Crude Oil, Natural Gas and Coal Bed Methane", prepared by Argonne National Laboratory for the US Department of Energy, National Energy Technology Laboratory, January 2004.

[3] MBPY corresponds to one million barrels of produced water per year – 1 MBPY is equivalent to 2,740 BPD or 80.0 gpm.

in produced water volume in a number of states where there are large coal reserves such as Colorado, Wyoming and Montana. Also note that produced water in Wyoming (refer to Table 9-1) has increased steadily as a result of CBM production. Refer to Figure 9-1 for a map of coal basins that produce (or could possibly produce) CBM. The map was prepared by ALL Consulting.[4]

Table 9-1
Annual Onshore Produced Water Generation by State (1000 bbl), prepared by Argonne National Laboratory, 2004



State	1985[a]	1995[b]	2002[c]	Source
Alabama	87,619	320,000	99,938	State
Alaska	97,740	1,090,000	813,367	State
Arizona	149	100	88	Estimate
Arkansas	184,536	110,000	90,331	Estimate
California	2,846,078	1,684,200	1,290,050	Estimate
Colorado	388,661	210,600	133,005	Estimate
Florida	No data available	76,500	48,990	Estimate
Illinois	1,282,933	285,000	212,098	Estimate
Indiana	No data available	48,900	34,531	Estimate
Kansas	999,143	683,700	1,174,641	State
Kentucky	90,754	3,000	2,411	Estimate
Louisiana	1,346,675	1,346,400	1,079,805	State
Michigan	76,440	52,900	33,207	Estimate
Mississippi	318,666	234,700	286,532	State
Missouri	No data available	100	1,200	State
Montana	223,558	103,300	104,501	Estimate
Nebraska	164,688	61,200	51,191	State
Nevada	No data available	6,700	2,765	Estimate
New Mexico	445,265	706,000	112,934	State
New York	No data available	300	844	State
North Dakota	59,503	79,800	78,236	Estimate
Ohio	No data available	7,900	6,416	State
Oklahoma	3,103,433	1,642,500	1,252,870	Estimate
Pennsylvania	No data available	2,100	5,842	State
South Dakota	5,155	4,000	3,293	State
Tennessee	No data available	400	275	Estimate
Texas	7,838,783	7,630,000	5,031,945	State
Utah	260,661	124,600	84,791	Estimate
Virginia	No data available	300	550	Estimate
W. Virginia	2,844	6,000	4,284	Estimate
Wyoming	785,221	1,401,000	2,119,394	State
TOTAL	20,608,505	17,922,200	14,160,325	

[a] 1985 produced water volume (barrels) from API (1988).
[b] 1995 produced water volume (barrels) from API (2000).
[c] 2002 produced water volume data from state oil and gas agencies/websites unless estimated based on historic water-to-oil ratio.

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[4] "Handbook on Coal Bed Methane Produced Water: Management and Beneficial Use Alternatives", prepared by ALL Consulting for Groundwater Protection Research Foundation and for the US Department of Energy, National Energy Technology Laboratory, July 2003.

Map of U.S. Coal Reserves/Basins

U.S. coal reserves and basins

Prepared by ALL Consulting, 2002

Figure 9-1
Coal Reserves and Basins in the United States

9.2.3 Produced Water Chemistry

The USGS has compiled a provisional Produced Waters Database.[5] The database contains well information (well name, well owner, state location, township and section numbers, longitude and latitude, etc.) and basic produced water chemistry. Some of the information dates back 80 years. Chemistry data provided by Veil 2003 (conventional and CBM sources), ALL 2003 (CBM sources) and the author's work in California and New Mexico fall well within the envelop of data provided by the USGS database.

One of the important values of the data is demonstration of the variability of the produced water resource. For example, produced water TDS in the database ranges from 500 mg/l to 400,000 mg/l. Refer to Figure 9-2 for a distribution of TDS values. About 37 percent of the produced water datasets have a TDS value of less than 30,000 mg/l. This is significant because produced water treatment for reuse in power plants is not economically feasible above 30,000 mg/l TDS (discussed next).

Only basic chemistry is provided in the database, i.e. pH, sodium, potassium, calcium, magnesium, alkalinity, chloride and sulfate. Other chemical information of interest, such as silica, barium, ammonia, volatile organic constituents, etc. is not available except for individual analyses recovered from producers and published technical reports, e.g. Veil 2003 and ALL 2003. Of the 58,700 individual water analyses in the USGS database, 48,600 were deemed useable because their cation/anion balance was within ±5 percent of neutrality.

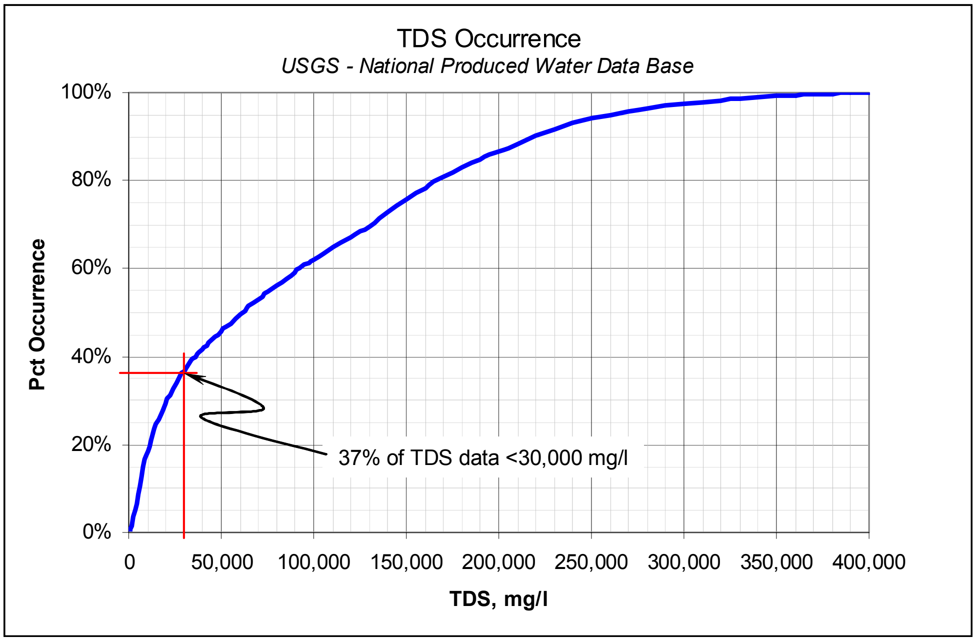

Figure 9-2
Total Dissolved Solids (TDS) Occurrence in Produced Water

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[5] The data are considered provisional because they have not received the approval of the Director of the USGS and are subject to revision. The database, which was posted in May 2002, can be found on the USGS website at energy.cr.usgs.gov/prov/prodwat/.

Given the limitations of the USGS database (along with its wealth of basic chemistry), a methodology is developed next in this section to predict the capital and operating costs of produced water treatment.

9.3 Produced Water Treatability

It is assumed in this analysis that produced water is treated for reuse at a power plant that is reasonably close to conventional oil and gas or CBM production.[6] In some cases, low-TDS produced water could be used with minimal treatment in a power plant, i.e. requiring de-oiling and filtration. Although low-TDS produced water exists, its occurrence is relatively rare. This section develops costs for membrane and evaporative technologies (evaluated for SJGS) to treat a range of saline produced waters.

Lastly, it is assumed that waste streams generated by produced water treatment would either be:

- Mixed with power plant ash and/or SO_2 scrubber sludge and landfilled
- Disposed of in new evaporation ponds
- Brought to dryness via crystallization and landfilled with power plant ash and/or SO_2 scrubber sludge.

9.3.1 Treatment Technology

For this analysis, high-efficiency reverse osmosis (HERO®) and brine concentrator (BC) technologies (discussed in Section 4, *Produced Water Use, Treatment, and Disposal Analysis*) were used to evaluate produced water treatment. Three treatment configurations were evaluated:

- HERO® + BC
- HERO® + BC + evaporation ponds
- HERO® + BC + crystallizer

Refer to Figure 9-3. HERO®, BC, and crystallizers are off-the-shelf technologies that have been used to treat high-TDS wastewater. The applicability of these configurations depends on how a power plant disposes of ash and SO_2 scrubber sludge and whether the climate is suitable for evaporation ponds. It is also assumed that reactor-clarifier sludge could be disposed of along with other treatment solids, since a $CaCO_3$-based waste product may not be suitable as a supplemental feedstock with all types of SO_2 scrubbers. Also, some plants might not have SO_2 scrubbers.[7]

In this analysis, all equipment is assumed new, i.e. no existing equipment is reassigned or refurbished for produced water treatment service.

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[6] Recall that the 28.5-mile pipeline in the produced water assessment for SJGS was almost 45 percent of the total project cost.

[7] In Section 4, *Produced Water Use, Treatment, and Disposal Analysis*, we assumed that SJGS would feed reactor-clarifier sludge to the SO_2 absorbers as supplemental limestone feed.

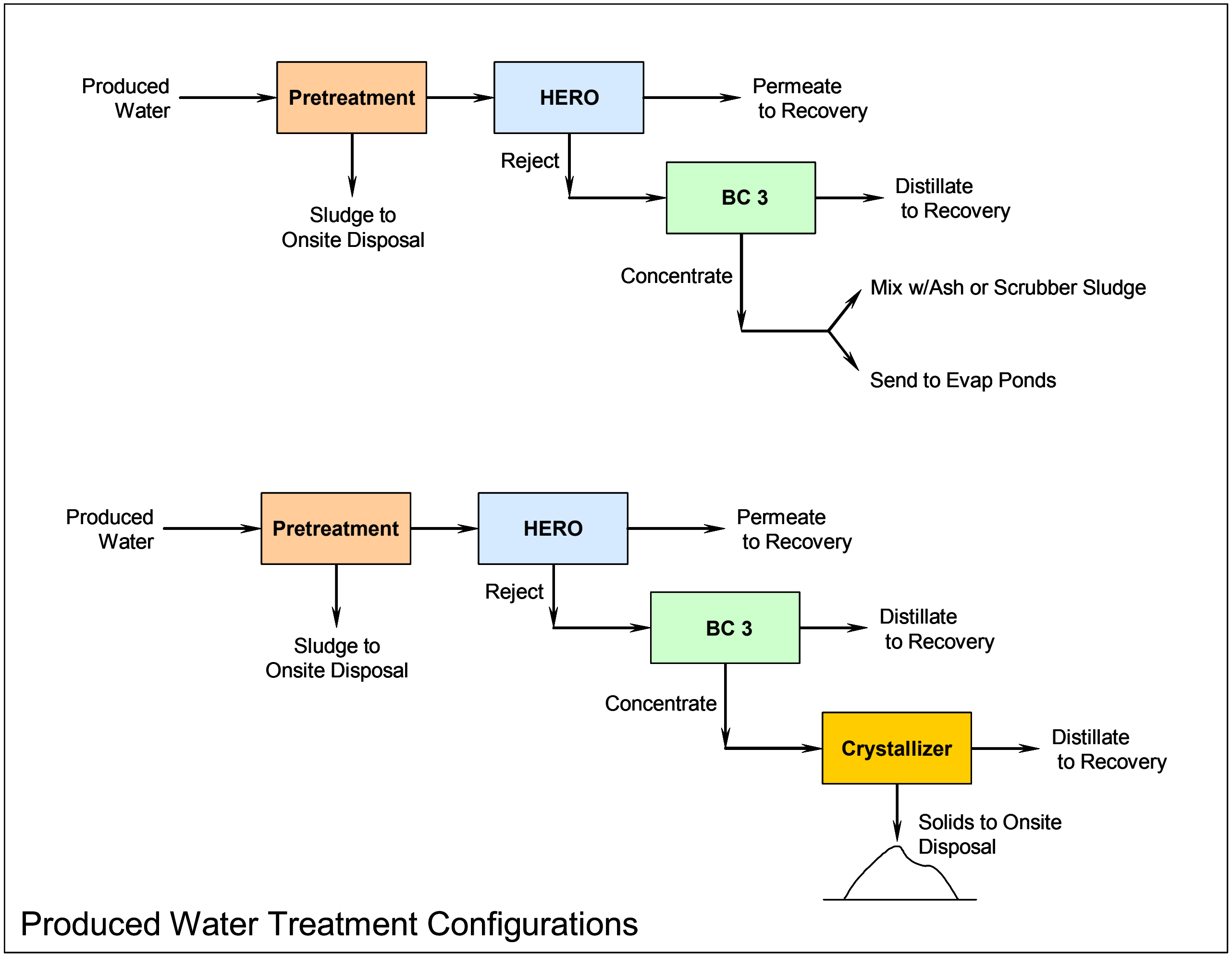

Figure 9-3
Produced Water Treatment Configurations

9.3.2 Treatability Criteria

Constituents evaluated for the treatability analysis are TDS, calcium, magnesium and alkalinity. These constituents drive the analysis because they determine the recovery parameters for treatment equipment as well as influencing operating parameters such as chemical consumption and power requirements.

The following general design criteria were used for the configurations outlined above:

- Reactor-clarifier solids are dewatered to 30 percent solids and landfilled onsite (with ash and/or SO_2 scrubber sludge)
- HERO® recovery is limited to 90 percent recovery or a reject concentration of 60,000 mg/l if 90 percent recovery is not achievable[8]
- BC recovery is limited to a brine concentration of 225,000 mg/l[9]
- The crystallizer is operated to produce a dry waste product consisting of 50 percent solids and landfilled onsite (with ash and/or SO_2 scrubber sludge).

Process criteria, although general, are closely associated with those used for the SJGS produced water project analysis.

The intent of this analysis is to maximize the recovery of the HERO® process and minimize the size of BC and crystallizer equipment and evaporation ponds. BC and crystallizer equipment is significantly more costly than the HERO® process (for a given flow rate) and more costly to operate. Evaporation ponds are capital intensive.

As outlined in Section 4, HERO® pretreatment softening and high-pH operation are well suited to treat a variety of produced waters with high TDS, hardness, silica, traces of oil, etc. HERO® recovery is calculated as follows:

$$HERO\text{®}\ \mathrm{Recov}ery,\ \% = \left[1 - \frac{TDS_{Feed},\ mg/l}{60{,}000\ mg/l}\right] \times 100$$

For this analysis, the HERO® process is limited to a feedwater TDS limit of 30,000 mg/l and a recovery of 50 percent. If the feedwater TDS limit were raised to just 35,000 mg/l, allowable recovery would drop to 42 percent, and at 40,000 mg/l, recovery would only be 33 percent.

For example, if 50,000 BPD of produced water with a TDS of 12,000 mg/l were to be treated, the HERO® process would recover 80 percent of the feedwater as permeate (40,000 BPD). Refer to

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[8] HERO® reject is limited to the osmotic pressure rating of the membranes, which is equivalent to 70,000 to 75,000 mg/l of TDS. A conservative operating limit of 60,000 mg/l was selected. This slightly increases the size of the equipment that must be installed to reduce total wastewater volume to the brine concentrator and evaporation ponds or crystallizers.

[9] This assumes the BC is operated at a pH of 10.0 to 11.0 with no chloride limitation.

the summary below. The BC would treat the remaining 20 percent of HERO® reject (10,000 BPD). Since the HERO® would be operated at a maximum reject concentration of 60,000 mg/l and BC brine concentration would be set at 225,000 mg/l, the BC would recover 73.3 percent in all cases. Therefore, 7,330 BPD of HERO® reject would be recovered by the BC. This would leave 2,670 BPD of BC brine to either be landfilled with ash or scrubber sludge, sent to an evaporation pond, or treated further by a crystallizer to dry salts.

Stream	**Flow Rate**	**TDS**
Feedwater	50,000 BPD	12,000 mg/l
HERO® Permeate	40,000 BPD	<500 mg/l
BC Feedwater (HERO® Reject)	10,000 BPD	60,000 mg/l
BC Distillate	7,330 BPD	<10 mg/l
BC Brine	2,670 BPD	225,000 mg/l
Total Recovered	47,330 BPD	(94.7% Recovered)

9.3.3 Chemistry Assumptions

Refer to Figures 9-4, 9-5 and 9-6 for relationships between TDS and calcium, TDS and magnesium, and TDS and alkalinity, respectively. Emphasis was placed on evaluating calcium, magnesium and alkalinity relationships because the cost of pre-softening produced water with a reactor clarifier usually dominates all other chemical costs. The sheer volume of information in the USGS database established well-defined, dense envelopes for each relationship (17,100 datasets were within the TDS range of 0 to 30,000 mg/l).

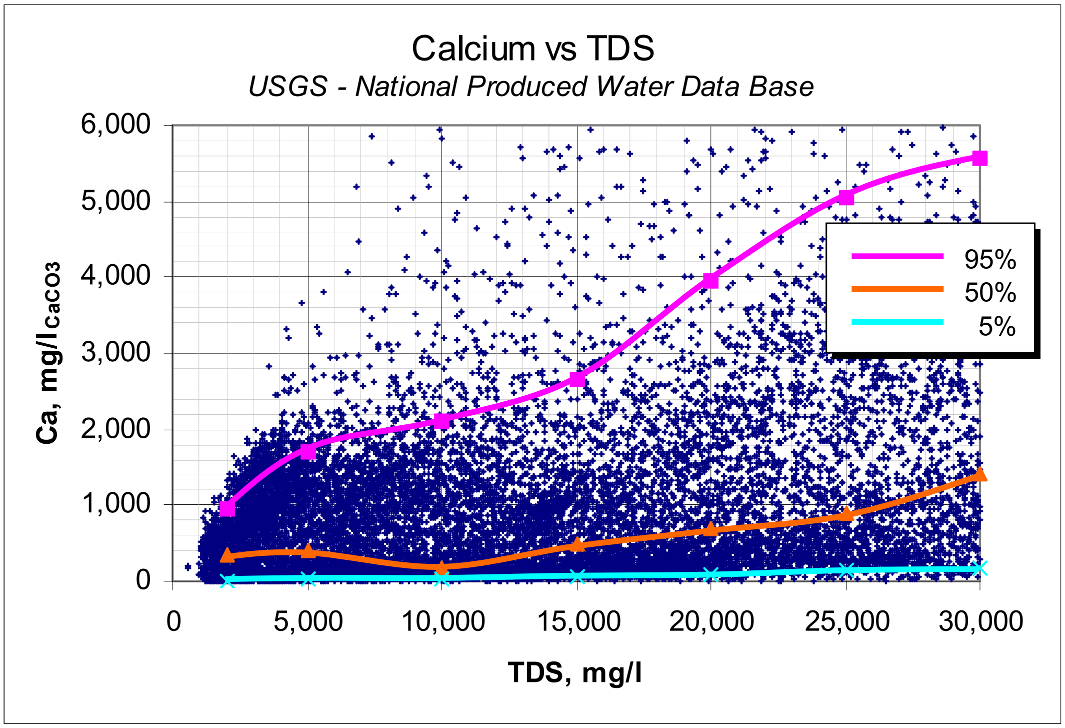

Figure 9-4
Calcium vs. Total Dissolved Solids Concentrations in Produced Water

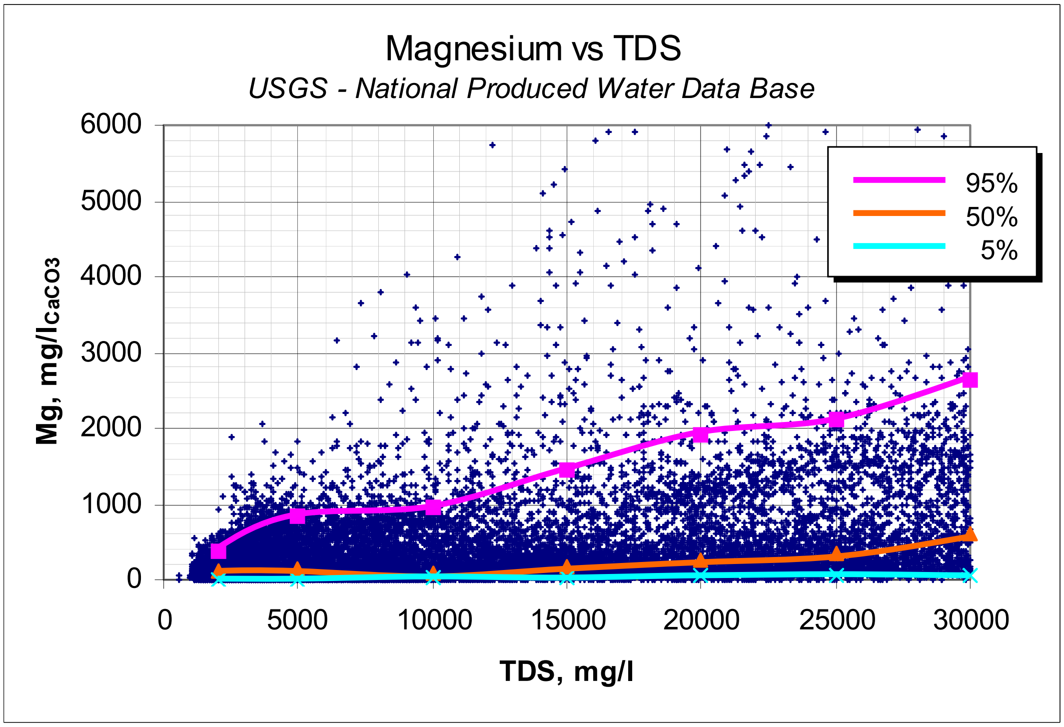

Figure 9-5
Magnesium vs. Total Dissolved Solids Concentrations in Produced Water

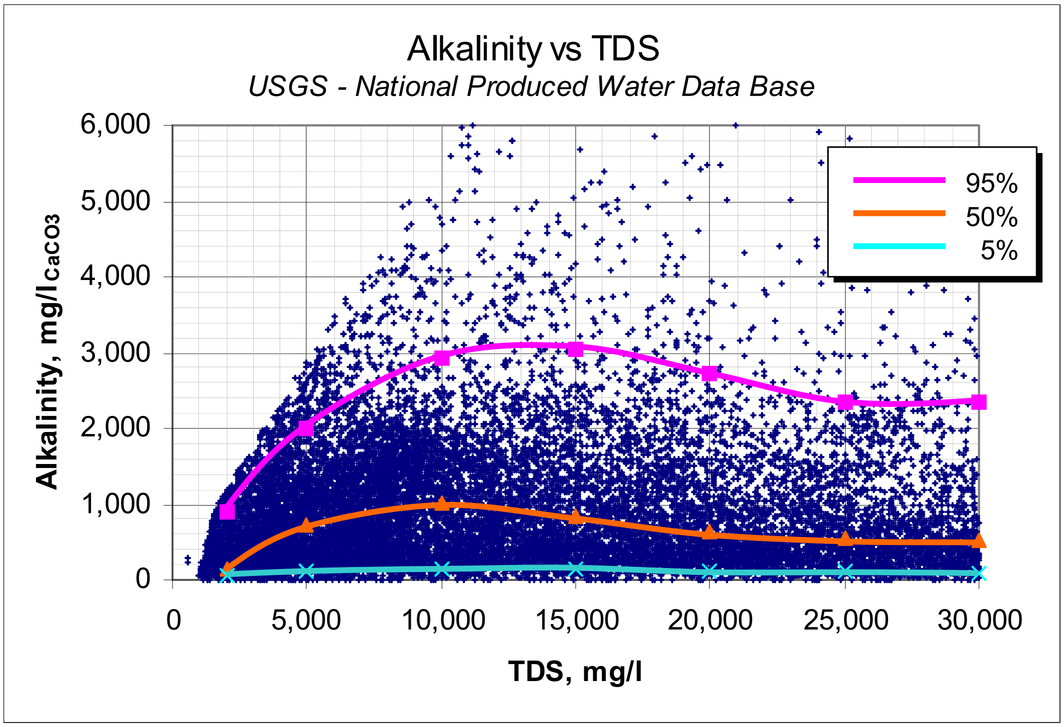

Figure 9-6
Alkalinity vs. Total Dissolved Solids Concentrations in Produced Water

Seven TDS scenarios were established to determine the capital and operating cost of each treatment configuration – 2,000, 5,000, 10,000, 15,000, 20,000, 25,000 and 30,000 mg/l. For each TDS scenario, the data were assessed to find the 95-, 50- and 5-percentile[10] concentrations of calcium, magnesium and alkalinity. These values are roughly equivalent to maximum, mean and minimum values. Table 9-2 presents the data summary for the seven TDS scenarios. For example, in the 10,000 mg/l TDS scenario[11], the 95-percentile calcium concentration was 2,110 mg/l_{CaCO3}, the 50-percentile calcium concentration was 190 mg/l_{CaCO3}, the 5-percentile calcium concentration was 34 mg/l_{CaCO3}.

The maximum concentration (100 percentile) for calcium, magnesium or alkalinity was not used in any of the TDS scenarios, because it was usually very high relative to the 95 percent value. For the 10,000 mg/l TDS scenario, the maximum value for calcium was 6,800 mg/l_{CaCO3} (3.2 times the 95-percentile value). Also, note that the 95-percentile value for calcium was usually 5 to 6 times that of the 50-percentile value (this applies to magnesium and alkalinity but at different levels of intensity). Conversely, the minimum concentrations (0 percentile) for calcium, magnesium and alkalinity were not used either, because all were 0 mg/l_{CaCO3}.

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[10] A 95 percentile value for calcium means that it is greater than 95 percent of all the calcium concentrations in a given TDS range.

[11] The 10,000 mg/l TDS scenario consists of calcium data within the TDS range of 9,001 to 10,000 mg/l. Depending on the scenario, the range was narrow (1,000 mg/l) for high-density areas within the data base and wider (2,000 mg/l) for less dense areas.

Table 9-2
Produced Water Chemistry – Data Summary

TDSmg/l	Percentile Concentrations								
	Ca, mg/l$_{CaCO3}$			Mg, mg/l$_{CaCO3}$			Alk, mg/l$_{CaCO3}$		
	95%	**50%**	**5%**	**95%**	**50%**	**5%**	**95%**	**50%**	**5%**
2,000	950	340	8	370	110	0	910	140	48
5,000	1,690	390	18	830	150	6	1,990	730	120
10,000	2,110	190	34	950	92	25	2,920	1,010	160
15,000	2,650	480	56	1,460	170	36	3,050	860	140
20,000	3,950	700	95	1,910	250	44	2,730	650	120
25,000	5,060	900	150	2,120	340	50	2,360	560	110
30,000	5,550	1,420	160	2,650	620	55	2,350	540	86

The data in Table 9-2 were used to evaluate a number of possible produced water chemistry and flow scenarios and is discussed in the next section.

9.4 Capital and Operating Cost of Produced Water Treatment

The chemistry developed in the previous section is used to assess a number of possible produced water flow and chemistry cases. Three treatment configurations (outlined previously) are evaluated for each TDS scenario and conceptual-level capital and operating costs are developed. Operating cost variations are bracketed to encompass the variability in the USGS database. The technology analysis in this section did not include equipment optimization, because optimization should be conducted when site-specific chemistry data is available.

Finally, as discussed in Section 7, *Cost/Benefit Analysis*, no operating-cost offsets were included in this analysis. For the SJGS produced water project, it was determined that a significant savings could be afforded by some of the producers, and those producers were willing to share the savings with Public Service of New Mexico (PNM). This approach is valid, however, this type of analysis is very site specific and should not be generally applied to all cases.

9.4.1 Capital Cost of Produced Water Treatment

This section presents costs for produced water treatment, de-oiling equipment and pipelines. No attempt was made to predict produced water gathering costs, because they are highly site specific and those costs would likely be borne by oil and gas producers. A number of flow and TDS scenarios were evaluated to determine the capital cost of a produced water project.

Produced Water Treatment Capital Costs

HERO®, BC and evaporation pond costs were factored from data obtained for Section 7, *Cost/Benefit Analysis* and previous work with PNM. Costs for crystallizers were obtained from equipment suppliers, information the author developed in previous work and with PNM. Three treatment configurations were evaluated:

- HERO® + BC
- HERO® + BC + evaporation ponds
- HERO® + BC + crystallizer

Refer to Figures 9-7 through 9-9 for the capital cost of each configuration for a range of feedwater rates (10,000 BPD to 100,000 BPD) and seven different TDS scenarios ranging from 2,000 mg/l to 30,000 mg/l. The costs include equipment and installation plus 25 percent contingency to cover project unknowns. Refer to Table E-1 in Appendix E for capital cost assumptions. Because this analysis is general (not specific to any particular site), costs should be considered “conceptual level” with a +50/-35 percent range of confidence. In other words, the capital costs derived from Figures 9-7 through 9-9 could be 50 percent greater or 35 percent less than the actual cost of installation.

Note that, at produced water TDS levels in excess of 20,000 mg/l, the cost of the equipment in scenarios with BCs and crystallizers jumps notably. In scenarios involving evaporation ponds, the cost variation is not as pronounced. Generally, as HERO® recovery drops at higher TDS levels, BC and crystallizer equipment and evaporation ponds must be sized larger. For example, if produced water TDS were 40,000 mg/l, the BC would be 50 percent larger than a HERO® operating with a feedwater TDS at 30,000 mg/l. For this analysis, the economic TDS limit was established at 30,000 mg/l.

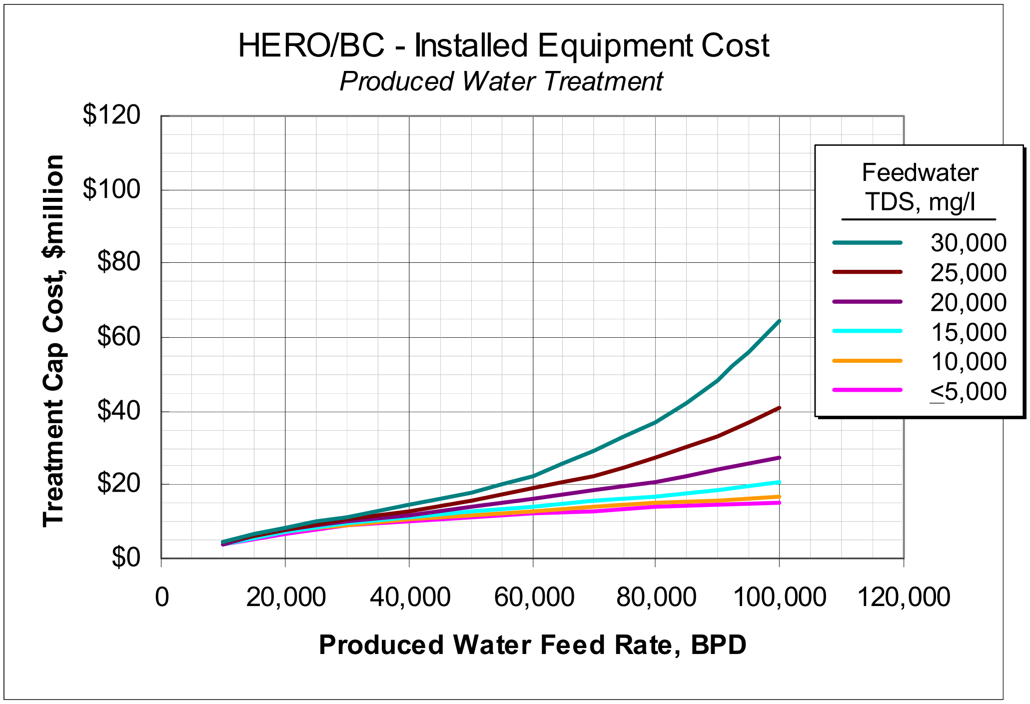

Figure 9-7
Installed Equipment Cost for High-Efficiency Reverse Osmosis and Brine Concentration System

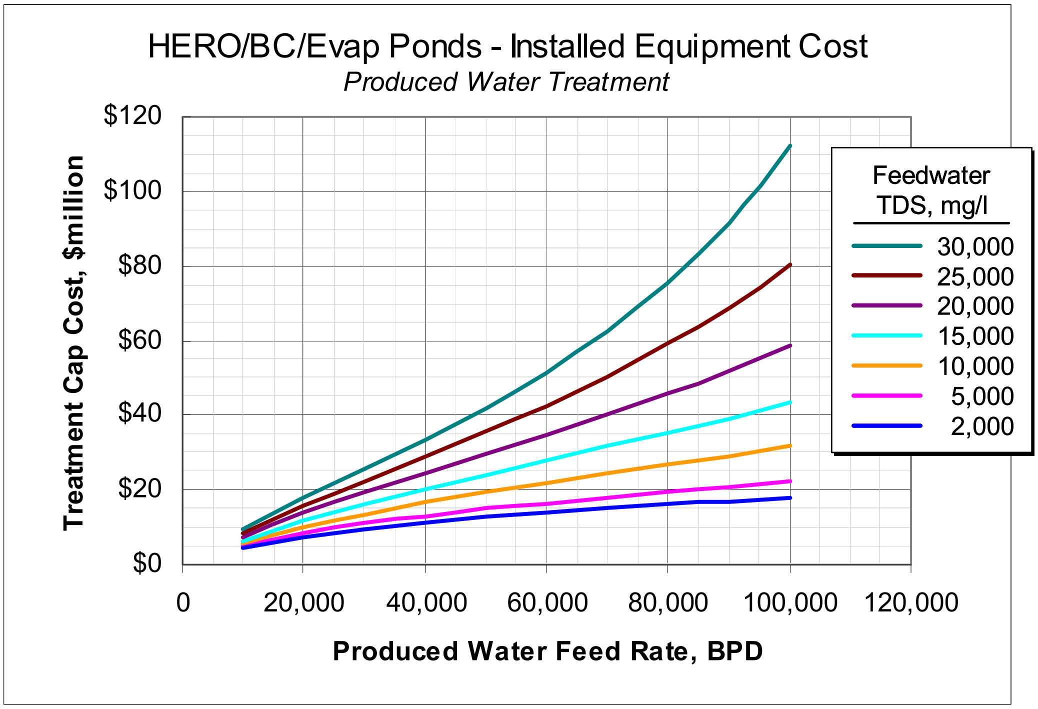

Figure 9-8
Installed Equipment Cost for High-Efficiency Reverse Osmosis, Brine Concentration, and Evaporation Pond System

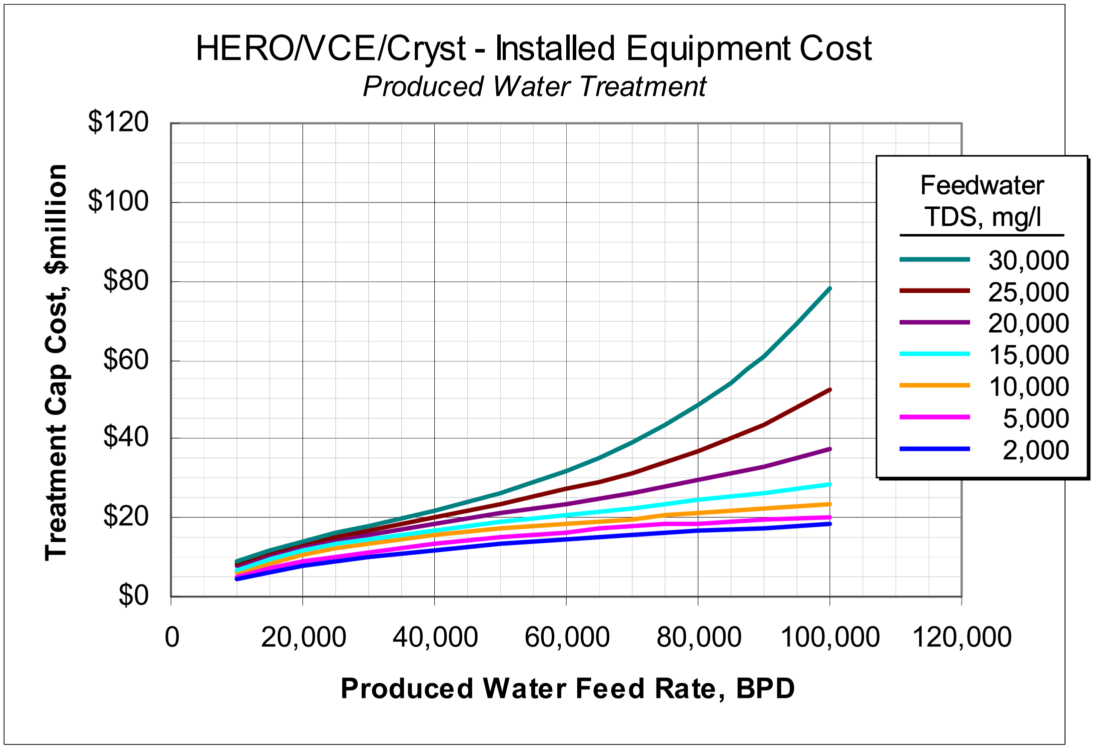

Figure 9-9
Installed Equipment Cost for High-Efficiency Reverse Osmosis, Brine Concentration, and Crystallization System

De-Oiling and Filtration Capital Costs

De-oiling equipment is only applicable to conventional oil and gas production in this analysis. Refer to Section 4.5, *Collection Center in Bloomfield* and Figure 4-10 for a process description and schematic for de-oiling equipment. The only exception would be covered tanks instead of the open basins proposed for SJGS. Some produced water could create a safety problem (and public nuisance) because of elevated levels of hydrogen sulfide gas (H_2S).[12] The occurrence of H_2S is highly site specific and cannot be predicted from the information in the USGS database.

CBM produced water is free of oily byproducts found in conventionally produced water, but typically contains coal fines. For this analysis, the process schematic would be similar to de-oiling but without gravity separation, oil recovery, gas flotation and off-spec produced water management.

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[12] Open basins were acceptable for the SJGS produced water project because H2S is typically at non-detectable levels.

Refer to Figure 9-10 for de-oiling equipment (conventional production) costs and filtration equipment (CBM production) costs.[13] Lastly, it is assumed that the de-oiling or filtration equipment is located at the produced water treatment plant.

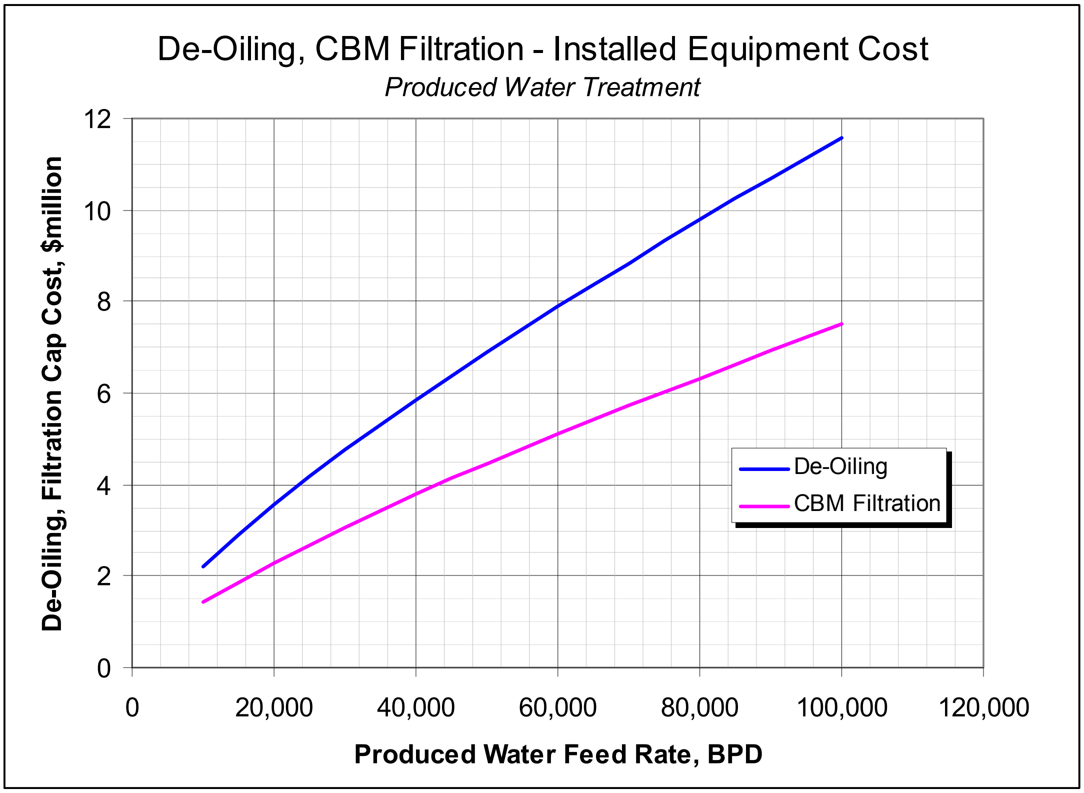

Figure 9-10
Installed Equipment Cost of Produced Water Deoiling and Coal Bed Methane Water Filtration Systems

Pipeline Capital Costs

Refer to Figure 9-11 for an estimate of pipeline costs. Three scenarios are presented – one, five and ten-mile pipelines. To simplify the analysis, the pipelines were assumed to be over flat terrain (no intermediate pump stations), constructed with HDPE[14] and operated at a relatively low pressure (to accommodate the HDPE). The pipeline headworks would consist of two tanks capable of holding 12 hours of daily inflow, one to three clean-out stations (pigging equipment), and a pump station to charge the line.

Cost criteria developed for the SJGS produced water pipeline were used in this analysis. For SJGS, it was determined that a pipeline would cost from $6.00 to $9.00 per inch-diameter per linear foot depending on the route. An average value of $7.50/inch-diameter/foot was used in

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[13] The costs for de-oiling and filtration equipment are not effected by produced water TDS.

[14] HDPE is high-density polyethylene – plastic pipe used for low-pressure corrosive-water service.

this analysis. The step-features of the cost lines are a result of line-size changes, i.e. the diameter of the line was increased at higher flow rates to minimize pressure drop. Costs were developed separately for the collection tanks and pump station (located at the head works) and were incorporated into the graphical analysis.

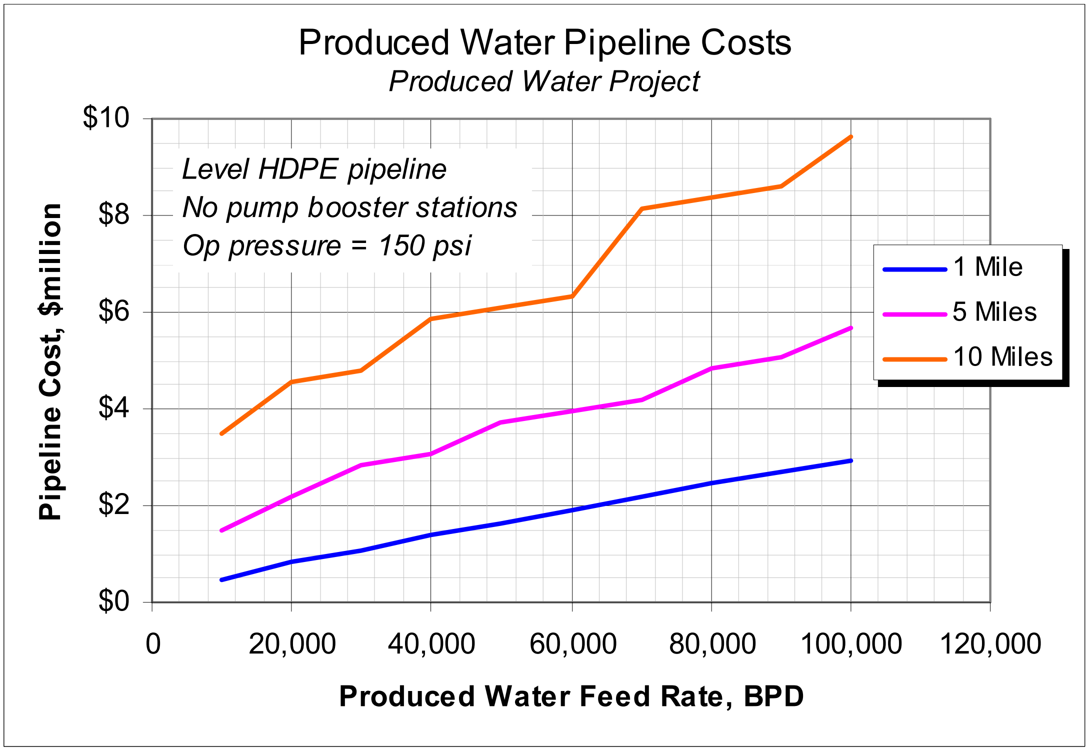

Figure 9-11
Produced Water Pipeline Costs

9.4.2 Operating Cost of Produced Water Treatment

For each of the seven TDS scenarios, 27 combinations of calcium, magnesium and alkalinity concentrations were assessed.[15] The chemistry derived from the USGS Produced Waters Database and presented in Table 9-2 provided the basis for the analysis. As stated previously, this analysis was designed to determine the performance and operating cost of a reactor clarifier. Since calcium, magnesium and alkalinity concentrations are lowered in a reactor clarifier, TDS was adjusted[16] to predict HERO® recovery and subsequently size the BC, crystallizer and evaporation ponds.

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[15] Three constituents (calcium, magnesium and alkalinity) by three concentrations (95-, 50- and 5-percentile) for a total of 27 combinations.

[16] When softening occurs in a reactor clarifier, effluent concentrations for calcium, magnesium and alkalinity are lowered, and depending on the chemicals used, sodium can increase. For each case within a scenario, TDS was recalculated. Then the 27 values were averaged to determine adjusted TDS (used to calculate HERO® recovery). This averaging method, although it reduces the case-by-case variability in the adjusted TDS, is more representative than the unadjusted value.

The chemical costs for the reactor clarifier, which typically dominate other chemical costs, were also averaged and added to the cost of other chemicals, power, membrane replacement, cleaning (RO membranes, BC internal surfaces and crystallizer internal surfaces as applicable), sludge/solids handling and onsite disposal, labor, and maintenance. Refer to Figures 9-12 through 9-14. Staffing to operate and maintain the treatment plant was also adjusted (to determine labor costs) based on the size of the plant. Refer to Table E-1 in Appendix E for operating cost assumptions.

The operating costs in Figures 9-12 through 9-14 do not include capital recovery costs. These costs were purposely left out to show how throughput capacity and TDS affect unit operating costs. Additionally, since there is no standard method to determine capital recovery, this calculation is left to the reader.

Unit operating costs are expressed as dollars per barrel ($/bbl). Therefore, in Figure 9-12, for a 50,000 BPD plant with a produced water TDS of 10,000 mg/l, the unit operating cost would be $0.14/bbl to operate a HERO® and BC. This translates to an operating cost of $7,000 per day (50,000 BPD x $0.14/bbl) or $2,555,000 per year. The costs include chemicals, power, membrane replacement, HERO® and BC cleaning, reactor-clarifier sludge handling and onsite disposal, labor, and maintenance.

Figures 9-15 and 9-16 present cost factors showing the impacts of total dissolved solids (TDS) concentration on the calculated operating cost. The cost factor ranges represent the variations between the 5- and 95-percentile calcium, magnesium and alkalinity concentrations. For these charts, a cost factor of 1.0 is equivalent to the operating costs found in Figures 9-12 through 9-14 (~50-percentile values). For the same example, the minimum and maximum operating cost factors from Figure 9-15 are 0.63 and 2.35, respectively. This translates to an operating cost range of $0.09/bbl ($0.14/bbl x 0.63) to $0.33/bbl ($0.14/bbl x 2.35). If the calcium, magnesium and alkalinity are known, the operating cost range could be roughly interpolated. It is prudent to apply the cost factors to general data until site-specific information can be assessed.

Lastly, the cost range is large because of the significant degree of calcium, magnesium and alkalinity variation in the USGS database. It should be noted that 50-percentile (mean) concentrations are much closer to the 5-percent concentrations than 95-percentile. Again, site-specific chemistry is required to rigorously evaluate treatability and costs. The approach developed here can be used to <u>conceptually bracket operating costs</u>.

De-Oiling and Filtration Operating Costs

The unit operating cost for this analysis[17] for de-oiling conventional oil and gas produced water is $0.035/bbl. The calculated values over the range of feedwater throughput vary little from a small to large de-oiling systems. Refer to Table E.1 in Appendix E for operating cost assumptions. The unit cost includes power, maintenance, chemicals and offsite transportation and disposal of off-spec produced water. Because of the unknowns, no recovered-oil credit was taken. Note that off-spec produced water disposal comprises 40 percent of the operating cost.

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[17] The analysis incorporated most of the assumptions used for the Bloomfield Collection Center for the SJGS produced water project. Refer to Section 4.5.

The unit operating cost for CBM water filtration is $0.014/bbl (applicable to small and large systems as well). Labor for de-oiling and CBM filtration was included in the produced water treatment plant staffing assumptions.

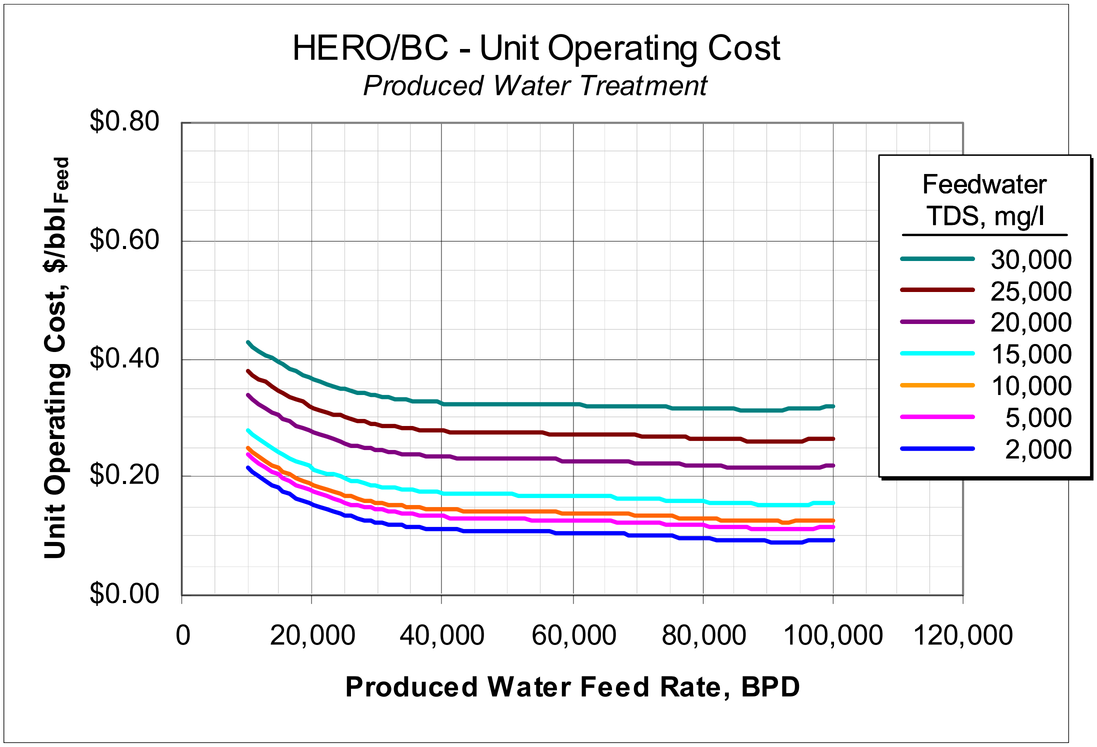

Figure 9-12
Unit Operating Cost of High-Efficiency Reverse Osmosis and Brine Concentration System

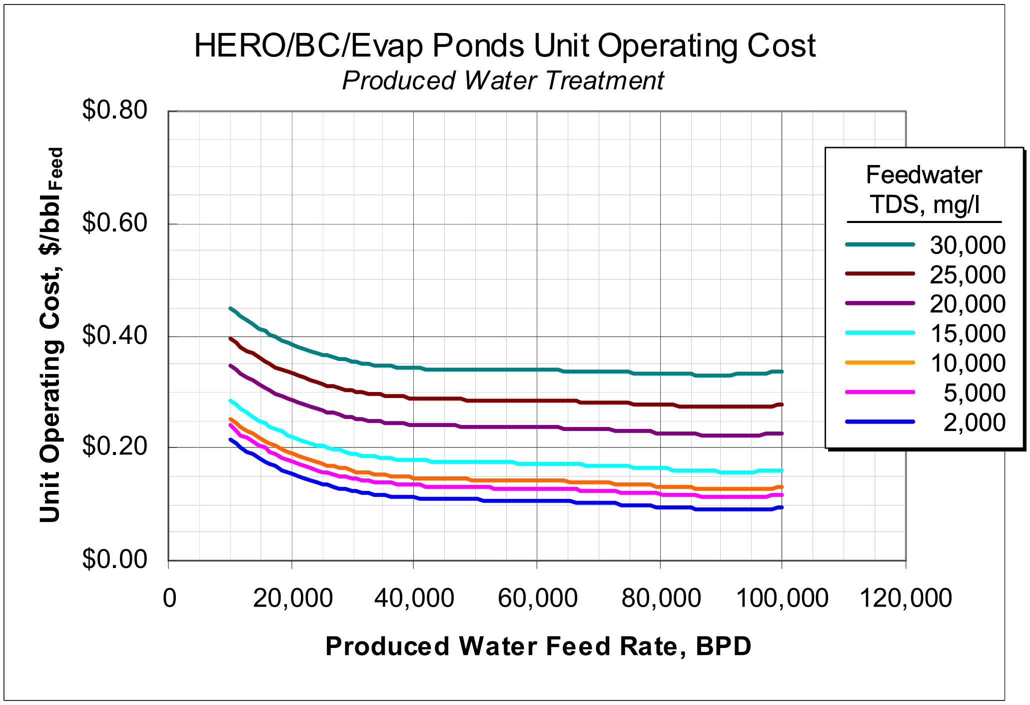

Figure 9-13
Unit Operating Cost of High-Efficiency Reverse Osmosis, Brine Concentration, and Evaporation Pond System

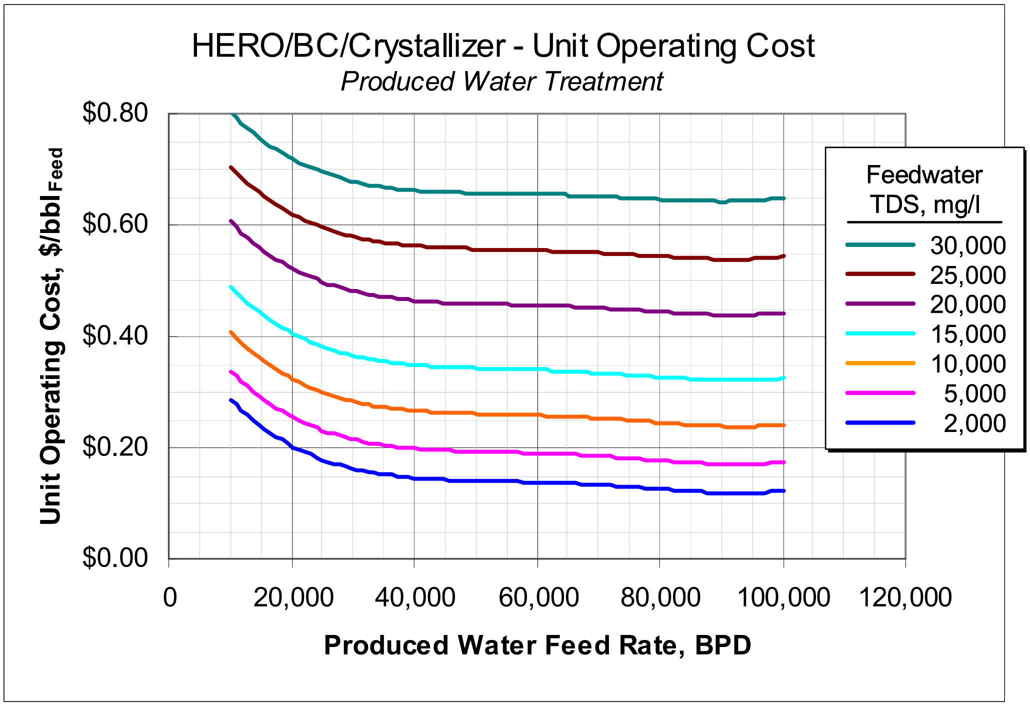

Figure 9-14
Unit Operating Cost of High-Efficiency Reverse Osmosis, Brine Concentration, and Crystallization System

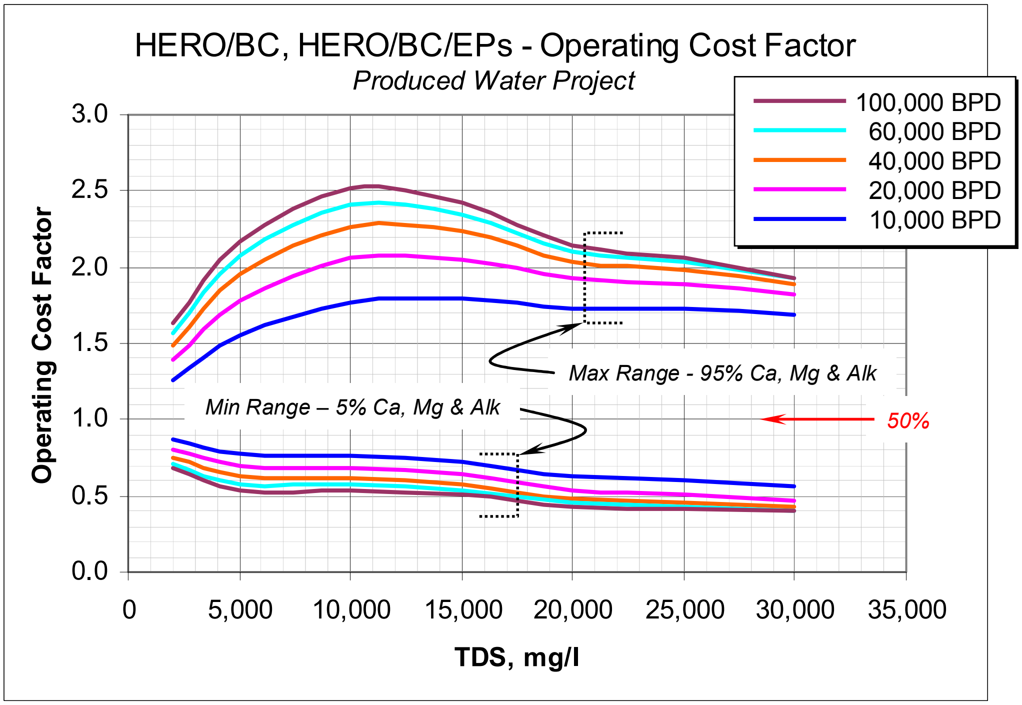

Figure 9-15
TDS Sensitivity Factors for High-Efficiency Reverse Osmosis/Brine Concentration (HERO/BC) and HERO/BC/Evaporation Pond Systems

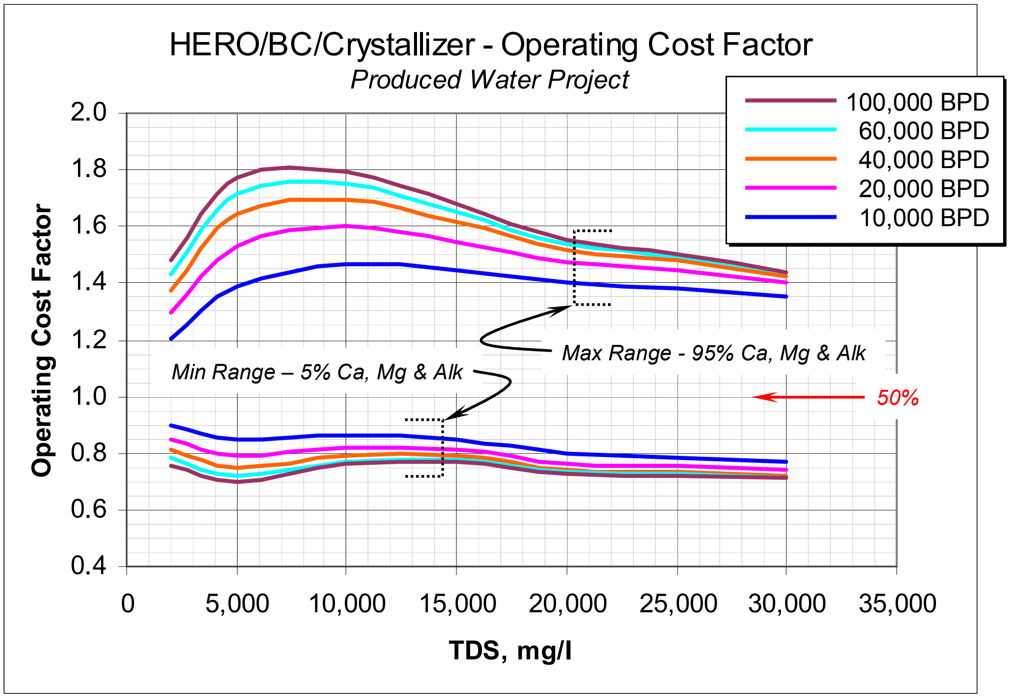

Figure 9-16
TDS Sensitivity Factors for High-Efficiency Reverse Osmosis, Brine Concentration, and Crystallization System

Pipeline Operating Costs

Pipeline operating costs are presented in Figure 9-17 (the analysis was smoothed with a curve fitting tool). The costs include pumping power and maintenance. Refer to Table E-1 in Appendix E for operating cost assumptions. Point-to-point cost variation is high in this analysis as a result of pipeline charging pressure. Line size selection and flow rate have a significant effect on pipeline pressure drop since transitions to larger diameter lines sizes are step-like and not smooth. Pipeline labor was included in the produced water treatment plant staffing assumptions.

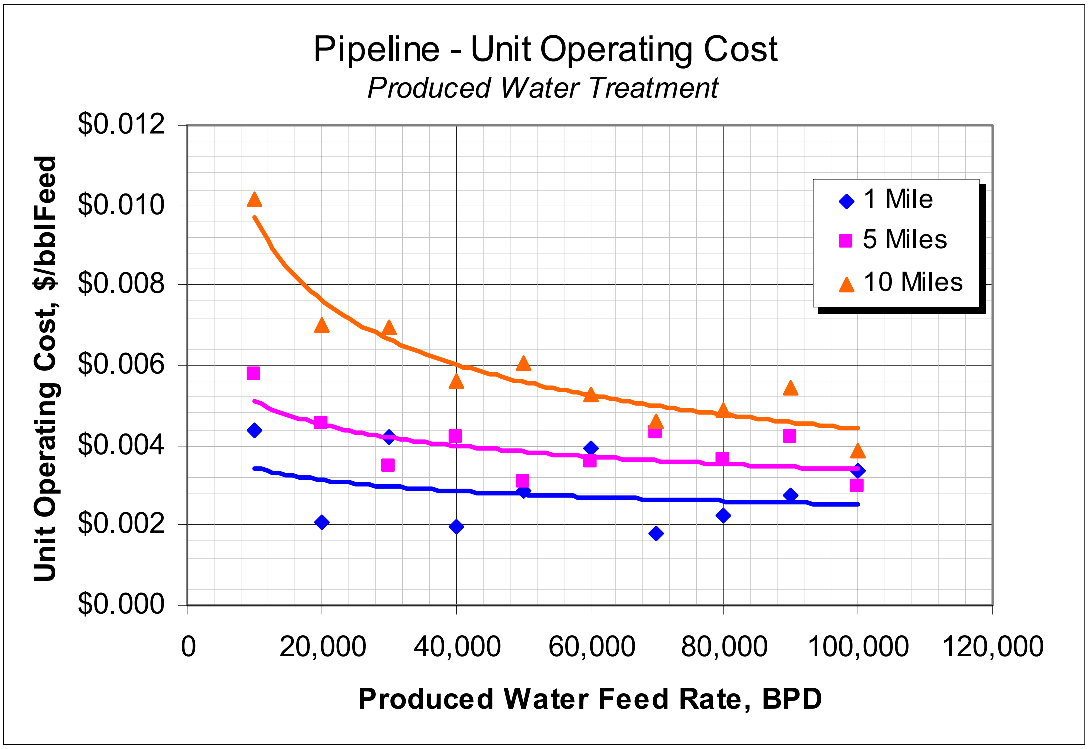

Figure 9-17
Unit Operating Cost of Pipeline

9.5 Plant Examples

Two plant examples are presented to show how the cost estimating charts could be used to evaluate conceptual-level produced water capital and operating costs.

9.5.1 Site 1 Example

A coal-fired power plant in the Southwest is approximately 7.5 miles from conventional oil production. The plant has an opportunity to treat and use 60,000 BPD of produced water with a TDS of 12,000 mg/l that would otherwise be disposed via injection. Assume that the existing de-oiling equipment (operated by the producers) is quite old and unreliable, so new equipment would be installed with the produced water treatment plant. The power plant has also determined that wastewater generated by produced water treatment must be sent to an evaporation pond. Table 9-3 describes the capital and operating cost elements of the analysis. Total installed cost is projected to be $37,200,000 for the produced water treatment plant, de-oiling equipment and a pipeline. Recall that the capital cost should be considered “conceptual level” with a +50/-35 percent range of confidence. Operating costs are expected to be within a range of $0.128/bbl to $0.426/bbl – this cost will be a function of produced water quality. Operating costs include chemicals, power, membrane replacement, equipment cleaning, maintenance and labor. Recall that the operating cost does not include capital recovery.

Table 9-3
Cost Analysis – Example 1

Design Basis			
Throughput	60,000 BPD		
Produced Water TDS	12,000 mg/l		
Distance to Source	7.5 miles		
Ultimate Disposal	Evaporation Pond		
Installed Cost Analysis			
Produced Water Treatment (Figure 8.8)	$24,000,000		
De-Oiling (Figure 8.10)	$8,000,000		
Pipeline (Figure 8.11)	$5,200,000		
Total Installed Cost (1)	$37,200,000		
Unit Operating Cost			
Mean Produced Water Treatment (Figure 8.13)	$0.16/bbl		
Cost Variation Factors (Figure 8.15)	0.55 (5 percentile)	2.41 (95 percentile)	
	Min	**Mean**	**Max**
Produced Water Treatment	$0.088/bbl	$0.160/bbl	$0.386/bbl
De-Oiling (same for all scenarios)	$0.035/bbl	$0.035/bbl	$0.035/bbl
Pipeline (Figure 8.17)	$0.005/bbl	$0.005/bbl	$0.005/bbl
Total Unit Operating Cost ($\$/\text{bbl}_{\text{Feed}}$)	$0.128/bbl	$0.200/bbl	$0.426/bbl
Annual Operating Cost (2)	$2,800,000	$4,380,000	$9,330,000

Notes:

Recall that the capital cost should be considered “conceptual level” with a +50/-35 percent range of confidence.

Does not include capital recovery costs.

Note, if the calcium, magnesium and alkalinity concentrations in the produced water were determined to be close to the mean values found in Table 9-2 (or Figures 9-4 through 9-6), the operating cost would be close to \$0.200/bbl. Therefore, knowing basic site-specific chemistry can be useful in narrowing the range of the operating costs by roughly interpolating the cost factor in Figures 9-15 and 9-16.

9.5.2 Site 2 Example

A coal-fired power plant in a Rocky Mountain state is approximately 2.5 miles from CBM production. They have an opportunity to treat and use 40,000 BPD of produced water with a TDS of 6,000 mg/l that would otherwise be disposed of. Assume that the existing filtration equipment (operated by the producers) is quite new, so filters would not be installed at the produced water treatment plant. The power plant has also determined that produced water treatment wastewater must be sent to crystallizers. The dried waste would be landfilled along with scrubber sludge. Table 9-4 describes the capital and operating cost elements of the analysis. Total installed cost is projected to be \$15,000,000 for the produced water treatment plant and a pipeline. Operating costs are expected to be within a range of \$0.169/bbl to \$0.371/bbl.

Again, if the calcium, magnesium and hardness concentrations in the produced water were determined to be close to the mean values found in Table 9-2, the operating cost would be close to \$0.226/bbl.

Table 9-4
Cost Analysis – Example 2

<table>
<tr><th colspan="4">Design Basis</th></tr>
<tr><td>Throughput</td><td colspan="3">40,000 BPD</td></tr>
<tr><td>Produced Water TDS</td><td colspan="3">6,000 mg/l</td></tr>
<tr><td>Distance to Source</td><td colspan="3">2.5 miles</td></tr>
<tr><td>Ultimate Disposal</td><td colspan="3">Crystallizer</td></tr>
<tr><th colspan="4">Installed Cost Analysis</th></tr>
<tr><td>Produced Water Treatment (Figure 8.9)</td><td colspan="3">$13,000,000</td></tr>
<tr><td>De-Oiling (Figure 8.10)</td><td colspan="3">N/A</td></tr>
<tr><td>Pipeline (Figure 8.11)</td><td colspan="3">$2,000,000</td></tr>
<tr><td>Total Installed Cost (1)</td><td colspan="3">$15,000,000</td></tr>
<tr><th colspan="4">Unit Operating Cost</th></tr>
<tr><td>Mean Produced Water Treatment (Figure 8.14)</td><td colspan="3">$0.22/bbl</td></tr>
<tr><td>Cost Variation Factors (Figure 8.15)</td><td colspan="2">0.75 (5 percentile)</td><td>1.67 (95 percentile)</td></tr>
<tr><td></td><td>Min</td><td>Mean</td><td>Max</td></tr>
<tr><td>Produced Water Treatment</td><td>$0.165/bbl</td><td>$0.220/bbl</td><td>$0.367/bbl</td></tr>
<tr><td>De-Oiling (same for all scenarios)</td><td>N/A</td><td>N/A</td><td>N/A</td></tr>
<tr><td>Pipeline (Figure 8.17)</td><td>$0.004/bbl</td><td>$0.004/bbl</td><td>$0.004/bbl</td></tr>
<tr><td>Total Unit Operating Cost ($\$/\text{bbl}_{\text{Feed}}$)</td><td>$0.169/bbl</td><td>$0.226/bbl</td><td>$0.371/bbl</td></tr>
<tr><td>Annual Operating Cost (2)</td><td>$2,470,000</td><td>$3,300,000</td><td>$5,360,000</td></tr>
</table>

Notes:

1. Recall that the capital cost should be considered “conceptual level” with a +50/-35 percent range of confidence.
2. Does not include capital recovery costs.

10

WET SURFACE AIR COOLING (WSAC) TEST AT SAN JUAN GENERATING STATION

10.1 Introduction

Wet surface air cooling (WSAC) technology was tested at the San Juan Generating Station (SJGS)[1] to determine its capacity to cool power plant circulating water using degraded water. WSAC is a commercial cooling technology and has been used for many years to cool and/or condense process fluids. In the WSAC, water is applied in dense spray patterns to the exterior of tubes carrying the liquid or gas to be cooled. At the same time, air is also drawn down and around the perimeter of the tubes in the same direction as spray water. What is unique about this type of cooling is that the dense flow of spray water continuously flushes the tube surfaces, preventing suspended matter and mineral scale from depositing. Scale and suspended solids eventually settle in the WSAC basin or on non-cooling surfaces. Lastly, WSAC operates similarly to a mechanical draft cooling tower in that it utilizes air to evaporate water to provide cooling.

The purpose of the pilot test was to determine if WSAC technology could cool process water at cycles of concentration considered excessive (highly scale forming) for mechanical draft cooling towers. To accomplish this, blowdown from the Unit 3 Cooling Tower at SJGS was fed to the WSAC pilot as make-up. An additional benefit of operating in this mode is that WSAC could also function as a concentrating device by reducing the volume of a large plant wastewater stream.[2]

The WSAC pilot was online for a total of 147 days – from July 5 to November 29, 2005. During this period, it was in service for 2,898 hours. It was configured to cool circulating water from Unit 3, and at the same time, use Unit 3 circulating water for make-up.[3] In this mode, the WSAC operated at an equivalent of 24 to 70 cycles of concentration (based on freshwater fed to SJGS cooling towers). Ten cycles of concentration is considered a safe limit at this plant to control

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[1] SJGS is four-unit 1500-MW coal-fired plant in the Four Corners Area of New Mexico and is operated by Public Service of New Mexico (PNM).

[2] SJGS is a zero-liquid discharge plant. The plant utilizes brine concentrators (wastewater evaporators) to reduce wastewater volume prior to disposal in evaporation ponds. Cooling tower blowdown is usually the largest wastewater stream at power plants.

[3] A small fraction of cooling tower circulating water is bled from the cooling system, as blowdown, to control salt build-up.

mineral scale formation. At the completion of testing, there was no visible scale on the heat transfer surfaces (tube externals) and cooling was sustained throughout the test period.

The report is presented in the following four sections – Pilot Design Parameters, Data Collection and Controls, Test Data and Summary of Findings. Section 10.2, Pilot Design Parameters, contains a detailed description of how WSAC technology cools water along with a process schematic showing flow paths for cooling water, air and process fluids. A description of the Unit 3 Cooling Tower at SJGS – the source of water for the testing – is also provided. Section 10.3, Data Collection and Controls, outlines WSAC instrumentation and design elements, measurement and analytical test parameters, and the operating approach for the testing. Section 10.4, Test Data, covers the flow, temperature and chemistry data collected during WSAC testing. The results of a corrosion analysis are presented at the end of this section. Section 10.5, Summary of Findings, summarizes observations, findings and conclusions as well as unresolved issues and recommended follow-on study.

10.2 Pilot Design Parameters

This section of the report describes the parameters that were used to design and operate the WSAC pilot equipment. The section begins with a more detailed description of how WSAC technology cools water along with a process schematic showing flow paths for cooling water, air and process fluids (to be cooled). A process flow schematic is provided that shows the pilot test equipment, metallurgy, instrumentation, etc. Pictures of the pilot equipment are also provided. Lastly, a description of the Unit 3 Cooling Tower at SJGS – the source of water for this phase of testing – is provided.

10.2.1 WSAC Technology

WSAC technology has been in commercial use for many years. There is one large WSAC installation in Massachusetts cooling a 150 MW power plant and there are thousands of smaller installations throughout the world.

Warm water enters the WSAC and passes through a heat exchanger (inside the tubes). Refer to Figure 10-1. WSAC cooling water is deluged via spray nozzles (high flow, low pressure) on the outside of the exchanger tubes. Spray nozzles are mounted on headers and located above the tube bundles (one header per bundle). Cooling air flows downward over the tubes in the same direction as the deluge water. The air is drawn through and around the tubes, through a demister (for drift control) and to the fans where it is exhausted. Cooling water is collected in a basin directly beneath the tube bundles and circulated back to the nozzles to be sprayed again. Cooling occurs as a small fraction of circulating water evaporates (into the air stream) while deluging the tubes and falling to the basin, i.e. evaporating water extracts heat from the system in the same manner as a mechanical draft cooling tower. As water evaporates, make-up water is fed to the WSAC to compensate for the loss, and at the same time, blowdown is withdrawn to control chemistry.

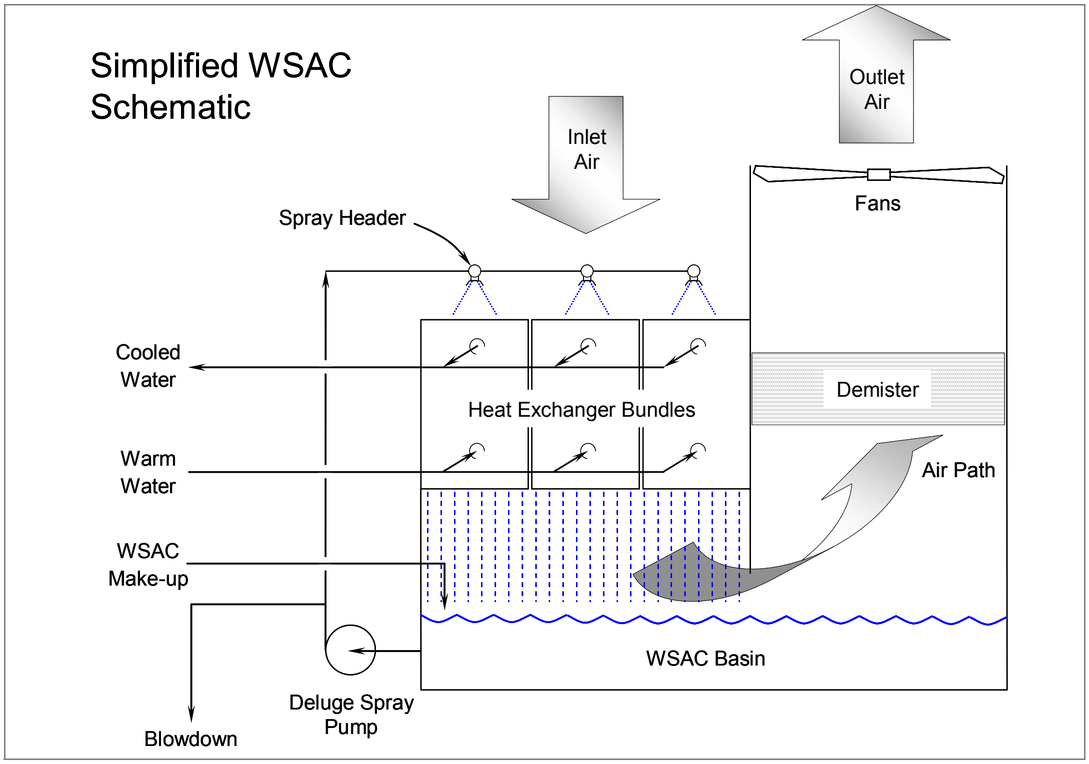

Figure 10-1
Schematic of the Wet-Surface Air Cooling Pilot Unit (WSAC)

The deluge spray configuration will in theory allow cooling water to be cycled much higher than in conventional mechanical draft cooling towers as described below.

- In a cooling tower, droplets and films of circulating water are encouraged to form. Cooling takes place on surfaces where films of water are formed (cooling tower packing) and on droplets in the freefall to the basin[4]. If the chemistry of the circulating water exceeds saturation limits, crystals of scaling salts (e.g. calcium sulfate) will start to nucleate. The nucleating crystals will adhere to rough surfaces in low-flow regions throughout the cooling circuit <u>including the condenser tubes</u>.
- The deluge water in a WSAC is in a separate circuit from the circulating water that cools the condensers, i.e. deluge water is sprayed onto the outside of the tube bundles that carry condenser circulating water. The WSAC is designed to spray relatively large volumes of water on the exterior surfaces of the tube bundles. The constant movement of the deluge water on the surfaces of tubes keeps scale (that may form) in the bulk spray water.

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[4] A small fraction of the circulating water evaporates into the air stream. As the water evaporates, it extracts heat from the circulating water. In the design of mechanical draft cooling towers, an effort is made to expose as much of the surface of the circulating water as possible (in the form of films and droplets) to maximize intimate air-water contact.

10.2.2 Pilot Test System Design Parameters

A pilot WSAC was built by Niagara Blower Company and delivered to SJGS in late May of 2005. It went through shakedown and was started on June 14, 2005. Testing started on July 5, 2005. This time was necessary to set-up WSAC sampling, instrumentation and equipment check-in schedules with plant laboratory and maintenance staff. Also, pH and conductivity instruments had to be installed and the data logger purchased and installed.

Three tube materials were selected – 316 stainless steel, 90-10 copper-nickel (Cu-Ni) and titanium (Grade 2) – to evaluate tube corrosion potential.[5] Stainless steel was selected because it is susceptible to chloride stress corrosion and would provide a good baseline as compared to more corrosion-resistant and costly metals. Copper-nickel is a metallurgy frequently used for once-through cooling at seawater and brackish-water power plants. Titanium is highly resistant to corrosion and is commonly used for condenser tubes in recently installed cooling systems using brackish or degraded water.

Table 10-1 summarizes the design specifications of the WSAC pilot unit.

Table 10-1
Pilot Wet Surface Air Cooler Design Specifications

Heat load, BTU/hour	3,000,000		
Flow Rate, gpm	260		
Spray Rate, gpm	650		
Spray Pump Motor, HP	7.5		
Spray Nozzles (total)	36		
Design Wet Bulb, °F	66		
Temperature In, °F	103		
Temperature Out, °F	80		
Tube Bundles	3		
Flow Passes	12		
Tube Material	316 SS (east bundle)	90-10 Cu-Ni (center bundle)	Ti Grade 2 (west bundle)
Tube OD, inches	1.25	1.25	1.25
Wall Thickness, inches	0.049	0.049	0.035
Demister Rating	0.005%		
Fans	3		
Total Fan Capacity, cfm	62,000		
Fan Power, HP	7.5 (each)		

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[5] It was not expected that cooling tower blowdown from Unit 3 would be a problem for any of the metallurgies selected. The metallurgy was selected for possible follow-on testing using a more saline degraded water.

Refer to Figures 10-2 and 10-3 for end and side elevations of the pilot equipment. Water from the Unit 3 Cooling Tower riser (water to be cooled) was fed the WSAC bundles (bottom) by way of a header. The unit was designed such that any or all of the bundles could be isolated, i.e. receive no flow. Cooled water exited through the top header and was returned to the Unit 3 Cooling Tower basin. Make-up to the WSAC was also Unit 3 circulating water and was supplied by the same feed line (from the riser at Unit 3). The spray pump, which was used to circulate the WSAC cooling water, can be seen in Figure 10-3.

Figures 10-4 and 10-5 are photographs of the WSAC pilot unit. Figure 10-6 shows the valve nest for inlet and outlet lines to the tube bundles. Figure 10-7 shows the WSAC circulating water (spray) pump with the Unit 3 Cooling Tower in the background. Tube bundles and spray headers can be seen in Figure 10-8.

10.2.3 Unit 3 Cooling Tower

The Unit 3 Cooling Tower provides cooling for a 550 MW coal-fired generating unit at SJGS. It cools 220,000 gpm of water and operated as follows during pilot testing:

Parameter	Range During Testing
Cycles of Concentration	5.6 to 7.9 (average = 7.0)
TDS, mg/l	2,400 to 3,000
Calcium, mg/l	200 to 420
Chloride, mg/l	75 to 135
Sulfate, mg/l	575 to 1,350
Silica, mg/l	20 to 60
Control pH Range	7.5 to 8.1

Continuous chlorination is used to maintain a residual of 0.1 to 0.2 mg/l_{Cl2} in the cooling tower circulating water. A phosphonate-type inhibitor is used for scale control[6] and an azole-type corrosion inhibitor is used to protect the copper-based admiralty brass metallurgy of the main condenser.

10.3 Data Collection and Controls

The WSAC pilot was designed to monitor a number of performance parameters and control pH and conductivity. This section outlines WSAC instrumentation and design elements, measurement and analytical test parameters, and the operating approach for the testing.

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[6] Calcium sulfate is the limiting scale-forming salt in the cooling towers at the San Juan Generating Station. Unit 3 Cooling Tower operates at 70 percent of calcium sulfate saturation. Calcium carbonate scale formation is controlled by pH.

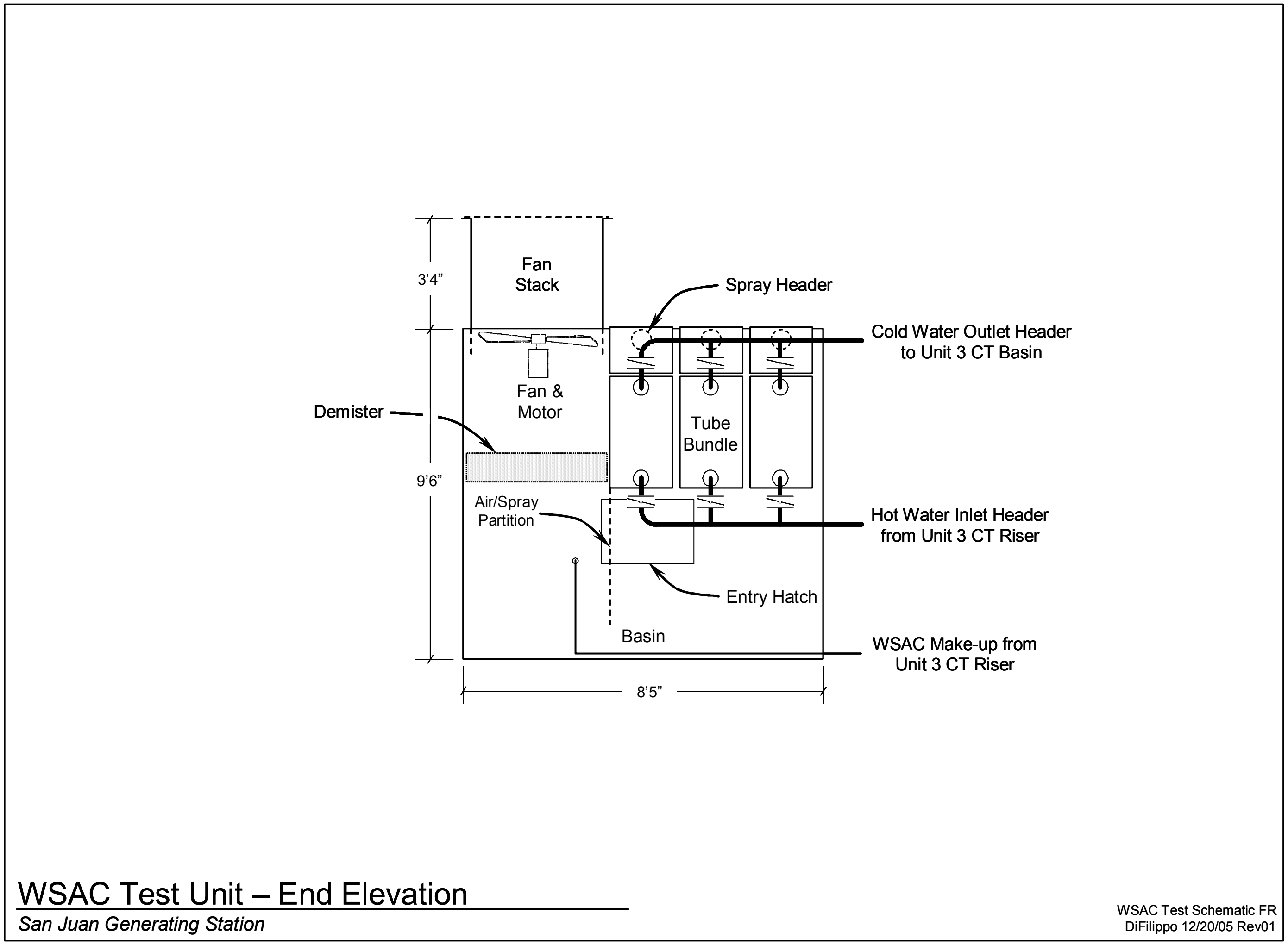

Figure 10-2
End Elevation of WSAC Test Unit

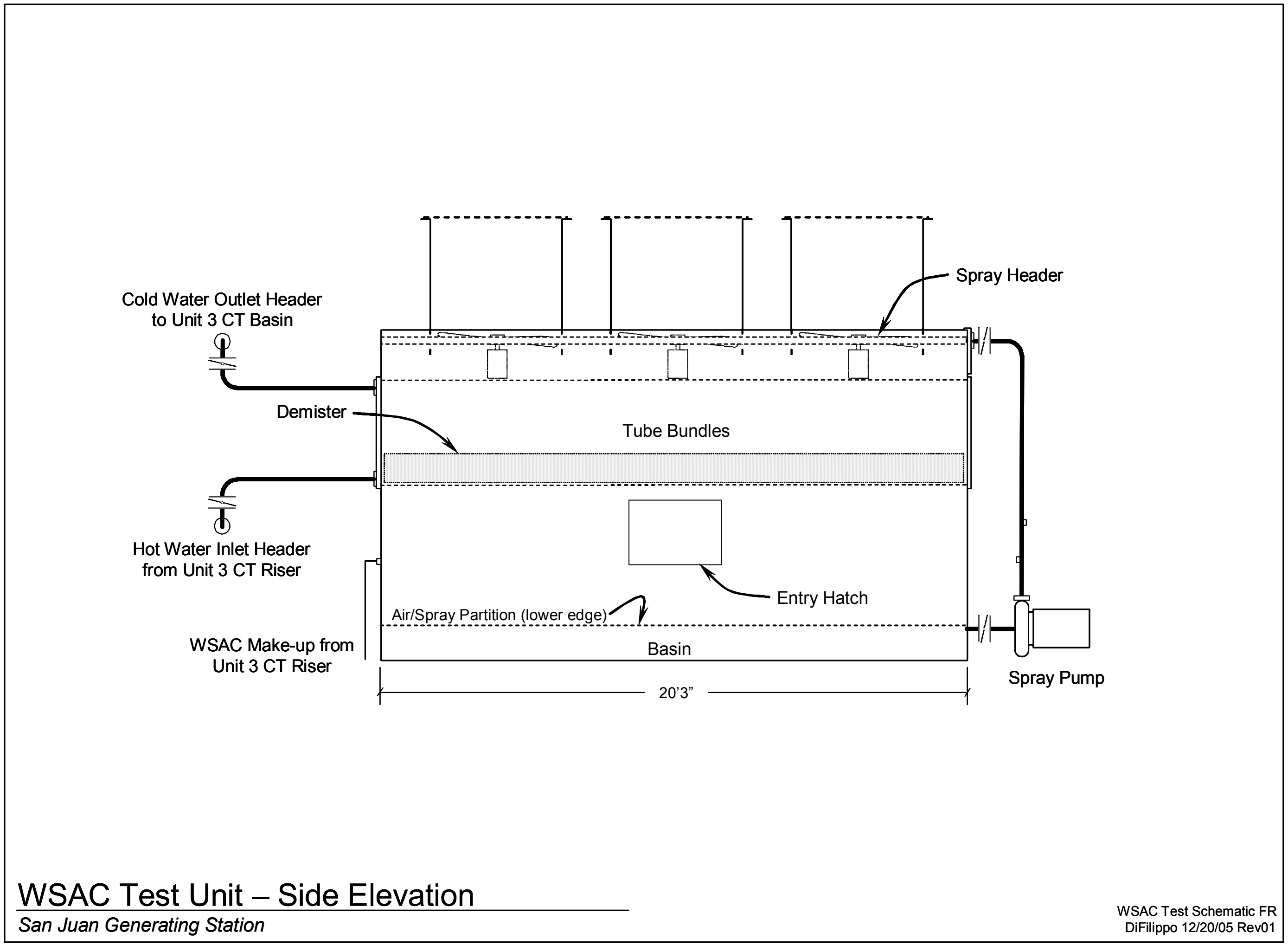

Figure 10-3
Front Elevation of WSAC Test Unit

Figure 10-4
WSAC Pilot Unit and Monitoring Shed

Figure 10-5
Inlet-Outlet Piping at WSAC Pilot Unit

Figure 10-6
Inlet-Outlet Valving at WSAC Pilot Unit

Figure 10-7
Spray Pump at WSAC Pilot Unit

Figure 10-8
Spray Headers and Tube Bundles at WSAC Pilot Unit

10.3.1 Instrumentation

Refer to Figure 10-9 for a process flow diagram (PFD) of the WSAC pilot. The PFD schematically shows the analytical probes, elements and instrumentation used for the test. The following parameters were monitored at ten-minute intervals throughout the testing:

- Inlet and outlet temperature (°F) of each tube bundle
- Skin temperature (°F) of two tubes in each bundle – top and mid-level
- Skin temperature of three dummy[7] tubes located at bottom of each tube bundle (a heating element was installed in these tubes to monitor heat flux)
- Hot-water flow rate (gpm) to each bundle
- Make-up water flow rate (gpm) to the WSAC (flow was also totalized)
- WSAC blowdown rate (gpm) (flow was also totalized)

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[7] A dummy tube was located below (and center to) each tube bundle. The dummy tube was open at each end so a heating element could easily be inserted to monitor a fixed rate of heat transfer.

- WSAC spray water pH
- WSAC spray water conductivity (µS/cm, Micro-Siemans per centimeter)
- Three removable tube sections (one per bundle) for metallurgical analysis at the completion of the testing

Flow, temperature, conductivity and pH were recorded every ten minutes during the operation of the WSAC. Data were stored on a data logger and downloaded once per week.

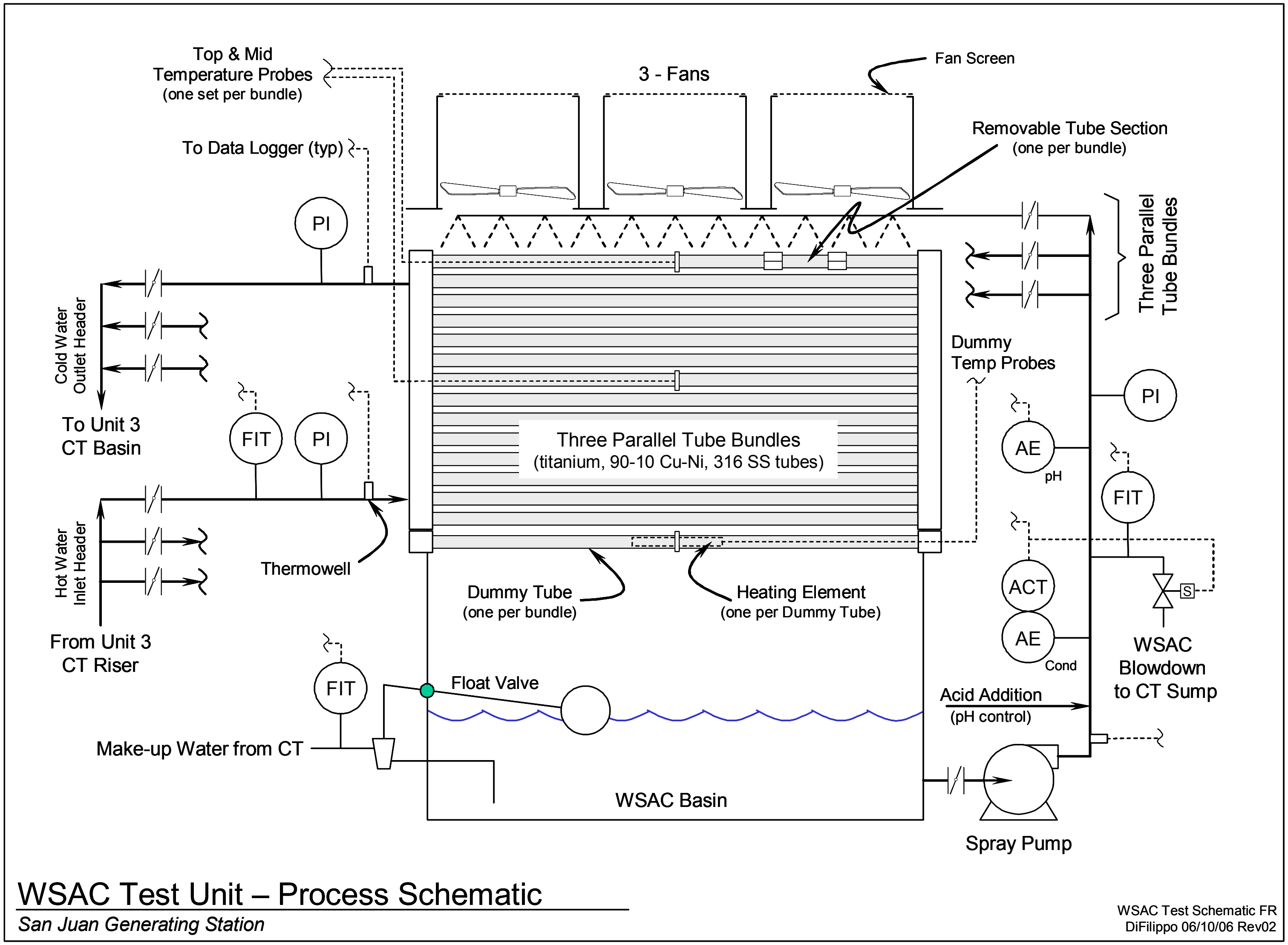

Figure 10-9
Process Flow Diagram of the WSAC Pilot Unit

10.3.2 Analytics

The following volume and chemistry control measures were implemented:

- Make-up was controlled by a float valve in the WSAC basin (on/off).
- A conductivity controller actuated the blowdown valve (on/off).
- pH was monitored (but not controlled) – 93 percent sulfuric acid was added at a constant rate.
- Chemical analyses were conducted on a daily basis (5 days per week) by onsite SJGS laboratory personnel: calcium, chloride (occasional), silica, pH, turbidity, Cl_2 residual and conductivity.

10.3.3 Operating Plan

The goal of the testing was to determine the degree to which blowdown from Unit 3 Cooling Tower could be concentrated in the WSAC and still maintain heat transfer. At the outset of testing it was decided to operate at 5,000 µS/cm conductivity (blowdown from Unit 3 to the WSAC had a conductivity of about 2,500 µS/cm). Three methods were routinely used to crosscheck and verify cycles of concentration during WSAC testing:

- The ratio of WSAC conductivity to Unit 3 Cooling Tower conductivity (operated on average at 7.0 cycles – approximately 70 percent of calcium sulfate saturation)
- Mass balance around the WSAC using totalized volume from make-up and blowdown flow
- The temperature difference across the bundles of the WSAC was used to calculate evaporation rate

Cycles of concentration were also calculated using the ratios of certain chemical constituents in the WSAC and Unit 3 blowdown – namely silica, calcium and chloride concentration.

As testing progressed, the conductivity control setting was to be increased on a stepwise basis. The intention was to determine the cycles of concentration (and chemistry) at which heat transfer would start to deteriorate. This threshold was never reached. As will be discussed later, conductivity control quickly became problematic; however, useful data was obtained.

10.4 Test Data

This section of the report covers the flow, temperature and chemistry data collected during WSAC testing. Refer to Figure 10-8 for the locations of flow meters, temperature probes and analysis elements. WSAC cycles of concentration are calculated for each type of data – flow (mass balance), flow/temperature (heat balance) and chemistry. The results of a corrosion analysis are presented at the end of this section.

10.4.1 Flow Measurement

Flow measurement was critical in determining WSAC performance. In particular, it was used to perform a mass balance around the pilot unit, estimate evaporation rate, calculate cycles of concentration and determine overall pilot heat transfer.

10.4.1.1 Tube Bundle Flow

Flow to the WSAC tube bundles is shown in Figure 10-10. Note that hot-water flow (blowdown from the Unit 3 Cooling Tower) to individual bundles was fairly constant during testing. The flow meter monitoring the east exchanger (316 SS tube bundle) failed after five weeks of operation. Flow to each bundle was assumed to be same (within 3 to 5 gpm of each other). Gaps in data represent pilot downtime (discussed in more detail later). Flow to the bundles dropped slightly over the test period (2 to 3 gpm). There was no apparent reason for this.

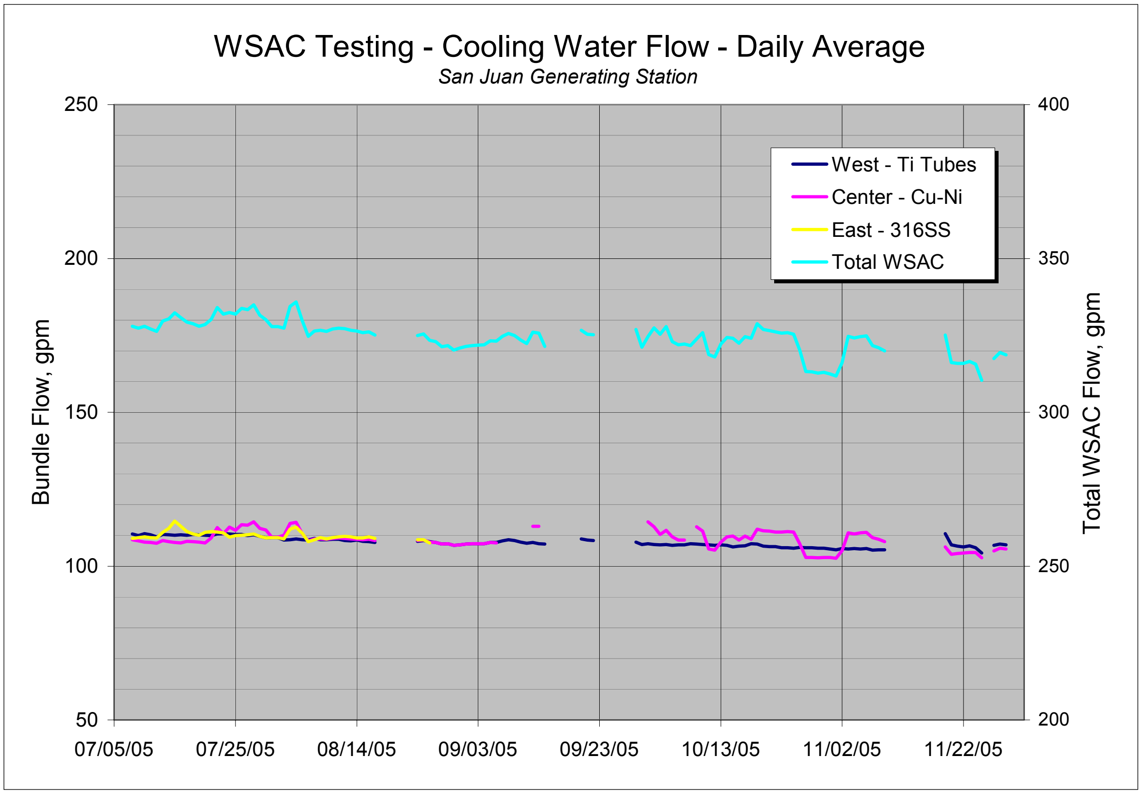

Figure 10-10
Tube Bundle Coolant Flow Rates during the WSAC Test Period

The east bundle (316 SS tubes) flow meter failed after one month of operation. The east bundle flow was estimated by averaging the flows to the other bundles. At one time the center bundle flow meter stopped working, but started again.

10.4.1.2 Flow-Based Cycles of Concentration

Refer to Figure 10-11 for make-up, blowdown and flow-based calculated cycles of concentration. The flow rates are based on 24-hour daily averages. Make-up flow was controlled by a float valve in the WSAC sump, i.e. when the sump level reached a low point, the make-up valve would open to admit feedwater (Unit 3 blowdown). WSAC blowdown was controlled by conductivity. When conductivity reached the control set point, a solenoid valve would open to release circulating spray water (from the discharge side of the Spray Pump).

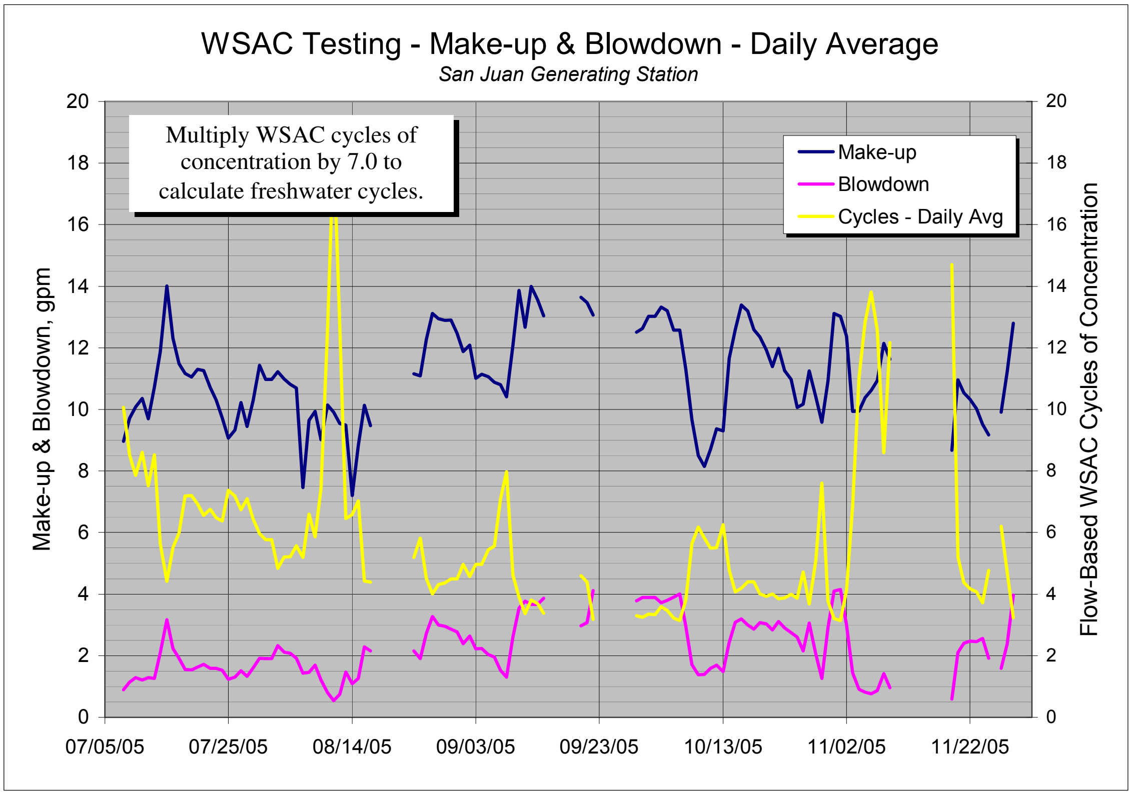

Figure 10-11
Make-up Water and Blowdown Flowrates and Cycles of Concentration during the WSAC Test Period

WSAC cycles of concentration in Figure 10-11 were calculated using make-up and blowdown flow rates as follows:

$$Cycles\ of\ Concentration = \frac{Make\cdot up\ Flow}{Blowdown\ Flow}$$

Cycles of concentration are highly sensitive to blowdown rate, and as can be seen in Figure 10-10, it fluctuated significantly at times – from 3.2 to 18.6 (4.4 cycles average). On August 21, 2005, daily average blowdown dropped to ~0.5 gpm while make-up remained at 9 to 10 gpm. As a result, cycles of concentration soared to 18.6.[8] Recall that WSAC cycles of concentration

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[8] 18.6 cycles of concentration was not supported by WSAC conductivity, which remained fairly constant.

are equivalent to seven times that of the fresh water being fed to the Unit 3 Cooling Tower. Duration-of-project WSAC cycles of concentration averaged 30.8 (on a freshwater basis).

Part of the observed erratic behavior was the inability of controller to maintain conductivity (which actuated the blowdown solenoid valve). Suspended matter "blinded" the conductivity probe at times (discussed later in more detail).

10.4.1.3 Heat-Based Cycles of Concentration

Figure 10-12 shows the evaporation rate of the WSAC unit using two calculation approaches:

- Flow based – calculated taking the difference of the make-up and blowdown rates
- Heat basis – calculated using bundles flow rate and bundle inlet/outlet temperature difference

Figure 10-12 also shows the ratio of flow-based to heat-based evaporation rates. Despite the somewhat erratic flow-based evaporation rate, there is generally good agreement between the two methods of measurement, i.e. the ratio averaged 1.04.

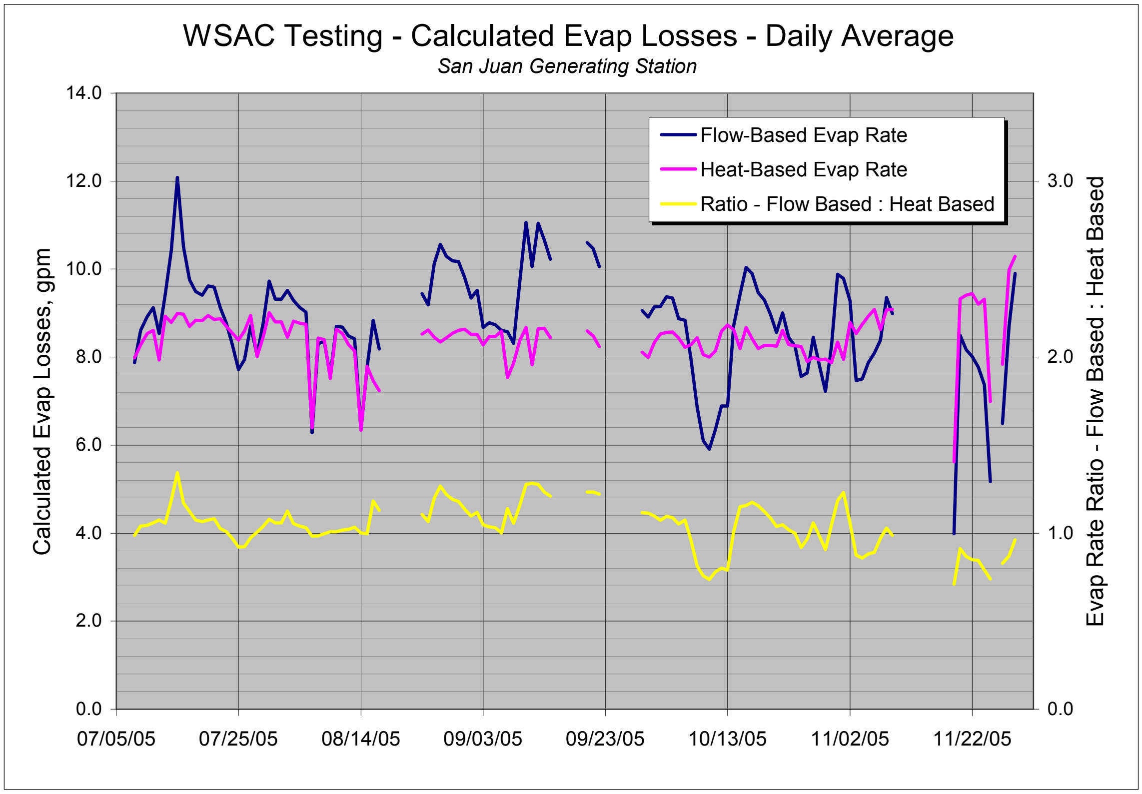

Figure 10-12
Calculated Evaporative Looses and Evaporation Rate Ratios during the WSAC Test Period

The WSAC releases heat in the same manner as a mechanical draft cooling tower, i.e. as circulating water (deluge/spray water) evaporates, latent heat is absorbed. Heat-based evaporation was calculated based on the heat flux of the WSAC tube bundles as follows:

$$Heat \cdot Based\ Evaporation\ Rate = \frac{Flow\,\Delta T\, C_{P,H_2O}}{H_{Evap,H_2O}}$$

where C_{P,H_2O} is the heat capacity of water (1 BTU/$\#_{H_2O}$/F) and H_{Evap,H_2O} is the latent heat of evaporation of water (1,000 BTU/$\#_{H_2O}$). Heat-based evaporation was less erratic because it is based on the relatively steady flow rate of water to be cooled and the fairly constant temperature drop across the tube bundles (discussed next).

10.4.2 Temperature Measurement

The inlet and outlet temperature differences across the WSAC bundles are found in Figure 10-13. Refer again to Figure 10-8 for the location of temperature probes and thermocouples. There are a number of data gaps which were caused by necessary repairs or upgrades to the WSAC pilot, a power plant outage and a data logger failure.

As early testing proceeded, significant fan vibrations developed. One or more fans were turned off to reset their motor starters (sometimes to reconnect the wiring).[9] As can be seen in Figure 10-13, when one or more fans were off, the temperature difference (delta T) clearly dropped. In mid August 2006, the pilot was shut down to clean the fans, which had significant mineral deposits on the blades. It was noted at that time that the demister section was not installed in the WSAC prior to shipping from the manufacturer. In mid September, the demister was installed and the fan vibrations ceased.

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[9] The motor starters were mounted on the side wall of the WSAC just below the fans. Refer to Figure 10-4.

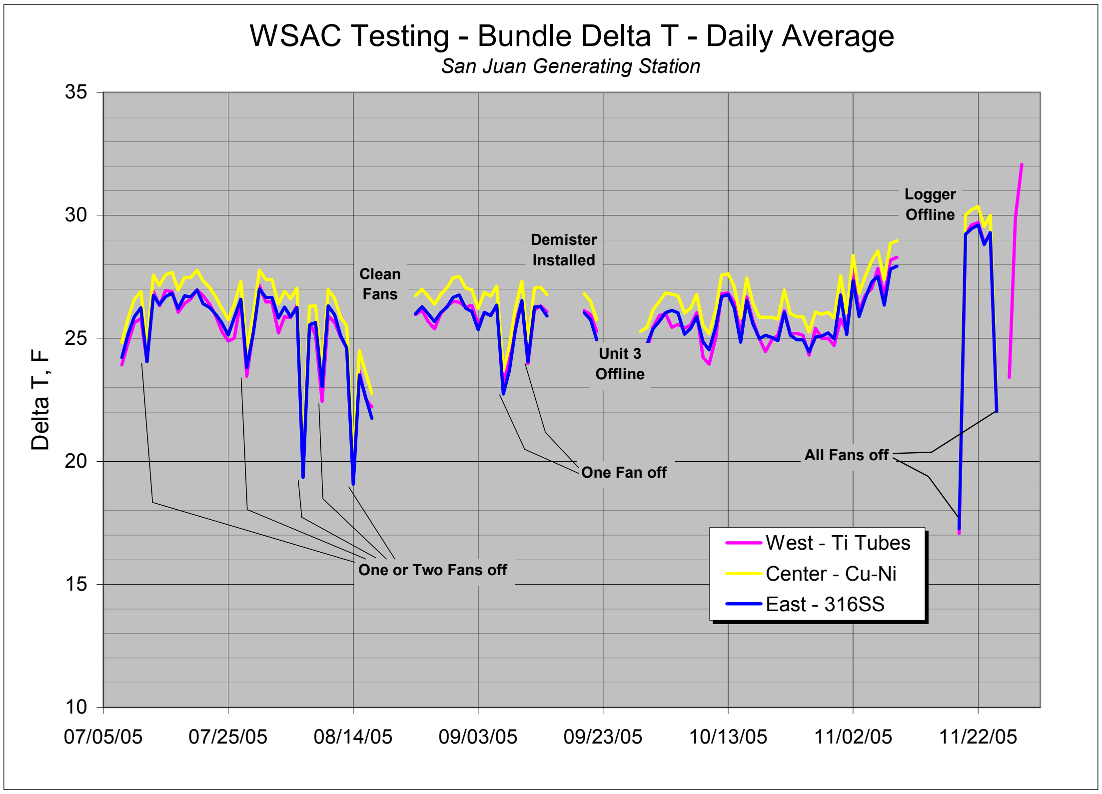

Figure 10-13
Temperature Differences between Coolant and Circulating Water the during WSAC Test Period

10.4.2.1 Tube Bundle Delta T

The fine detail of tube-bundle inlet-outlet temperature differences in Figure 10-13 shows that the 90-10 Cu-Ni tubes had the best relative heat transfer characteristics – a difference of about 1 ºF. The performance of the Ti and 316SS tubes tracked closely. The sudden rise in delta T in November 2006 was the result of the onset of fall weather and much colder ambient air temperatures. Interferences and data gaps aside, tube-bundle delta T remained fairly constant from mid-July through October 2005.

10.4.2.2 Tube Bundle Heat Transfer

Refer to Figure 10-14 for tube bundle heat transfer (expressed as 10^6 BTU/hour). As in Figure 10-13, there are data gaps and spikes in heat transfer where one or more fans were offline and there were test interruptions. This analysis shows that the bundle with Cu-Ni tubes had the best heat transfer. The bundle with stainless steel tubes had slightly better heat transfer than the bundle with titanium tubes. Tube-bundle heat transfer remained reasonably constant from mid-July through October 2006.

10.4.2.3 Tube Skin Temperature

Skin temperature was measured at two points on each tube bundle – center top and center mid-bundle. Thermocouples were attached to the tubes using standard stainless-steel hose-band clamps. Refer to Figure 10-15 for a four-day interval comparing skin temperatures of the Ti, 90-10 Cu-Ni and 316 SS tube bundles. Diurnal temperature effects are evident (cyclic variations) – data were recorded at ten-minute intervals. Bundle delta T was 29°F to 30°F, and as stated previously, the Cu-Ni tubes provided slightly better heat transfer, i.e. lower outlet temperatures.

The skin temperature was the coldest at the top of the bundle, where cooled spray water first meets the tube bundles. Bundle outlet water temperature (accounting for diurnal variations) was consistently close for all the bundles (within 1 ºF).

There were significant differences in mid-bundle skin temperatures. The Ti tubes were significantly warmer at mid-bundle by 8°F to 9°F compared to Cu-Ni and 316 SS. The probe measuring mid-bundle Ti skin temperature could have been sensing incorrectly (probe defect, insufficient probe/tube contact, etc.). Also, there could have been a flaw in the flow pattern of deluge water with reduced flow in the vicinity of the temperature probe. Again, outlet water temperatures were very close.

Some thermocouples started to fail shortly after start-up – corrosion was observed at the thermocouple-tube interface. It was not clear whether the thermocouples or the hose bands were corroding. Top and dummy-tube thermocouples (located below the center bottom of each tube bundle) were replaced with self-adhering thermocouples. Mid-bundle thermocouples were not changed because they could not be reached. The failures did not recur.

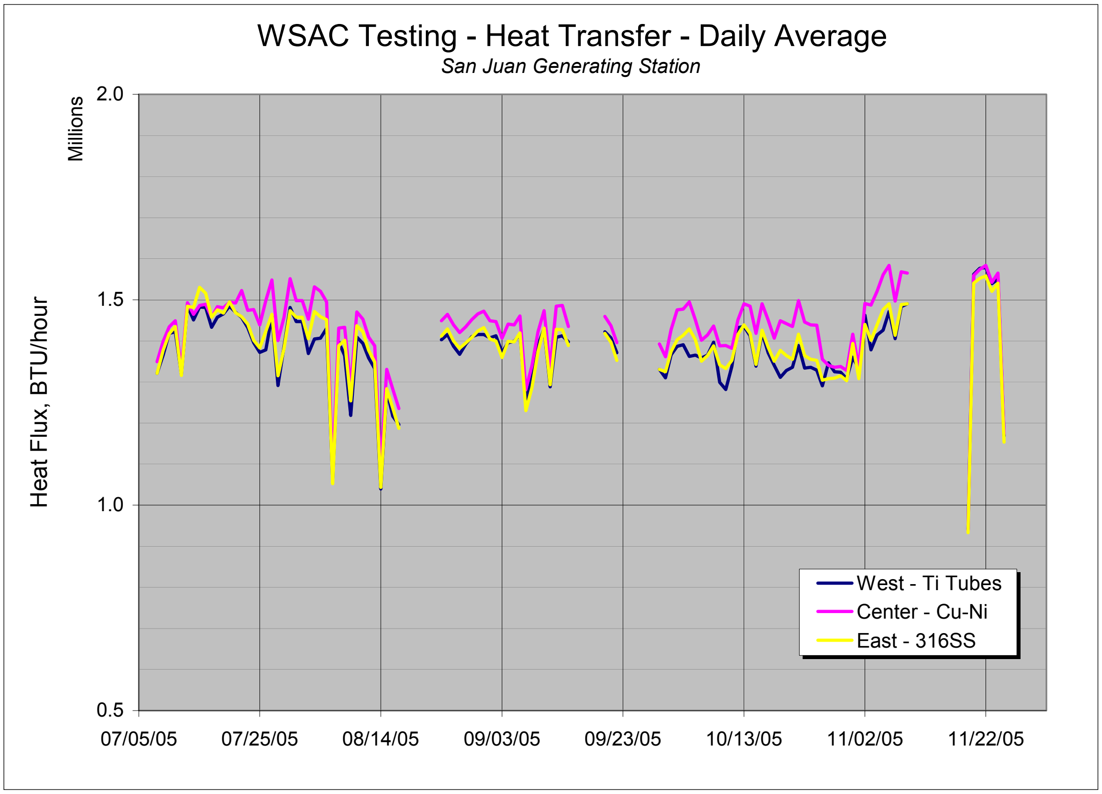

Figure 10-14
Heat Flux to Tube Bundles during WSAC Test Period

10.4.2.4 Heat Transfer Coefficient

Heating elements were installed in the dummy tubes located below the center of each tube bundle. The tubes were designed such that a heating element could be inserted into either end of the tube. The 14-inch heating element would impose a constant heat load on the tube section where a thermocouple would monitor skin temperature. This arrangement would allow the measurement of the heat transfer coefficient of the tubes under typical water flow conditions across the exterior of the tube. Changes in skin temperature would identify slight changes (degradation) in heat transfer.

The heating elements, which arrived after start-up, were not installed until the testing was well underway. Problems were encountered in centering the elements in line with the thermocouples. Only one element worked properly (titanium dummy tube). Instead of a repeating diurnal pattern, skin temperature rose and fell dramatically. As discussed previously, this could have been due to probe/tube contact or uneven flow patterns. across the surface of the tube. Also, the heat output of the element was too high (elements were not sized correctly), therefore, the skin temperature, which was 40°F to 50°F higher than the surrounding tubes, was not representative of the heat load of working the tubes in the bundles. Given these circumstances, the data were deemed unusable.

Figure 15

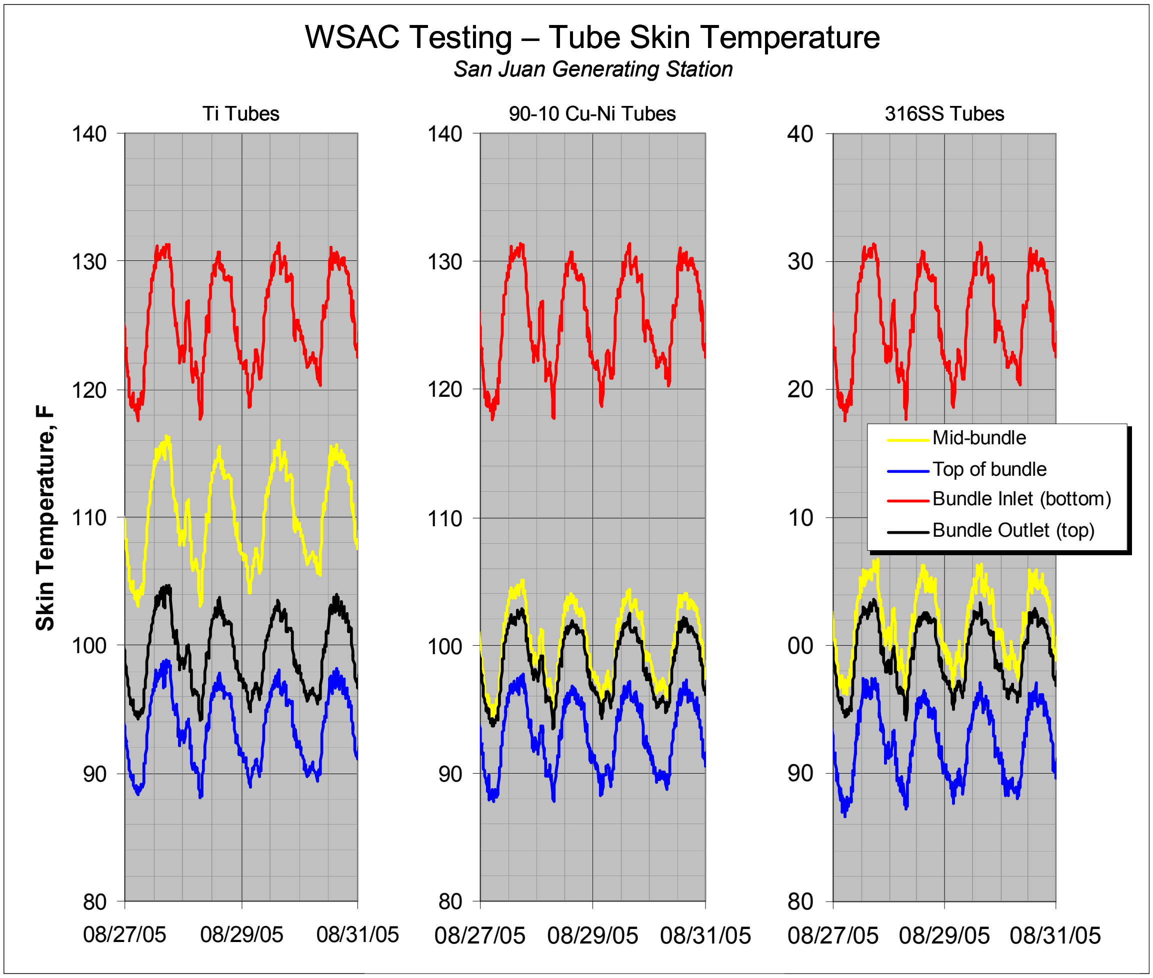

Figure 10-15
Tube Bundle Skin Temperatures during the WSAC Test Period

10.4.3 Chemistry

Chemistry was monitored throughout the testing. The following control and sampling measures were implemented:

- pH and conductivity were monitored on ten-minute intervals.
- A conductivity controller actuated the blowdown valve (on/off).
- pH was monitored (but not controlled) – 93 percent sulfuric acid was added at a constant rate to achieve a somewhat constant pH of 7.6 (roughly that of the Unit 3 Cooling Tower).
- Chemical analyses were conducted on a daily basis (5 days per week) by onsite SJGS laboratory personnel: calcium, chloride (occasional), silica, pH, turbidity, Cl_2 residual and conductivity

- A chlorine residual (0.05 to 0.2 mg/l_{Cl2}) was maintained to control biological growth – none was observed.[10]

10.4.3.1 pH & Conductivity

Refer to Figure 10-16 for WSAC pH and conductivity observations. pH was easily controlled in a relatively narrow range throughout the testing.[11] There were two episodes of high pH as a result of losing acid feed (the injection quill was plugged). Conductivity for the WSAC and Unit 3 Cooling Tower are also shown in Figure 10-16. WSAC conductivity control was problematic while Unit 3 conductivity was fairly constant. WSAC conductivity remained at 5,000 µS/cm despite frequent adjustments to the controller set point (to increase conductivity) even after recalibrating the instrument. The conductivity probe was being blinded by high levels of suspended matter in the WSAC deluge water. In hindsight, the sample stream should have been filtered (with a strainer) to keep the conductivity cell free of debris and operable at all times. Toward the end of the testing conductivity was increased to 6,000 µS/cm after repeated attempts to clean and calibrate the instrument.

There were a number of small process leaks, especially around the deluge/spray pump. These leaks, which were very small and immeasurable, could have contributed to the loss of conductivity control in that they are comparable to a steady blowdown stream.

There was secondary effect that impeded conductivity. As the cycles of concentration of the Unit 3 cooling water were raised in the WSAC, constituents like calcium and sulfate were dropping out of solution. Recall that in Unit 3, calcium sulfate is the limiting parameter at 70 percent of saturation. Therefore, as the cycles increased in the WSAC, soluble calcium and sulfate ions were being removed from solution, i.e. their removal from solution as mineral scale reduced the conductivity of the WSAC circulating water.

10.4.3.2 Constituent-Based Cycles of Concentration

In Figure 10-17, constituent-based cycles of concentration for calcium, silica and chloride are presented with flow-based and heat-based cycles. There is relatively good agreement among flow-based and heat-based cycles of concentration. For the most part, these parameters paralleled each other – from 3.5 to over 10 cycles of concentration (24 to 70 cycles based on Unit 3 freshwater make-up). The limited number of data points for chloride-based cycles agreed with flow-based and heat-based cycles (except for one point). As discussed previously, conductivity-based cycles of concentration were practically constant (varying between 2 to 2.5 cycles) and were always less than flow-based and heat based cycles. This was likely the result of problems with the conductivity probe/controller and ionic constituents (calcium and sulfate) dropping out of solution and forming mineral scale. Calcium-based and silica-based cycles of

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[10] A chlorine residual was maintained by adding a chlorine puck (tablet of calcium hypochlorite) every day or two. This type of addition is commonly used to control chlorine levels in swimming pools. This control method added very little chloride and calcium to the background levels in the WSAC bulk fluid volume.

[11] Sulfuric acid was fed at a fixed rate for pH control. The acid pump was powered to start and stop with the Deluge/Spray Pump.

concentration paralleled flow-based and heat-based cycles, but were almost always less as a result of dropping out of solution.

Figure 16

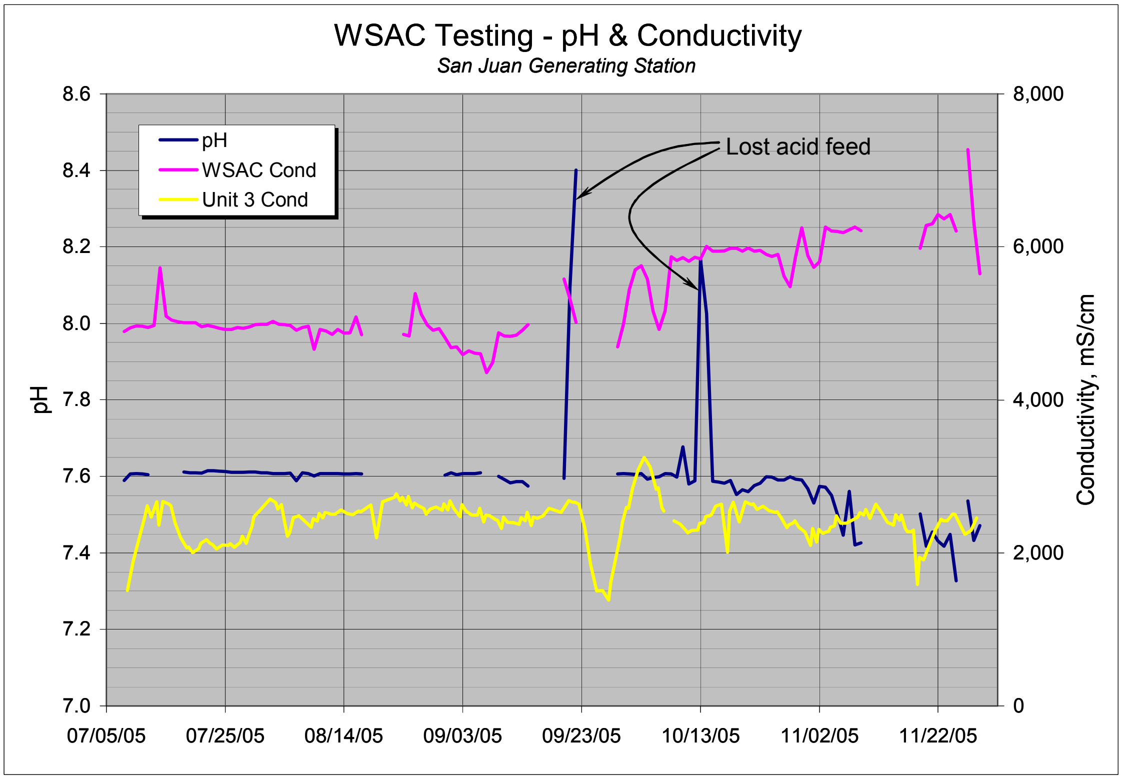

Figure 10-16
Coolant pH and Conductivity during the WSAC Test Period

Figure 17

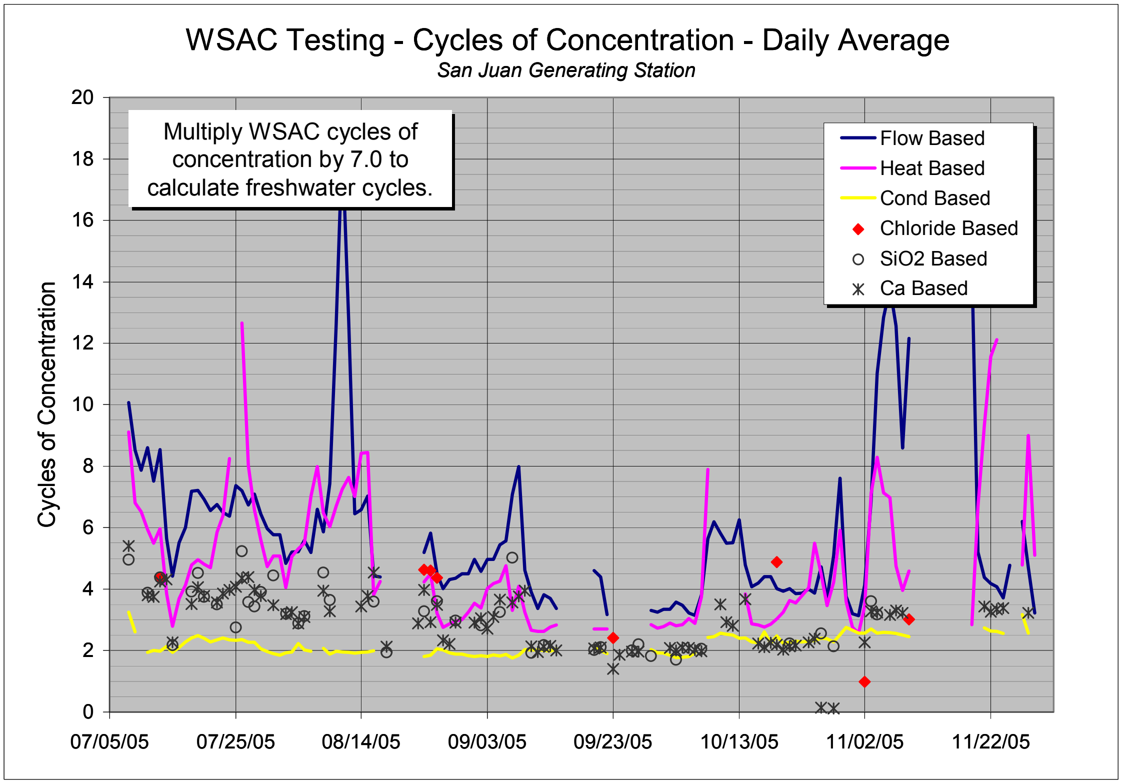

Figure 10-17
Cycles of Concentration of Cooling Water based on Heat, Conductivity, Chloride, Silica, and Calcium during the WSAC Test Period.

10.4.3.3 Suspended Matter

During the testing, solids build-up on tube bundle side walls was fairly significant. High levels of suspended matter (TSS, total suspended solids) were carried in the bulk fluid and continuously washed over tubes surfaces. Turbidity, an indirect measure of TSS, was monitored daily (one grab sample per day).[12]

Figure 10-18 shows how the turbidity of the WSAC deluge/spray water and Unit 3 Cooling Tower circulating water varied during day-to-day testing. WSAC Turbidity exceeded 1,000 NTU at the outset of testing, averaged 350± NTU in the early stages and attenuated to 200± NTU in the last half of the testing. Based on extrapolated data, the turbidity of Unit 3 Cooling tower blowdown at 200 NTU is equivalent to about 400 mg/l of TSS. Refer to the inset in Figure 10-18 for a relationship between TSS and turbidity in the Unit 3 cooling system.[13]

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[12] Turbidity is expressed as Nepthlometric Turbidity Units (NTU) which are a measure of light interference.

[13] This TSS-NTU relationship was developed from 44 data points taken from Unit 3 cooling Tower chemistry records.

Extrapolated values for TSS were not attempted for turbidity levels above 200 NTU because of the limited range of the data.

Using the TSS-NTU relationship in Figure 10-18, TSS levels were very high during the first few months of testing and settled to around 400± mg/l (200± NTU). These levels of TSS are considered very high for a cooling system.[14] What can be inferred from the data is that the WSAC was "moving" significant amounts of suspended matter in the bulk deluge/spray water. At the same time, solids were being deposited on and in quiescent areas, i.e. tube bundle side walls and the basin. Refer to Figure 10-19.[15] Mineral scale was deposited on tube-bundle side walls to thicknesses up to 0.25 inches. At places in the WSAC basin, sludge depths were in excess of 3 to 4 inches. Even with all the scale on non-wetted surfaces, the stainless steel tubes in Figure 10-19 remained clean – visually and to the touch. Refer to the insert in Figure 10-19 to see the condition of the spray header, bundle side walls and tubes before testing.

10.4.3.4 Residual Scale and Corrosion Inhibitors

The WSAC pilot unit may have benefited from residual concentrations of scale and corrosion inhibitors in Unit 3 Cooling Tower blowdown. A phosphonate-type scale inhibitor is used to control mineral scale and an azole-type corrosion inhibitor is used to protect copper-bearing metallurgy in all of the cooling towers at San Juan Generating Station The scale inhibitor may have benefited the pilot, but it was not possible to determine its effects (if any). Given the degree of mineral scale formed during testing, the scale inhibitor was likely overwhelmed and had no measurable beneficial effect. As will be discussed next, the metallurgical limits of the tubes were not challenged during WSAC testing, i.e. none of the tube specimens showed any signs of external surface corrosion. It was assumed that the salinity[16] of the WSAC water was not high enough to initiate corrosion.

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[14] Cooling tower operators usually try to keep TSS below 100 mg/l with a nominal operating average of 50 mg/l.

[15] This photograph of the WSAC East Cell (316 SS tubes) was taken in November 2005 – note the ice formation along the side walls.

[16] For the 316 SS tube bundle, the concern was high chloride content. High levels of chloride can cause stress-cracking corrosion.

Figure 18

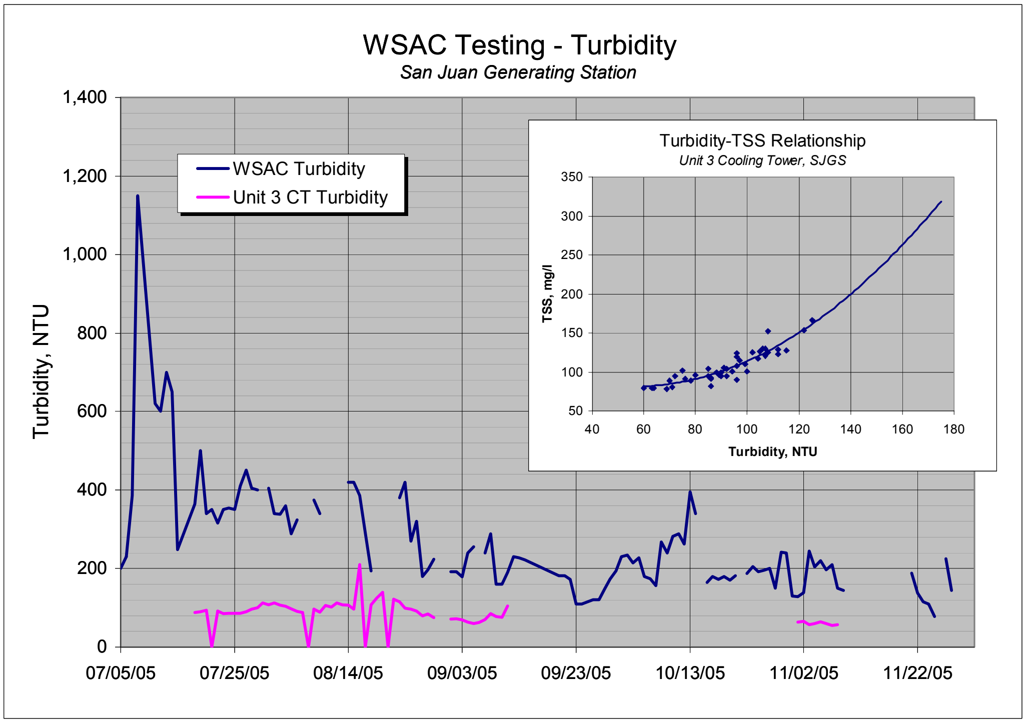

Figure 10-18
WSAC Turbidity Data

Figure 19

Mineral Scale Deposition - 316 SS Tube Bundle

WSAC Pilot Unit

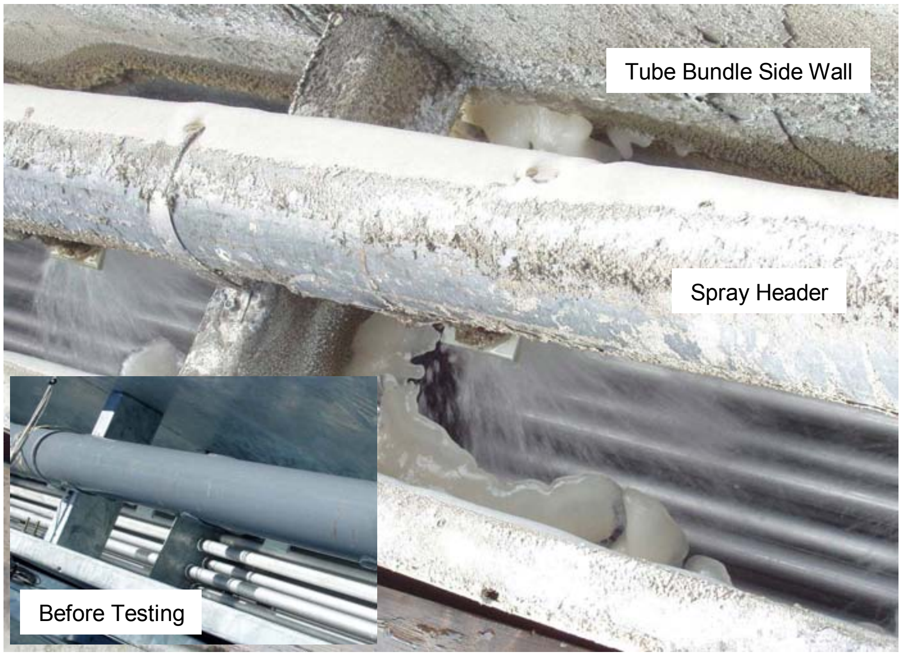

Figure 10-19
Mineral Scale Deposition - 316 SS Tube Bundle

10.5 Corrosion Analysis

The WSAC pilot unit has removable 12-inch tube sections – one per bundle located in the top row. Each tube section[17] was removed and sent to an outside laboratory[18] for a complete metallurgical assessment (the report is found in Appendix A). During the assessment, each tube specimen was cut in half lengthwise to inspect its interior surfaces.

The laboratory concluded that the external surfaces of the tubes were smooth and had no visible signs of corrosion. It was estimated that a thin layer of deposition – 0.001 to 0.003 inches – had formed on the external surfaces. This deposition layer could have occurred at shut down. The unit was to be flushed immediately after the fans and deluge/spray pump were turned off to remove residual suspended matter and mineral scale found in the bulk spray water. The flushing procedure could not be followed because of other maintenance demands in the plant at the time of shut down.

The interior surfaces of titanium and 316 SS tube sections were smooth with no visible signs of corrosion (there were occasional deposits). Conversely, the 90-10 Cu-Ni tube section showed signs of pitting corrosion on the interior surface. Recall that the interior surfaces of the WSAC tubes were only exposed to circulating water from Unit 3 Cooling Tower. Interestingly, the Cu-Ni tubes showed the best heat transfer characteristics during testing.

10.6 Summary of Findings

The goal of the WSAC testing at San Juan Generating Station was to determine the extent to which degraded water could be used for cooling without impeding heat transfer. Although the testing never achieved this goal, the WSAC operated at very high cycles of concentration with significant levels of suspended matter without degrading heat transfer.

10.6.1 Observations, Findings and Conclusions

The following observations, findings and conclusions summarize the test results:

- Throughout the testing, the tube exterior surfaces appeared visually clean. The surfaces were always smooth to the touch, i.e. no apparent mineral deposits or biological film.
- The maximum cycles of concentration that would impede heat transfer were not determined. The inability to progressively increase conductivity over time in a stepwise manner (i.e. control conductivity) prevented the testing from achieving higher sustained cycles of concentration.
- There were a number of small process leaks, especially around the Deluge/Spray Pump. These could have contributed to the inability to control conductivity in that they were comparable to a steady (albeit small) blowdown stream.

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[17] The tube sections were held in place with water-tight metal collars.

[18] GE Infrastructure Water & Process Technologies performed the analysis.

- Cycles of concentration varied (somewhat erratically at times) from 3.5 to over 10 during the test period – the average for the test period was 4.4 cycles. This was equivalent to 24 to 70 cycles of concentration (30 cycles average) based on Unit 3 Cooling Tower freshwater make-up. This was far in excess of the control level for the cooling system – 10 cycles of concentration (based on calcium sulfate saturation).
- Cycles of concentration were determined by a number of methods – make-up and blowdown flow rates, evaporative loss via tube-bundle temperature difference, and chloride concentration. All methods showed general agreement.
- Heat transfer was maintained throughout the testing, i.e. the temperature difference across the bundles was steady. Heat transfer was only a problem when one or more of the three fans were taken out of service.
- The bundle with Cu-Ni tubes had slightly better heat transfer than the titanium and 316 SS bundles (even with signs of corrosion in the interior Cu-Ni surfaces).
- Calcium and silica cycles of concentration paralleled flow-based and heat-based cycles, *but they were always less*. This was attributed to calcium and silica dropping out of solution to form mineral salts at elevated cycles.
- Conductivity and pH (by acid addition) were the only variables controlled during the testing, i.e. no other means of scale control were employed.
- There was no measurable benefit to the WSAC pilot from residual concentrations of scale and corrosion inhibitors in Unit 3 Cooling Tower blowdown.
- The exterior surfaces of the tubes fared well with respect to visible scale and corrosion. Only the Cu-Ni tubes showed signs of corrosion on the internal surfaces (non-WSAC side).
- Mineral scale and suspended matter (measured as turbidity) were very evident throughout the testing. TSS levels greatly exceeded cooling system standards. It was clear that solids control is important if WSAC technology is to be used in this manner.

Given the operating conditions of high cycles of concentration, formation of extensive mineral scale and very high levels of suspended matter in the bulk cooling water, the WSAC performed well by maintaining heat transfer throughout the test period.

10.6.2 Unresolved Issues

The following issues were not resolved during WSAC testing in 2005 at San Juan Generating Station.

- The WSAC did not reach the heat transfer threshold where mineral scale deposition on external tube surfaces would interfere with heat transfer.
- The WSAC did not challenge the 316 SS, 90-10 Cu-Ni or Ti tube metallurgy (this was not anticipated in this round of testing).
- Suspended matter interfered with conductivity control (i.e. the ability to control cycles of concentration) and created a huge solids burden in non-heat transfer areas of the pilot unit.

If further testing is pursued for the WSAC pilot, higher cycles of concentration, potential tube corrosion and solids control should be evaluated by:

- Protecting instruments with upstream strainers to enable conductivity control and higher cycles of concentration.
- Finding a degraded source water with higher salt content than Unit 3 Cooling Tower blowdown at San Juan Generating Station to approach the corrosion limit of tube bundle metals.
- Installing a filter on a slip stream off the deluge/spray riser to continuously remove mineral scale (in the bulk fluid) as it is formed. Mounting freshwater spray headers with low pressure nozzles above the bundle side walls and below the demister to intermittently flush these surfaces and keep them free of mineral scale.

Specific areas of concern are discussed next.

10.6.2.1 Instrumentation

Critical instrumentation needs to be shielded from the high solids environment of degraded water at elevated cycles of concentration. This is especially applicable to analysis probes – in particular, conductivity and pH. Sample streams that feed these instruments should have upstream strainers to remove suspended material. Suspended material blinds the analysis element making it difficult to obtain usable information.

Flow meters should also be selected carefully for high TSS and high salinity duty. Paddlewheel-type flow meters should be avoided because of suspended material in the bulk fluid. Vortex meters should be erosion resistant (from suspended matter). For any type of flow meter, materials of construction are critical, i.e. the material (metal or plastic) must be compatible with WSAC chemistry and possibly abrasive solids.

10.6.2.2 Solids Control

Solids generation and accumulation in the WSAC were significant. Blowdown (which was intermittent) did not remove enough solids to control deposition on bundle side walls and accumulation in the basin. Additionally, it was shown that, when the WSAC was operated without demisters, there was significant carryover to the fans. Therefore, deposition on the underside of the demisters will likely be a problem as well.

Consideration must be given to removing solids from the WSAC as they are generated over and above the capability of the blowdown stream. One method could be to filter a slip stream of deluge/spray water. Side-stream filtration has been used for years on a number of full-scale industrial cooling towers to control solids build-up. Filtrate from these systems is returned to the cooling tower basin. For the WSAC, the flow capacity of the Deluge/Spray Pump would have to be increased by the amount of flow for the side-stream filter. Side-stream filters are sized anywhere from 5 to 10 percent of the circulating flow of a cooling system.

Side-stream filtration should control solids accumulation, especially in the basin. Filtration may not completely control deposition on bundle side walls and on the inlet side of the demisters, since some of the deposition in these areas is coming from droplet carryover (to the demister) or splashing at and around the spray header. These areas could be wetted intermittently with a small volume of freshwater to wash away deposition that has not dried in place.

10.6.2.3 Metallurgy

The metallurgical limits were not challenged during WSAC testing, i.e. none of the tube specimens showed signs of external surface corrosion. The 316 SS was expected to be the least capable metal. Chloride stress-cracking corrosion was expected, but did not occur.

Water to be cooled must be compatible with the WSAC tube-bundle metallurgies (316 SS, 90-10 Cu-Ni and Ti). Pitting corrosion, which was not expected, was found on the interior tube surfaces of the Cu-Ni test specimen. Unit 3 Cooling Tower circulating water was apparently not compatible with Cu-Ni.

10.6.3 Need for Further Study

As a result of the testing at San Juan Generating Station, WSAC showed great potential to cool degraded water significantly beyond normal cooling tower operating limits. Further testing is required to demonstrate the operating limits of the technology, i.e. the point at which heat transfer starts to degrade either from mineral scale or corrosion. Therefore, additional testing would be used to validate this application of WSAC technology, i.e. determine its operating limits and identify operating and reliability issues.

Upgrades will be necessary to conduct further WSAC testing. Improvements include:

- Strainers to protect conductivity and pH analysis elements for better controls of cycles of concentration
- Side-stream filter to remove suspended matter from the bulk cooling fluid
- Spray headers to intermittently wash tube-bundle side walls and the inlet side to the demister

In addition to these improvements, flow meters and analysis elements will be replaced to improve data gathering and data quality. Pipe connections and hatchways will be tightened to minimize process leaks. Lastly, the unit will be thoroughly cleaned using high pressure water to remove mineral scale and deposited suspended matter.

If it can be demonstrated with further testing that WSAC can reliably use degrade water as a cooling source without performance degradation, an economic analysis would be conducted to determine efficacy of using WSAC technology in lieu of mechanical draft cooling towers.[19] The economic analysis was to be part of the original phase of testing, but given the problems

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[19] Standard cooling technology, in addition to special materials of construction for degraded water, would require pre-treatment to protect the cooling system from mineral scale. Also, post-treatment is sometimes required to reduce the volume of blowdown for waste disposal at inland plants that require zero-liquid discharge.

encountered with scale generation and instrument blinding, it was decided to postpone the analysis until these critical issues could be proven manageable.

The economic analysis would evaluate several configurations, e.g. WSAC could be utilized as an auxiliary cooler and use power plant wastewater as a make-up source (e.g. at inland plants). In this mode, it would be a pre-concentrating device to a zero-liquid treatment system and could provide significant economic benefit by reducing capital and operating costs of wastewater treatment. WSAC could also be incorporated into the main cooling circuit and share the load with the main cooling tower(s) while using their blowdown as make-up.

11

ADAPTATION OF A DETERMINISTIC WATERSHED MODEL FOR CLIMATE-HYDROLOGY

11.1 Introduction

The development and application of useful decision support systems (DSS) are critical for water resources decision making and management, particularly in assessing the impacts of drought and climate change. Large scale modeling has typically been used to address the impact of climate change. General circulation models or global climate models (GCMs) numerically simulate changes in climate on a coarse spatial scale (on the order of 300 x 300 km); however, they generally do not provide a good estimation of hydrological response to climate change on a watershed scale. Past studies have developed linkages between GCMs and other models using stochastic or regression based techniques (Strzepek and Yates 1997, Dibike and Coulibaly 2004, Zhu et al. 2005). Limitations of these studies include a recognized simplicity in the approach taken with respect to spatial scale and detail of human-influenced water management.

Concurrently, a set of models have been developed to address watershed management and calculation of total maximum daily loads (TMDLs) (Chen et al. 2001, Bicknell et al. 1993, Neitsch et al. 2001). These tools, not historically developed for climate studies, use a deterministic modeling approach. The models characterize a watershed as a network of land catchments, apply historical precipitation and temperature data, and simulate resulting streamflow. Significant detail regarding managed flows (e.g. diversions, reservoir releases) may be incorporated into these tools at a small spatial scale. Historical records can be selected for drought periods and used to produce a single, deterministic prediction of resulting streamflow. This approach provides a “snap shot” picture of the potential impact of drought or climate change; however conclusions drawn from the simulation may be limited due to uncertainty regarding how well a limited multiyear record of temperature and precipitation data represent a basin’s climate.

This section discusses a unique approach to adapt a mature, map-based watershed simulation model to handle the complexity of water diversion activities and reservoir operations under varying climatic conditions using a historic sampling approach. Motivation for this work was driven by the recent severe drought in the southwestern United States. An application of the tool to the San Juan Basin (CO, NM) demonstrates an analysis of the impact of projected drought coupled with increasing mean annual temperature associated with projected climate change. While this modeling framework does not attempt to predict the full suite of impacts associated with climate change, it does enable water managers to explore potential impacts of climate change within a user-friendly, PC-based modeling framework.

11.2 Methods

11.2.1 Background

The ZeroNet Water-Energy Initiative was initiated by the U. S. Department of Energy to help address critical issues facing water users in arid watersheds. One of the goals of the ZeroNet Initiative is to develop a decision support system (DSS) to support water management (Rich et al. 2005). The ZeroNet DSS was designed to provide tools that synthesize critical water supply and demand information and assist water owners and managers with planning for shortages. Watershed stakeholders can use the ZeroNet DSS to understand four aspects of water management: 1) the functioning of their river basin; 2) how new technologies and use of degraded water for power plant cooling can affect freshwater availability; 3) how to formulate and evaluate management options for water shortage sharing schemes under drought conditions; and 4) how future growth and development and possible climate change may affect water availability. The ZeroNet DSS has three major components: Watershed Tools, a Quick Scenario Tool (QST), and a Knowledge Base, based on foundations of stakeholder input, watershed data, and geographic information system (GIS) capabilities. This paper emphasizes the development of the Watershed Tools component of the ZeroNet DSS via adaptation of a mature watershed model to simulate impacts of drought and climate change via a historical sampling technique to construct drought scenarios.

11.2.2 Watershed Tools: WARMF

The Watershed Analysis Risk Management Framework (WARMF) was selected as the Watershed Tools component of ZeroNet DSS. WARMF contains a dynamic watershed simulation model that calculates daily surface runoff, ground water flow, non-point source loads, hydrology, and water quality of river segments and stratified reservoirs (Chen et al. 2005, Chen et al. 2004, Herr et al. 2003, Keller 2000, Weintraub et al. 2001). In WARMF, a watershed is divided into a network of land catchments, river segments, and reservoir layers. Land catchments are further divided into land surface and soil layers. These watershed compartments are seamlessly connected for hydrologic and water quality simulations. The land surface is characterized by its land uses and cover, which may include forested areas, agriculture lands, or urbanized cities. Daily precipitation, which includes rain and snow, is deposited on the land catchments. WARMF performs daily water balance calculations to partition precipitation between evapotranspiration, surface runoff and groundwater accretion to river segments. The water entering the river is then routed from one river segment to the next downstream river segment until it reaches the terminus of the watershed. The associated point and nonpoint loads are also routed through the system. Heat budget and mass balance calculations are performed to calculate the temperature and concentrations of various water quality constituents in each soil layer, river segment, and lake layer. WARMF provides a robust framework to address the complex issues proposed by the ZeroNet project.

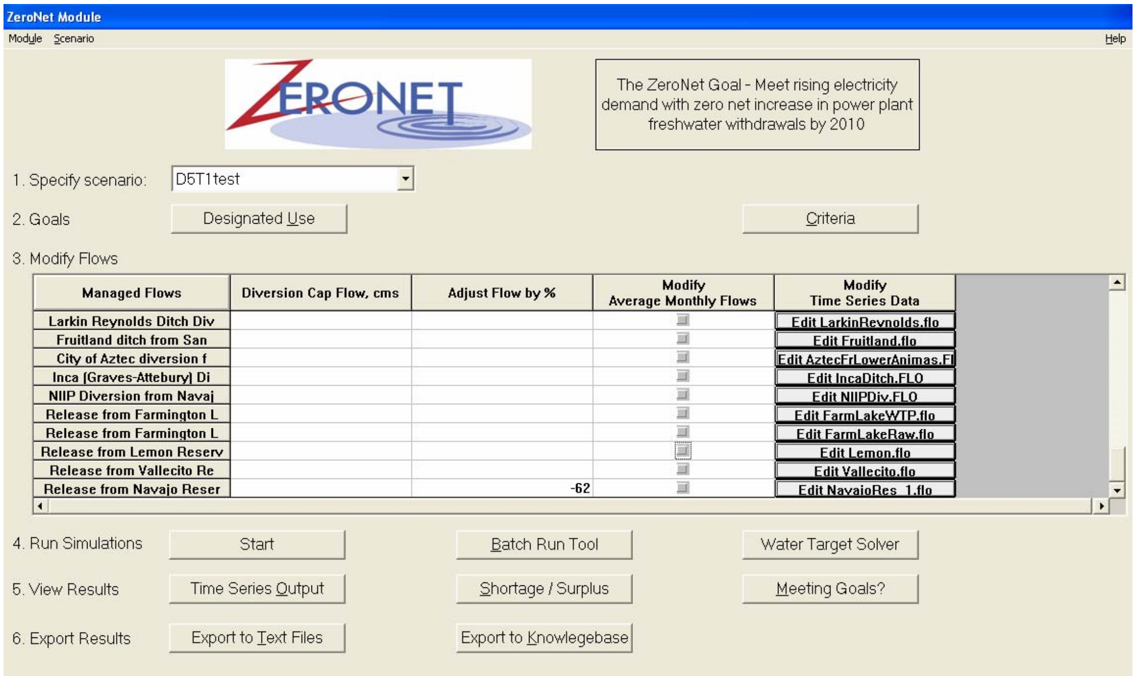
WARMF has been applied to over nineteen watersheds in the United States and internationally and was originally designed to support modeling and planning for total maximum daily loads (TMDLs). Past watershed applications have mainly focused on single, deterministic model runs. Strengths of WARMF include its comprehensive hydrologic model, built-in database, map-based user interface, and strong stakeholder decision making tools. In order to predict the relative impact of climate change and extended drought conditions on water supply as part of the ZeroNet Initiative, we enhanced WARMF with a new approach for scenario construction and execution.

11.2.3 WARMF ZeroNet Module

To analyze the complexity of water diversion activities and reservoir operations under varying climatic conditions, WARMF was enhanced to include climate and water supply scenario generators to examine consequences of changes in climate, and water allocation. This was accomplished by development of a ZeroNet Module and a Batch Scenario Tool.

The main structure of the ZeroNet module is a series of steps, 1 through 6, which guide the user through setting up simulations and viewing model output (Figure 11-1). In these steps, new scenarios can be created, and designated uses (e.g. fish habitat) and criteria (e.g. minimum streamflow or reservoir elevation) can be established. In a spreadsheet tool, adjustments can be made to specified diversions and reservoir releases by setting maximum flow allocation values or percent adjustments to historical data. Scenarios may be run using a traditional deterministic approach which will produce output for a single simulation. Or the Batch Run Tool may be selected to run a set of simulations. Though not a true stochastic model, the batch scenario tool enables a user to construct hypothetical climate scenarios based on historical climate data. WARMF samples a set of historical years of data to create a long term simulation which reflects a user-specified pattern of wet, normal and dry years. After viewing a table which segregates the historical meteorologic data into wet, normal, and dry years, the user selects the number of years for simulation, the number of simulation repetitions, and a pattern of meteorologic conditions by classifying each year of the simulation as either dry, wet or normal. Parameters for degree of temperature increase to reflect climate change, and modification of reservoir releases and diversions can also be set. For each simulation, WARMF randomly selects each year from the appropriate bin of historical wet, dry or normal years. WARMF will create the needed input files for meteorology, reservoir releases, diversions, point sources and upstream boundary conditions. The sampled years are linked together to create a continuous simulation and results are saved in standard output files. Multiple simulations are run to produce a probabilistic distribution of results.

After simulations are complete, model predictions may be viewed as time series output of flow, shortage and surplus at various watershed locations. Also, the shortage/surplus tool graphically displays the magnitude of shortages and available pass through water via a color coded map. Compliance with criteria set for the watershed is also viewable via map displays. Finally, the ZeroNet module includes a mechanism for exporting results for single deterministic or full sets of batch simulations.



The ZeroNet Goal - Meet rising electricity demand with zero net increase in power plant freshwater withdrawals by 2010

1. Specify scenario: D5T1test

Managed Flows	Diversion Cap Flow, cms	Adjust Flow by %	Modify Average Monthly Flows	Modify Time Series Data
Larkin Reynolds Ditch Div			☐	Edit LarkinReynolds.flo
Fruitland ditch from San			☐	Edit Fruitland.flo
City of Aztec diversion f			☐	Edit AztecFrLowerAnimas.Fl
Inca (Graves-Attebury) Di			☐	Edit IncaDitch.FLO
NIIP Diversion from Navaj			☐	Edit NIIPDiv.FLO
Release from Farmington L			☐	Edit FarmLakeWTP.flo
Release from Farmington L			☐	Edit FarmLakeRaw.flo
Release from Lemon Reserv			☐	Edit Lemon.flo
Release from Vallecito Re			☐	Edit Vallecito.flo
Release from Navajo Reser		-62	☐	Edit NavajoRes_1.flo

Figure 11-1
ZeroNet Module of WARMF

11.2.4 Study Watershed: San Juan Basin

The San Juan Basin was selected for demonstration of the modified WARMF. It is located in the four corners region of New Mexico, Colorado, Utah, and Arizona (Figure 11-2). Major tributaries of the San Juan River (the Piedra, Animas, and La Plata Rivers) drain melted snow pack from southern Colorado mountains into New Mexico. Drainage from the Upper San Juan and Piedra Rivers fill Navajo Reservoir, which holds approximately 1.8 million acre-feet of water and brings recreational benefits as well as water supply for drought management irrigation projects. The ZeroNet project focused primarily on the portion of the San Juan Basin in northwestern New Mexico.

A diverse set of stakeholders have immediate interest in the management of water resources in the New Mexico portion of the San Juan Basin including local, state, and federal government agencies, industry, Native American tribes, irrigation districts, and private individuals. In addition, there is a need to protect endangered aquatic species. The southwest U.S. has experienced five years of severe drought (2000-2004), during which reservoir levels continually have dropped. Stakeholders have been faced with the real situation of dwindling supplies and increased demand, including new in-stream flow recommendations for endangered species Efforts to mitigate drought impacts involved creation of a "shortage sharing" group in which stakeholders voluntarily agreed to curtail withdrawals by a set percentage.

Comprehensive analysis of drought and its impact in the basin is needed to support the decision making efforts of diverse stakeholders.

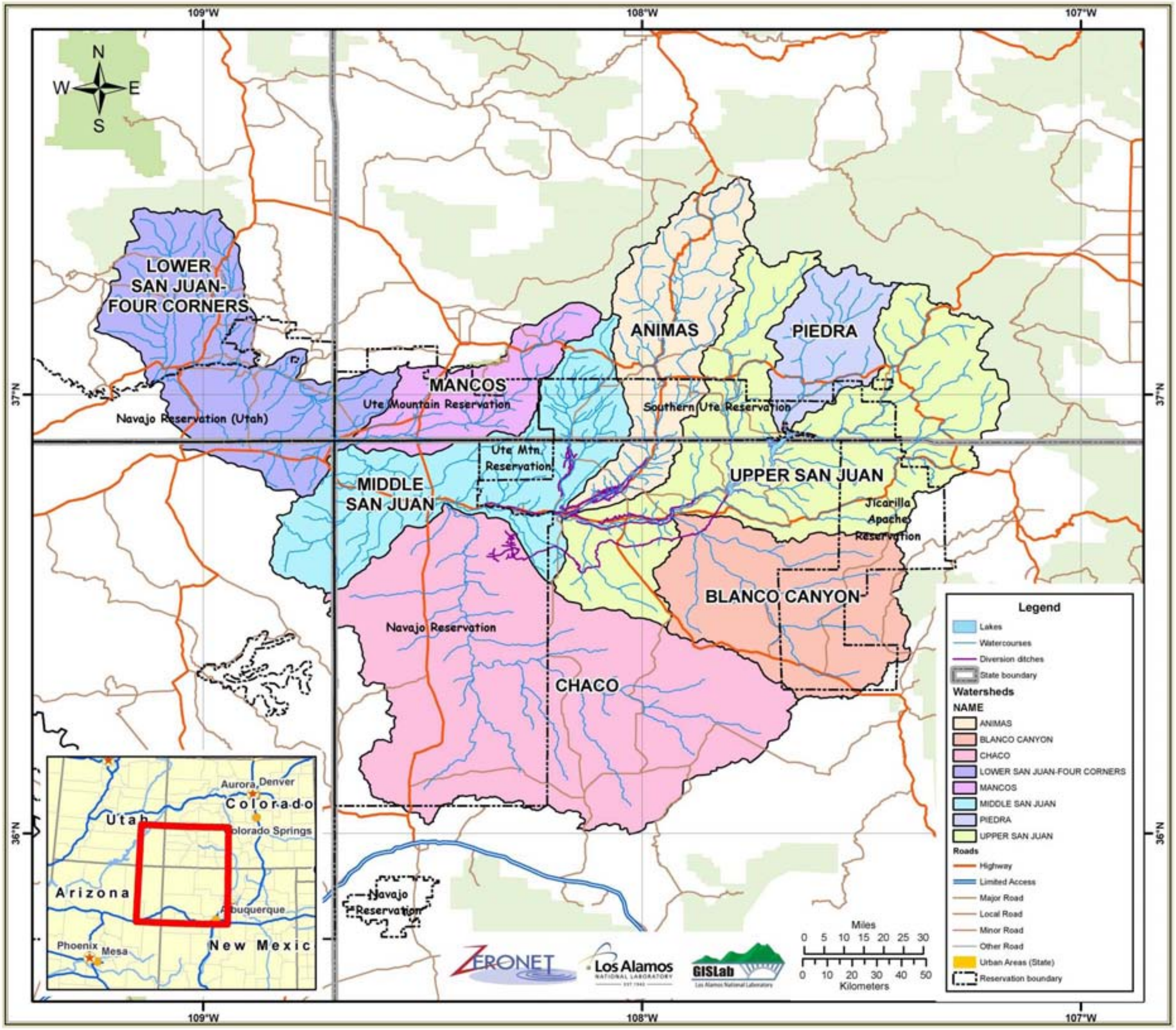


Figure 11-2
The San Juan Basin is located in the Four-Corners area where New Mexico, Colorado, Utah, and Arizona meet.

11.2.5 Simulations

WARMF was applied to the San Juan Basin by importing watershed data (e.g. topography, meteorology, land use, observed streamflow, diversions, point sources, and reservoir releases) and performing a hydrology calibration and validation for the time period of 1984 through 2004. After model calibration, a set of simulations was performed using the Batch Scenario Tool of WARMF to examine the impact of extended drought on water availability. Historic data for each year in the dataset (1984-2004) were organized and ranked as “wet”, “normal”, and “dry” based on annual average precipitation. The use of the tool is demonstrated by assessing impacts of a 3-year drought versus a 5-year drought, with 0, 1, and 2 degree temperature increase. A set of 6 scenarios were set up and run (Table 11-1). These scenarios represent a range from no climate change (0-degree temperature increase) to significant warming. A 2-degree increase in mean annual temperature is a high value often predicted by models such as the Community Climate Model (Kiehl et al. 1997) roughly 100 years into the future. In WARMF, this temperature change is applied as a uniform increase to historical minimum and maximum daily temperatures. When computing hydrology, these temperature data are used to dynamically calculate changes in snowmelt timing and the quantity of evapotranspiration.

Table 11-1
Scenario Matrix for Drought and Climate Simulations

Scenario	Temp. +/- (°C)	Year 1	Year 2	Year 3	Year 4	Year 5	Year 6	Year 7	Year 8	Year 9
1 – d3t0	0	Normal	Normal	Dry	Dry	Dry	Normal	Normal	na	na
2 – d3t1	1	Normal	Normal	Dry	Dry	Dry	Normal	Normal	na	na
3 – d3t2	2	Normal	Normal	Dry	Dry	Dry	Normal	Normal	na	na
4 – d5t0	0	Normal	Normal	Dry	Dry	Dry	Dry	Dry	Normal	Normal
5 – d5t1	1	Normal	Normal	Dry	Dry	Dry	Dry	Dry	Normal	Normal
6 – d5t2	2	Normal	Normal	Dry	Dry	Dry	Dry	Dry	Normal	Normal

Given the large year-to-year variation in climate data, random sampling of historical years will result in a large variation in the resulting water budget. Therefore, a repeated sampling approach was taken to reduce the model uncertainty. WARMF performed a set of runs for each scenario, sampling different combinations of "normal" and "dry" years. Preliminary testing was performed to determine the minimum number of iterations (runs) that is required in order to get a stabilized mean output. The ending elevation of Navajo Reservoir after the 3-year or 5-year drought was used as the main variable for evaluation. A comparison of the mean ending elevation as a function of the number of runs suggests that for a 3-year drought, mean ending elevation stabilizes after 50 runs. For a 5-year drought, the running mean of the elevation starts to stabilize after 50 runs, however it still shows slight variation. We also compared the cumulative distributions of elevations for three sets of runs with 25, 50 and 75 iterations. For the 3-year drought, the cumulative distribution of the ending elevation for the 50 runs is very close to that for the 75 runs. For the 5-year drought, the cumulative distribution of elevation for the 50 runs shows slight deviation from the 75 runs. With consideration of these results and associated simulation time to produce results, it was decided to use a repeated sampling of 50 runs to calculate a representative mean of the results for both the 3- and 5-year drought simulations.

The results were summarized as means and standard deviations of the 50 runs. Initially, all six scenarios yielded average reservoir elevations which violated the minimum elevation criteria of 5990 ft (1825 m) in Navajo Reservoir that must be met to accommodate the siphon pump for the Navajo Indian Irrigation Project (NIIP) diversion. Therefore, a reservoir release adjustment factor was introduced during the second (2°C scenarios) or third (0 and 1°C scenarios) year of drought to avoid minimum elevation violation. This factor reduces the historic prescribed reservoir releases by a user-specified percentage. A trial and error process was used to determine the approximate release adjustment that was necessary for each scenario so that the mean reservoir elevation would not fall below the minimum elevation criteria.

11.3 Results

The value of the historical sampling technique added to WARMF is demonstrated by looking at the large variability in predicted streamflow and reservoir elevation for the 50 runs of one scenario. For example, Figure 11-3 shows the predicted reservoir elevation for the D3T1 scenario at the end of the 3-year drought ranges from 1769.1 m to 1853.4 m for the 50 runs. Also, the predicted elevation for the whole simulation period also shows large variation, as suggested by the relatively large standard deviation (Figure 11-4).

Figures 11-5 and 11-6 show the mean projected elevation for Navajo Reservoir for each of the six scenarios. As stated in the figures, in order to keep the mean reservoir elevation above the minimum criterion, it was necessary to decrease the prescribed reservoir releases by 18% to 70% during specific drought years under various drought and temperature conditions. For the scenarios with zero and one degree temperature increases (D3T0, D3T1, D5T0, and D5T1), reservoir release reductions began during the third year of drought and continued through the end of the drought. For the scenarios with two degree temperature increases (D3T2 and D5T2), it was necessary to start the reservoir release adjustment during the second year and continue through the end of the drought period to avoid minimum reservoir elevation violations. The estimated percent reduction in reservoir releases was set to keep predicted mean elevation above minimum. For a more conservative approach, the reservoir release adjustments could have been set based on the mean reservoir elevation minus the standard deviation (Figure 11-4). This would have resulted in a larger percent reduction in reservoir elevation.

The adjustment in reservoir release will cause shortage downstream. Here we summarize the basin-wide shortage for the four scenarios, both on annual and cumulative bases (Table 11-2). Because the La Plata tributary watershed has historically experienced significant shortages during all years (including normal and wet years) due to over allocation and under delivery from Colorado, this portion of the watershed was excluded from the water shortage analysis.

11.4 Discussion

We successfully adapted the deterministic WARMF model to produce simulations based on repeated sampling of climatic and watershed data. If only a single deterministic run was performed, as represented by any single curve in Figure 11-4, the output could have fallen on the upper or lower end of the curve and could result in decision makers implementing a plan that would be either overly aggressive or ineffective. The repeated sampling approach enables us to examine output variability as a function of input variability. A set of drought scenarios sampled with sufficient iteration will provide both a mean result and a representation of variation about the mean, enabling more informed decisions. The limitations of this approach are two. First, the set of input data must satisfactorily represent a probability distribution for key parameters such as precipitation and temperature. And second, calculation time can limit the number of iterations that can be performed.

The scenario results show that under both three- and five-year drought conditions, significant modifications in reservoir release operations may be necessary to meet the minimum elevation criteria. Also, the watershed is likely to experience water shortages downstream of Navajo

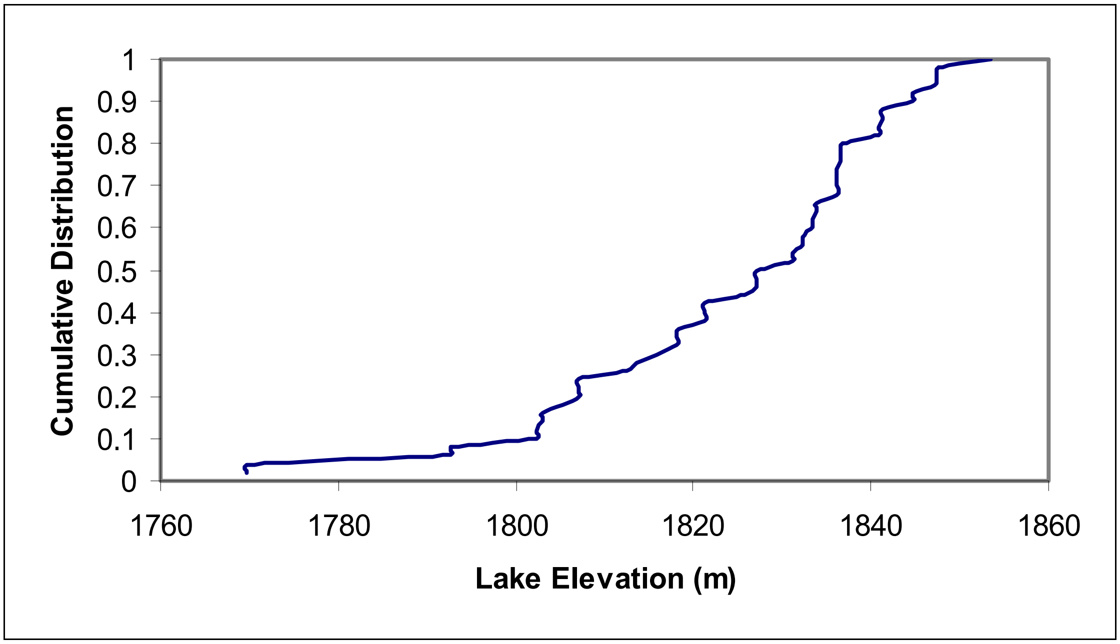


Figure 11-3
Projected cumulative distribution of surface elevation of Navajo Reservoir at the end of year 5 for a 3-year drought and 1-degree temperature increase (D3T1).

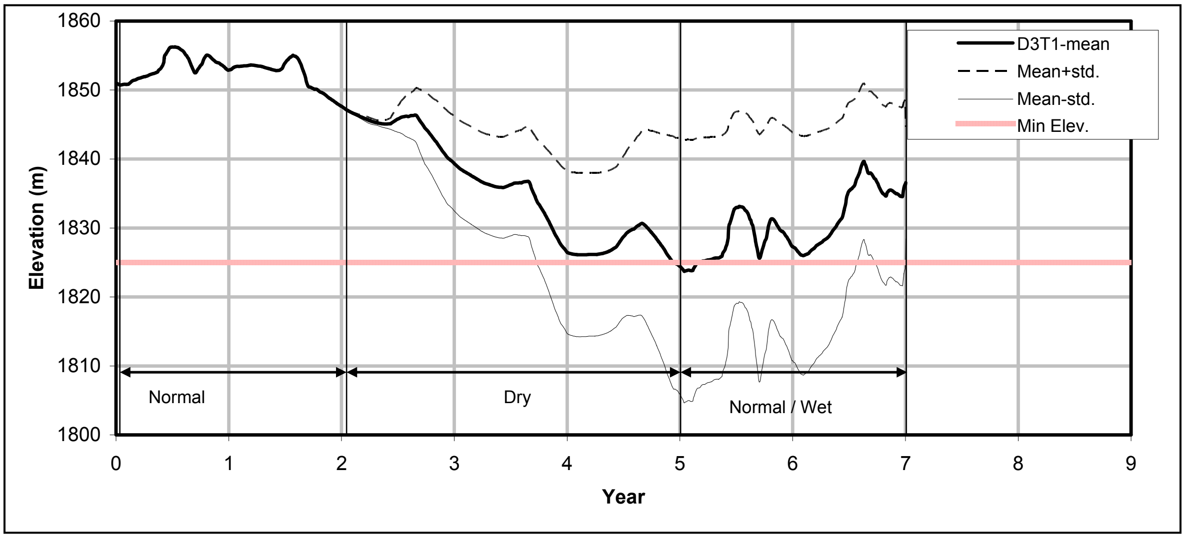


Figure 11-4
Projected mean surface elevation of Navajo Reservoir under a 3-year drought and 1-degree temperature increase (D3T1) with ± standard deviation

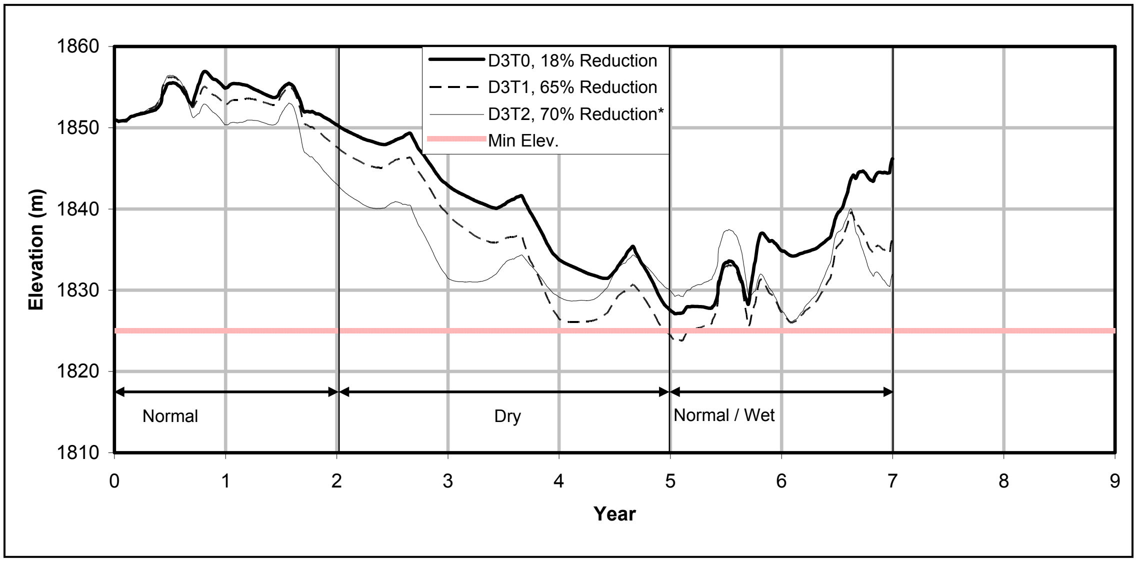


Figure 11-5
Projected Mean Surface Elevation of Navajo Res. for 3-Yr Drought Scenarios. Note: Reservoir release reduction was implemented for years 4 and 5 for scenario D3T2 and for year 5 only for scenarios D3T0 and D3T1.

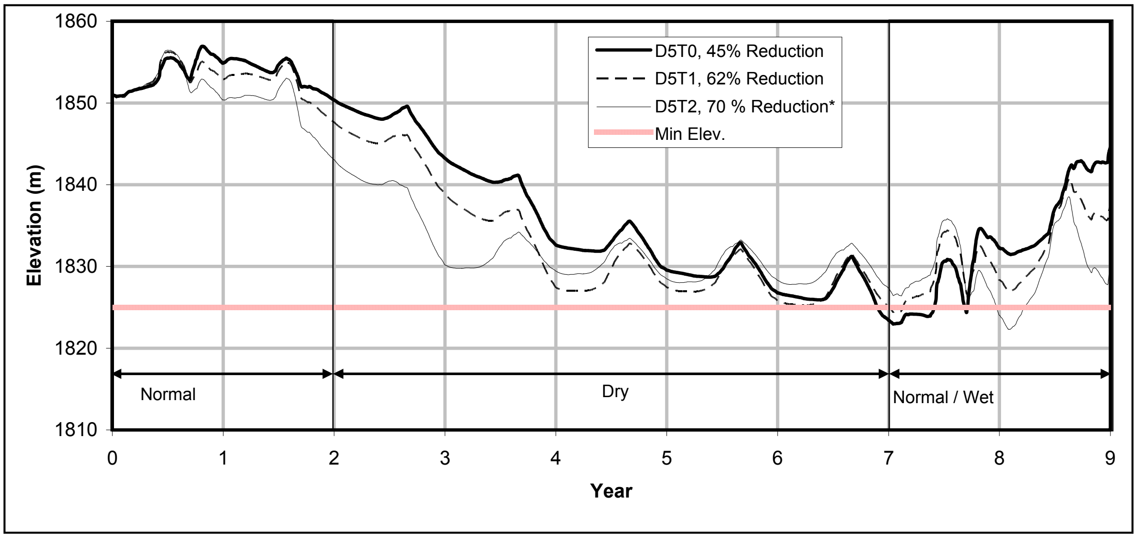


Figure 11-6
Projected Mean Surface Elevation of Navajo Res. for 5-Yr Drought Scenarios. Note: Reservoir release reduction was implemented for years 4, 5, 6, 7 for scenario D5T2 and for years 5, 6, 7 for scenarios D5T0 and D5T1.

Reservoir during drought conditions. A projected climate change with an increase in average temperature by just 1 or 2°C demonstrated a more severe impact on reservoir elevation and downstream shortages. The water shortage analysis shows a predicted shortage ranging from 266 to 62,071 AF/year under various scenario conditions. During the recent drought in the San Juan Basin, a reported "shortage sharing" agreement of approximately 3,400 AF/year effectively helped the region meet the minimum reservoir elevation criteria and eliminate the need to shut down lower priority diverters. This shortage falls within the range of shortages predicted by WARMF. WARMF could be used to further refine potential shortage sharing agreements in future drought situations. Likewise, the implementation of other management alternatives such as reduced power plant withdrawals due to the use of produced (degraded) water or advanced cooling techniques could be added to model scenarios. Though the approach demonstrated here used the mean calculated reservoir elevations to establish required reservoir release adjustments, the tool provides the capability to take a more conservative approach and base decisions on the mean +/- the calculated standard deviation for the multiple runs. Without a mechanism to efficiently set up multiple runs which sample the full range of potential drought conditions, the model would predict a single deterministic run which could be an upper or lower extreme condition. A "snap shot" analysis such as this could lead water managers to make planning decisions which are too conservative or not conservative enough.

Table 11-2
Projected Water Shortages for Drought / Climate Scenarios

Scenario	Reservoir Release Adjustment	Volume Held Back in Reservoir (AF)	Number of Years with Reservoir Release Adjustment	Total Shortage Downstream of Reservoir During Drought Period (AF)	Average Shortage Downstream of Reservoir (AF/yr)[1]
D3T0	18%	86,786	1	266	266
D3T1	65%	314,010	1	45,050	45,050
D3T2	70%	689,548	2	129,173	64,586
D5T0	45%	656,699	3	27,767	9,256
D5T1	62%	896,268	3	109,713	36,571
D5T2	70%	1,358,127	4	248,282	62,071

[1] Averaged for years when reservoir release adjustment was implemented.

Therefore, the approach presented above demonstrates a valuable improvement to an established watershed model, which is now capable of predicting climate change impacts using a historical sampling approach to build scenarios. Limitations of the approach taken include a sampling from a limited database of 20 years (1984-2004) for the watershed, and a lack of efficiency and flexibility in the trial and error method used for setting the reservoir release adjustment. In addition, a simplified approach was taken to account for climate change with the use of a uniform temperature increase. Large scale climate change models predict that warming may be accompanied by more precipitation in the southwest, but not more snowpack (Thomson et al. 2005). Increased evapotranspiration may or may not outweigh increases in precipitation and snowpack is predicted to melt sooner, and contain less water than in the past. These changes in precipitation and snowmelt patterns may have greater or lesser impact than the temperature induced evapotranspiration impact examined in this analysis. Never-the-less, this is the first tool that water managers, farmers and tribes have had access to, to even begin to examine potential water planning problems associated with climate change. WARMF will be extended to examine some of these other impacts in the future.

11.5 Conclusions

As part of the ZeroNet DSS development, a mature physically-based watershed simulation model, WARMF, was successfully enhanced to provide a mechanism for integrating critical water supply and demand information and assisting water owners and managers with planning for drought and climate change. WARMF can now execute a historic sampling technique via a batch tool to provide mean and standard deviation output of a set of simulation runs to characterize resulting stream flow, shortage, and surplus of water in watersheds faced with potential drought and climate change. The tool also provides a record of where violations of minimum stream flow or minimum reservoir elevation may occur under such conditions. The resultant scenario library generated by the Watershed Tools is linked with the other components of the ZeroNet DSS (Knowledge Base and Quick Scenario Tool). These other tools will provide mechanisms for organizing scenarios and performing related economic analysis.

The ZeroNet DSS was distributed to the stakeholders of the San Juan Basin for their use in water planning. Potential future work includes continued development of the ZeroNet Module of WARMF to include a more efficient method for calculating the required reservoir release adjustment for meeting minimum elevation criteria. Additional applications of the ZeroNet DSS to watersheds with other environmental characteristics and water use situations need to be conducted to test the robustness of the tool and analysis methods. Also, linkage of WARMF with climate change prediction data from a GCM could prove to be useful. Finally, the climate generation scenario technique presented here could be expanded beyond hydrology simulations to include water quality analyses.

11.6 Acknowledgements

The ZeroNet partnership includes participation from Los Alamos National Laboratory (LANL), the Electric Power Research Institute (EPRI), the Public Service Company of New Mexico (PNM), the University of New Mexico (UNM), Sandia National Laboratories (SNL), Systech Engineering, Inc., and the Department of Energy National Energy Technology Laboratory (DOE NETL).

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A

SECTION 3, INFRASTRUCTURE AVAILABILITY AND TRANSPORTATION ANALYSIS

Table A-1
Pipeline Installation and Operating Costs

Pipeline Installation and Operating Costs

PNM - Produced Water Project - SJGS

Installation Costs		
Charge Pumps		$580,000
Pavement Replacement		$110,000
Boring & Casing		$1,190,000
Pipe		$5,420,000
Valves, Fittings, etc.		$1,050,000
Lift Station		$580,000
Mob, Staking, Surveying, etc.		$850,000
Other		$920,000
Right of Way		$950,000
Design, Const Oversight		$1,250,000
Subtotal		$12,900,000
Contingency	15%	$1,940,000
NMGRT (4)	6.125%	$790,000
PNM G&A (5)	5.5%	$710,000
Total Installed Cost		$16,340,000
Annual Operating Costs		
Power (1)		$144,000
Operators (2)		$0
Maintenance (3)		$65,000
Total Operating Cost		$209,000

Notes…..

1. Offsite power at $0.05/kwh.
2. Operator coverage from SJGS and the Collection Center.
3. Maintenance at 0.5% of capital cost.
4. NMGRT is the New Mexico Gross Receipts Tax.
5. G&A is a "general and Administrative" charge applied to all PNM projects.

B

SECTION 4, TREATMENT & DISPOSAL ANALYSIS

Table B-1
Produced Water Treatment Alternatives Summary

Produced Water Treatment Alternatives Summary

PNM - Produced Water Project - SJGS

	Alt 1	Alt 2	Alt 3	Alt 4	Alt 5	Alt 6	Alt 7	Alt 8	Alt 9	Alt 10
	CRO	HERO	BC	CRO-BC	HERO-BC	CRO	HERO	BC	CRO-BC	HERO-BC
System Flows										
Produced Water, gpm (10)	1,216	1,216	1,160	1,216	1,216	1,216	1,216	1,060	1,216	1,216
Purge Water, gpm (1)	N/A	N/A	N/A	N/A	N/A	100	100	100	100	100
Softener Feed Rate, gpm	1,280	1,250	1,160	1,280	1,250	1,385	1,352	1,160	1,385	1,352
Water Lost to Sludge, gpm	1.6	2.6	N/A	1.6	2.6	7.4	8.6	N/A	7.4	8.6
UF Recycle, gpm (2)	63.9	N/A	N/A	63.9	N/A	68.9	N/A	N/A	68.9	N/A
Media Filter, WAC Recycle, gpm (2,4)	N/A	34.4	N/A	N/A	34.4	N/A	36.4	N/A	N/A	36.4
RO Feed Rate, gpm	1214.4	1247.8	N/A	1214.4	1247.8	1308.6	1343.9	N/A	1308.6	1343.9
RO Net Permeate, gpm (4)	935.1	1047.5	N/A	935.1	1047.5	981.5	1082.7	N/A	981.5	1082.7
RO Reject, gpm	279.3	174.7	N/A	279.3	174.7	327.2	234.5	N/A	327.2	234.5
BC Distillate, gpm	N/A	N/A	1009.1	133.0	126.8	N/A	N/A	998.9	173.7	171.9
Total Recovered Water,gpm	935.1	1047.5	1009.1	1068.2	1174.2	981.5	1082.7	998.9	1155.2	1254.6
BC Brine, gpm	N/A	N/A	150.9	146.3	47.9	N/A	N/A	161.1	153.5	62.6
Pretreatment										
Lime, $Ca(OH)_2$, tpd (1)	2.39	3.68	N/A	2.39	3.68	11.3	12.7	N/A	11.3	12.7
Coagulant Aide, ppd (1)	231	225	N/A	231	225	250	244	N/A	250	244
Cationic Polymer, ppd	46.1	45.1	N/A	46.1	45.1	49.9	48.8	N/A	49.9	48.8
Sludge (dry basis), tpd	5.02	8.38	N/A	5.02	8.38	23.8	27.6	N/A	23.8	27.6
Sludge Moisture Content	65%	65%	N/A	65%	65%	65%	65%	N/A	65%	65%
Thickened Sludge (wet basis), tpd	14.3	23.9	N/A	14.3	23.9	68.0	78.9	N/A	68.0	78.9
Sludge - $CaCO_3$ Content, tpd	4.05	6.91	N/A	4.05	6.91	15.9	19.1	N/A	15.9	19.1
Sludge - $Mg(OH)_2$ Content, tpd	0.48	0.65	N/A	0.48	0.65	6.48	6.66	N/A	6.48	6.66
93% Sulfuric Acid, H_2SO_4, tpd	19.9	0.58	N/A	19.9	0.58	9.05	0.65	N/A	9.05	0.65
Sodium Hypochlorite, NaOCl, ppd	76.9	75.1	N/A	76.9	75.1	83.2	81.3	N/A	83.2	81.3
Sodium Bisulfite, $NaHSO_3$, ppd	55.0	53.8	N/A	55.0	53.8	59.6	58.2	N/A	59.6	58.2
Anti-Scalant, ppd	76.9	0.00	N/A	76.9	0.00	83.2	0.00	N/A	83.2	0.00

Table B-1 (page 2 of 3)
Produced Water Treatment Alternatives Summary

Produced Water Treatment Alternatives Summary

PNM - Produced Water Project - SJGS

	Alt 1	**Alt 2**	**Alt 3**	**Alt 4**	**Alt 5**	**Alt 6**	**Alt 7**	**Alt 8**	**Alt 9**	**Alt 10**
	CRO	HERO	BC	CRO-BC	HERO-BC	CRO	HERO	BC	CRO-BC	HERO-BC
Pretreatment (continued)										
Strainer, UF Op Pressure, psi	100	N/A	N/A	100	N/A	100	N/A	N/A	100	N/A
Media Filter, WAC Op Pressure, psi	N/A	100	N/A	N/A	100	N/A	100	N/A	N/A	100
Power Requirement, kw (5)	90	90	N/A	90	90	100	90	N/A	100	90
RO System (3)										
Sodium Hydroxide, NaOH, tpd	N/A	0.91	N/A	N/A	0.91	N/A	1.44	N/A	N/A	1.44
Recovery	77.0%	83.9%	N/A	77.0%	83.9%	75.0%	80.6%	N/A	75.0%	80.6%
Permeate TDS, mg/l	270	260	N/A	270	260	260	270	N/A	260	270
Reject TDS, mg/l	51,400	61,800	N/A	51,400	61,800	46,100	60,400	N/A	46,100	60,400
Cleanings per Year (9)	18	1	N/A	18	1	18	1	N/A	18	1
Op Pressure, psi	400	400	N/A	400	400	400	400	N/A	400	400
Inter-Stage Op Pressure, psi	N/A	800	N/A	N/A	800	N/A	800	N/A	N/A	800
Power Requirement, kw (5)	560	510	N/A	560	510	620	570	N/A	620	570
Brine Concentrator										
93% Sulfuric Acid, H_2SO_4, tpd	N/A	N/A	22.4	0.39	0.00	N/A	N/A	20.7	0.18	0.00
Anti-Scalant, ppd	N/A	N/A	209	50.3	0.00	N/A	N/A	209	59.0	0.00
Calcium Chloride, $CaCl_2$, tpd	N/A	N/A	16.0	15.2	0.00	N/A	N/A	21.6	15.3	0.00
Recovery	N/A	N/A	86.99%	47.63%	72.57%	N/A	N/A	86.12%	53.10%	73.29%
Distillate TDS, mg/l	N/A	N/A	10	10	10	N/A	N/A	10	10	10
BC Op pH	N/A	N/A	5.0	5.0	11.0	N/A	N/A	5.0	5.0	11.0
BC Op Cl, mg/l	N/A	N/A	50,000	50,000	124,270	N/A	N/A	50,000	50,000	101,020
BC Op Total Solids, mg/l	N/A	N/A	117,800	115,200	225,000	N/A	N/A	126,000	114,700	225,000
Cleanings per Year	N/A	N/A	1	1	0.3	N/A	N/A	1	1	0.3
Power Requirement, kw (6)	N/A	N/A	4,830	640	610	N/A	N/A	4,780	840	830

Table B-1 (page 3 of 3)
Produced Water Treatment Alternatives Summary

Produced Water Treatment Alternatives Summary

PNM - Produced Water Project - SJGS

	Alt 1	**Alt 2**	**Alt 3**	**Alt 4**	**Alt 5**	**Alt 6**	**Alt 7**	**Alt 8**	**Alt 9**	**Alt 10**
	CRO	HERO	BC	CRO-BC	HERO-BC	CRO	HERO	BC	CRO-BC	HERO-BC
Overall										
Feed Rate, gpm (10)	1,216	1,216	1,160	1,216	1,216	1,316	1,316	1,160	1,316	1,316
Recovery	76.90%	86.14%	86.99%	87.84%	96.57%	74.58%	82.27%	86.12%	87.78%	95.33%
Recoverable Water, gpm	935	1,047	1,009	1,068	1,174	981	1,083	999	1,155	1,255
Total Power Requirement, kw	650	600	4,830	1,718	1,774	720	660	4,780	1,875	1,915
Recoverable Water TDS, mg/l	270	260	10	240	240	260	270	10	230	230
Wastewater to Evap Ponds, gpm	279	175	151	146	48	327	235	161	153	63
Additional Evap Ponds, acres (7,8)	140	87.3	75.4	73.1	24.0	114	67.3	30.5	26.7	0.00

Notes…..

1. N/A = not applicable, tpd = tons per day, ppd = pounds per day.
2. Recycle is sent to softener and comprised of UF bleed, media filter backwash and WAC spent regenerant.
3. All RO systems contain spiral wound, thin-film polyamide membranes.
4. A portion of the last stage RO permeate is used for filter backwash and WAC regeneration.
5. Includes 5% allowance for miscellaneous process power and rounded up to nearest 10 kw.
6. Assume 78.1 kwh/1,000 distillate. Includes 2% allowance for miscellaneous process power and rounded up to nearest 10 kw.
7. SJGS assumes that the equivalent of 2 gpm/acre evaporates from the ponds.
8. Alternatives 6 to 10 take a 50-acre credit for freed-up Purge Water capacity.
9. For alternatives with UF and CRO, assume 9 RO and 9 UF cleanings per year.
10. The flow basis is approximate and within the likely produced water recovery range.

Table B-2
Process Chemistry – Alternative 1 – Produced Water – CRO

		Average FW	5% UF Bleed (to R-C)	Lime Clarifier Feed	Lime Clarifier Effluent	UF Eff	Decarb Effluent	CRO Feed pH	1st Stg Perm	1st Stg Rej	2nd Stg Perm	2nd Stg Rej	3rd Stg Perm	3rd Stg Rej	Total CRO Perm
Flow Rate	gpm	1216	63.9	1279.9	1278.4	1214.4	1214.4	1214.4	470.6	743.9	288.2	455.6	176.3	279.3	935.1
Solids	tpd														
Waters of Moisture	gpm				1.55										
Na	mg/l_{CaCO3}	9858	390	9386	9386	9386	9386	9386	149	15229	241	24711	390	40063	223
K	mg/l_{CaCO3}	97.8	4.9	93.2	93.2	93.2	93.2	93.2	1.86	151	3.02	244.5	4.89	396	2.79
Ca	mg/l_{CaCO3}	198	0.70	188	52.5	52.5	52.5	52.5	0.26	85.5	0.43	139	0.70	227	0.40
Mg	mg/l_{CaCO3}	126	0.16	119	12.0	12.0	12.0	12.0	0.06	19.6	0.10	31.9	0.16	51.9	0.09
Ba	mg/l_{CaCO3}	0.05	0.00	0.05	0.01	0.01	0.01	0.01	0.00	0.02	0.00	0.04	0.00	0.06	0.00
Sr	mg/l_{CaCO3}	15.5	0.05	14.7	4.13	4.13	4.13	4.13	0.02	6.72	0.03	11.0	0.05	17.8	0.03
Fe	mg/l_{CaCO3}	20.2	0.00	19.2	0.10	0.10	0.10	0.10	0.00	0.16	0.00	0.27	0.00	0.43	0.00
HCO_3	mg/l_{CaCO3}	3019	4.64	2869	124	49.7	57.4	57.4	1.89	88.6	2.98	138.6	4.64	218.9	2.77
CO_3	mg/l_{CaCO3}	32.0	0.00	30.0	2366	0.00	0.06	0.06	0.00	0.15	0.00	0.38	0.00	0.98	0.00
Cl	mg/l_{CaCO3}	7111	355	6773	6773	6773	6773	6773	135	10972	219	17774	355	28770	203
Br	mg/l_{CaCO3}	8.95	0.45	8.52	8.52	8.52	8.52	8.52	0.17	13.8	0.28	22.4	0.45	36.2	0.26
NO_3	mg/l_{CaCO3}	1.60	0.08	1.53	1.53	1.53	1.53	1.53	0.03	2.48	0.05	4.01	0.08	6.49	0.05
SO_4	mg/l_{CaCO3}	196	36.8	188	188	2771	2771	2771	13.9	4515	22.6	7356	36.8	11976	20.9
Total SiO_2	mg/l	18.5	1.60	17.7	17.66	17.66	17.66	17.66	0.62	28.43	1.00	45.79	1.60	73.69	0.92
Total NH_3, mg/l_N	mg/l_N	15.8	0.38	15.0	15.03	15.03	15.03	15.03	0.06	24.50	0.15	39.91	0.38	64.86	0.15
Total Alk, mg/l_{CaCO3}	mg/l_{CaCO3}	3105	21.1	2951	2490	2490	68.1	68.1	19.5	98.9	20.1	149	21.1	229	20.0
B, mg/l_B	mg/l_B	2.51	0.22	2.40	2.40	2.40	2.40	2.40	0.08	3.86	0.14	6.21	0.22	10.0	0.12
o-PO_4, mg/l_P	mg/l_P	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
TDS	mg/l	13757	481	13095	11033	12020	12030	12030	183	19519	297	31673	481	51358	275
pH		7.84	5.72	7.84	11.09	4.40	6.82	6.82	5.33	7.02	5.52	7.23	5.72	7.44	5.49

System Net Recovery **76.90%**

Table B-2 (page 2 of 10)
Process Chemistry – Alternative 2 – Produced Water – HERO®

PNM – Produced Water Project – SJGS

		Average FW	5% UF Bleed (to R-C)	Lime Clarifier Feed	Lime Clarifier Effluent	UF Eff	Decarb Effluent	CRO Feed pH	1st Stg Perm	1st Stg Rej	2nd Stg Perm	2nd Stg Rej	3rd Stg Perm	3rd Stg Rej	Total CRO Perm
Flow Rate	gpm	1216	63.9	1279.9	1278.4	1214.4	1214.4	1214.4	470.6	743.9	288.2	455.6	176.3	279.3	935.1
Solids	tpd														
Waters of Moisture	gpm				1.55										
Na	mg/l_{CaCO3}	9858	390	9386	9386	9386	9386	9386	149	15229	241	24711	390	40063	223
K	mg/l_{CaCO3}	97.8	4.9	93.2	93.2	93.2	93.2	93.2	1.86	151	3.02	244.5	4.89	396	2.79
Ca	mg/l_{CaCO3}	198	0.70	188	52.5	52.5	52.5	52.5	0.26	85.5	0.43	139	0.70	227	0.40
Mg	mg/l_{CaCO3}	126	0.16	119	12.0	12.0	12.0	12.0	0.06	19.6	0.10	31.9	0.16	51.9	0.09
Ba	mg/l_{CaCO3}	0.05	0.00	0.05	0.01	0.01	0.01	0.01	0.00	0.02	0.00	0.04	0.00	0.06	0.00
Sr	mg/l_{CaCO3}	15.5	0.05	14.7	4.13	4.13	4.13	4.13	0.02	6.72	0.03	11.0	0.05	17.8	0.03
Fe	mg/l_{CaCO3}	20.2	0.00	19.2	0.10	0.10	0.10	0.10	0.00	0.16	0.00	0.27	0.00	0.43	0.00
HCO_3	mg/l_{CaCO3}	3019	4.64	2869	124	49.7	57.4	57.4	1.89	88.6	2.98	138.6	4.64	218.9	2.77
CO_3	mg/l_{CaCO3}	32.0	0.00	30.0	2366	0.00	0.06	0.06	0.00	0.15	0.00	0.38	0.00	0.98	0.00
Cl	mg/l_{CaCO3}	7111	355	6773	6773	6773	6773	6773	135	10972	219	17774	355	28770	203
Br	mg/l_{CaCO3}	8.95	0.45	8.52	8.52	8.52	8.52	8.52	0.17	13.8	0.28	22.4	0.45	36.2	0.26
NO_3	mg/l_{CaCO3}	1.60	0.08	1.53	1.53	1.53	1.53	1.53	0.03	2.48	0.05	4.01	0.08	6.49	0.05
SO_4	mg/l_{CaCO3}	196	36.8	188	188	2771	2771	2771	13.9	4515	22.6	7356	36.8	11976	20.9
Total SiO_2	mg/l	18.5	1.60	17.7	17.66	17.66	17.66	17.66	0.62	28.43	1.00	45.79	1.60	73.69	0.92
Total NH_3, mg/l_N	mg/l_N	15.8	0.38	15.0	15.03	15.03	15.03	15.03	0.06	24.50	0.15	39.91	0.38	64.86	0.15
Total Alk, mg/l_{CaCO3}	mg/l_{CaCO3}	3105	21.1	2951	2490	2490	68.1	68.1	19.5	98.9	20.1	149	21.1	229	20.0
B, mg/l_B	mg/l_B	2.51	0.22	2.40	2.40	2.40	2.40	2.40	0.08	3.86	0.14	6.21	0.22	10.0	0.12
o-PO_4, mg/l_P	mg/l_P	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
TDS	mg/l	13757	481	13095	11033	12020	12030	12030	183	19519	297	31673	481	51358	275
pH		7.84	5.72	7.84	11.09	4.40	6.82	6.82	5.33	7.02	5.52	7.23	5.72	7.44	5.49

System Net Recovery **76.90%**

Table B-2 (page 3 of 10)
Process Chemistry – Alternative 3 – Produced Water – BC 2 + BC 3

PNM – Produced Water Project – SJGS

		Produced Water	BC FW	BC Distillate	BC Brine
Flow Rate	gpm	1160	1160	1009.1	150.9
Solids	tpd				
Waters of Moisture	gpm				
Na	mg/l_{CaCO3}	9858	9858	8.55	75730
K	mg/l_{CaCO3}	97.8	97.8		752
Ca	mg/l_{CaCO3}	198	2265		17413
Mg	mg/l_{CaCO3}	126	126		966
Ba	mg/l_{CaCO3}	0.05	0.05		0.42
Sr	mg/l_{CaCO3}	15.5	15.5		119
Fe	mg/l_{CaCO3}	20.2	20.2		156
HCO_3	mg/l_{CaCO3}	3019	0.04		0.99
CO_3	mg/l_{CaCO3}	32.0	0.00		0.00
Cl	mg/l_{CaCO3}	7111	9178	8.55	70501
Br	mg/l_{CaCO3}	8.95	8.95		68.8
NO_3	mg/l_{CaCO3}	1.60	1.60		12.3
SO_4	mg/l_{CaCO3}	196	3298		25353
Total SiO_2	mg/l	18.5	18.5		142
Total NH_3, mg/l_N	mg/l_N	15.8	15.8		121
Total Alk, mg/l_{CaCO3}	mg/l_{CaCO3}	3105	3.00		23.1
B, mg/l_B	mg/l_B	2.51	2.51		19.3
o-PO_4, mg/l_P	mg/l_P	0.00	0.00		0.00
TDS	mg/l	13757	15331	10.0	117791
pH		7.84	4.50	7.00	5.00

System Net Recovery **86.99%**

Table B-2 (page 4 of 10)
Process Chemistry – Alternative 4 – Produced Water – CRO + BC 2 (Alternative 1 + BC 3)

PNM – Produced Water Project – SJGS

		Alternative 1 3rd Stg Rej	Total Conv RO Perm	BC FW	BC Distillate	BC Brine	Total Recov'd Water
Flow Rate	gpm	279.3	935.1	279.3	133.0	146.3	1068.2
Solids	tpd						
Waters of Moisture	gpm						
Na	mg/l_{CaCO3}	40063	223	40063	8.55	76494	196
K	mg/l_{CaCO3}	396	2.79	396		756	2.44
Ca	mg/l_{CaCO3}	227	0.40	8381		16004	0.35
Mg	mg/l_{CaCO3}	51.9	0.09	51.9		99.0	0.08
Ba	mg/l_{CaCO3}	0.06	0.00	0.06		0.12	0.00
Sr	mg/l_{CaCO3}	17.8	0.03	17.8		34.1	0.03
Fe	mg/l_{CaCO3}	0.43	0.00	0.43		0.83	0.00
HCO_3	mg/l_{CaCO3}	219	2.77	0.04		0.24	2.16
CO_3	mg/l_{CaCO3}	0.98	0.00	0.00		0.00	0.00
Cl	mg/l_{CaCO3}	28770	203	36924	8.55	70500	179
Br	mg/l_{CaCO3}	36.2	0.26	36.2		69.1	0.22
NO_3	mg/l_{CaCO3}	6.49	0.05	6.49		12.4	0.04
SO_4	mg/l_{CaCO3}	11976	20.9	12203		23301	18.3
Total SiO_2	mg/l	73.7	0.92	73.7		141	0.81
Total NH_3, mg/l_N	mg/l_N	64.9	0.15	64.9		124	0.13
Total Alk, mg/l_{CaCO3}	mg/l_{CaCO3}	229	20.0	3.00		5.73	17.5
B, mg/l_B	mg/l_B	10.0	0.12	10.0		19.1	0.11
o-PO_4, mg/l_P	mg/l_P	0.00	0.00	0.00		0.00	0.00
TDS	mg/l	51358	275	60353	10.0	115237	241
pH		7.44	5.49	4.50	7.00	5.00	5.50

BC Recovery	47.63%
System Net Recovery	87.84%

Table B-2 (page 5 of 10)
Process Chemistry – Alternative 5 – Produced Water – HERO® + BC 3 (Alternative 2 + BC 3)

PNM – Produced Water Project – SJGS

		Alternative 2 3rd Stg Rej	Total HERO Perm	BC FW	BC Distillate	BC Brine	Total Recov'd Water
Flow Rate	gpm	174.7	1047.5	174.7	126.8	47.9	1174.2
Solids	tpd						
Waters of Moisture	gpm						
Na	mg/l_{CaCO3}	51208	182	51208	8.55	186634	163
K	mg/l_{CaCO3}	661	3.00	661		2410	2.68
Ca	mg/l_{CaCO3}	0.00	0.00	0.00		0.00	0.00
Mg	mg/l_{CaCO3}	0.00	0.00	0.00		0.00	0.00
Ba	mg/l_{CaCO3}	0.00	0.00	0.00		0.00	0.00
Sr	mg/l_{CaCO3}	0.00	0.00	0.00		0.00	0.00
Fe	mg/l_{CaCO3}	0.00	0.00	0.00		0.00	0.00
HCO_3	mg/l_{CaCO3}	48.0	2.26	137		330	2.03
CO_3	mg/l_{CaCO3}	466	0.06	377		1543	0.04
Cl	mg/l_{CaCO3}	48077	218	48077	8.55	175223	196
Br	mg/l_{CaCO3}	60.5	0.27	60.5		221	0.24
NO_3	mg/l_{CaCO3}	10.9	0.05	10.9		39.6	0.04
SO_4	mg/l_{CaCO3}	2969	3.32	2969		10821	2.96
Total SiO_2	mg/l	123	0.99	123		447	0.88
Total NH_3, mg/l_N	mg/l_N	24.2	14.1	24.2		88.3	12.6
Total Alk, mg/l_{CaCO3}	mg/l_{CaCO3}	514	2.33	514		1873	2.08
B, mg/l_B	mg/l_B	16.6	0.13	16.6		60.6	0.12
o-PO_4, mg/l_P	mg/l_P	0.00	0.00	0.00		0.00	0.00
TDS	mg/l	61777	264	61763	10.0	225000	236
pH		10.77	8.63	10.77	7.00	11.00	8.63

BC Recovery	**72.57%**
System Net Recovery	**96.57%**

Table B-2 (page 6 of 10)
Process Chemistry – Alternative 6 – PW/PW Blend – CRO

PNM – Produced Water Project – SJGS

		Average FW	5% UF Bleed (to R-C)	Lime Clarifier Feed	Lime Clarifier Effluent	UF Eff	Decarb Effluent	CRO Feed pH	1st Stg Perm	1st Stg Rej	2nd Stg Perm	2nd Stg Rej	3rd Stg Perm	3rd Stg Rej	Total CRO Perm
Flow Rate	gpm	1316	68.9	1384.9	1377.5	1308.6	1308.6	1308.6	484.3	824.3	305.1	519.3	192.1	327.2	981.5
Solids	tpd														
Waters of Moisture	gpm				7.36										
Na	mg/l_{CaCO3}	9552	360	9095	9096	9096	9096	9096	145	14355	228	22655	360	35745	213
K	mg/l_{CaCO3}	107.6	5.1	102.5	102.5	102.5	102.5	102.5	2.05	162	3.23	254.6	5.09	401	3.01
Ca	mg/l_{CaCO3}	275	0.66	261	52.5	52.5	52.5	52.5	0.26	83.2	0.42	132	0.66	209	0.39
Mg	mg/l_{CaCO3}	1417	0.15	1347	12.0	12.0	12.0	12.0	0.06	19.0	0.10	30.1	0.15	47.7	0.09
Ba	mg/l_{CaCO3}	0.01	0.00	0.01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.01	0.00
Sr	mg/l_{CaCO3}	32.3	0.08	30.7	6.17	6.17	6.17	6.17	0.03	9.77	0.05	15.5	0.08	24.5	0.05
Fe	mg/l_{CaCO3}	18.8	0.00	17.9	0.10	0.10	0.10	0.10	0.00	0.16	0.00	0.25	0.00	0.40	0.00
HCO_3	mg/l_{CaCO3}	2797	1.74	2659	46	18.3	24.5	24.5	0.72	35.9	1.12	53.2	1.74	79.8	1.05
CO_3	mg/l_{CaCO3}	30.8	0.00	29.0	876	0.00	0.01	0.01	0.00	0.02	0.00	0.05	0.00	0.12	0.00
Cl	mg/l_{CaCO3}	6996	331	6665	6665	6665	6665	6665	133	10502	210	16549	331	26072	196
Br	mg/l_{CaCO3}	8.82	0.42	8.41	8.41	8.41	8.41	8.41	0.17	13.2	0.26	20.9	0.42	32.9	0.25
NO_3	mg/l_{CaCO3}	20.2	0.95	19.2	19.2	19.2	19.2	19.2	0.38	30.3	0.61	47.7	0.95	75.1	0.56
SO_4	mg/l_{CaCO3}	1603	32.9	1525	1525	2617	2617	2617	13.1	4146	20.7	6570	32.9	10409	19.3
Total SiO_2	mg/l	19.6	1.61	18.7	18.7	18.7	18.7	18.7	0.65	29.3	1.03	45.9	1.61	72.0	0.96
Total NH_3, mg/l_N	mg/l_N	16.6	0.13	15.8	15.8	15.8	15.8	15.8	0.02	25.0	0.05	39.7	0.13	63.0	0.05
Total Alk, mg/l_{CaCO3}	mg/l_{CaCO3}	2877	19.4	2735	922	922	36.8	36.8	18.8	47.4	19.1	64	19.4	90	19.0
B, mg/l_B	mg/l_B	12.1	0.99	11.5	11.5	11.5	11.5	11.5	0.40	18.1	0.63	28.4	0.99	44.4	0.59
o-PO_4, mg/l_P	mg/l_P	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
TDS	mg/l	15062	448	14336	11202	11709	11716	11716	180	18491	284	29184	448	46052	265
pH		7.84	5.26	7.84	11.09	4.40	6.39	6.39	4.90	6.57	5.08	6.77	5.26	6.97	5.05

System Net Recovery **74.58%**

Table B-2 (page 7 of 10)
Process Chemistry – Alternative 7 – PW/PW Blend – HERO®

PNM – Produced Water Project – SJGS

		Average FW	Media Filter B/W (to R-C)	WAC B/W (to R-C)	Avg WAC Regen+ S Rinse (to R-C)	Avg WAC F Rinse (to R-C)	Lime Clarifier Feed	Lime Clarifier Effluent	H-Form WAC Effluent	Decarb Effluent	HERO Feed pH Adjust	1st Stg Perm	1st Stg Rej	2nd Stg Perm	2nd Stg Rej	3rd Stg Perm	Net 3rd Stg Perm	3rd Stg Rej	Total HERO Perm
Flow Rate	gpm	1316	15.4	5.0	6.3	9.8	1352.4	1343.9	1343.9	1343.9	1343.9	593.1	750.8	331.3	419.5	185.0	158.3	234.5	1082.7
Solids	tpd																		
Waters of Moisture	gpm							8.55											
Na	mg/l_{CaCO3}	9552	406	406	406	406	9306	9305	8494	8494	8605	112	15314	204	27248	406		48421	179
K	mg/l_{CaCO3}	107.6	6.60	6.60	6.60	6.60	105	105	105	105	104.9	2.10	186	3.72	330	6.60		586	3.18
Ca	mg/l_{CaCO3}	275	0.00	0.00	32647	1632	433	52.5	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00		0.00	0.00
Mg	mg/l_{CaCO3}	1417	0.00	0.00	7462	373	1417	12.0	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00		0.00	0.00
Ba	mg/l_{CaCO3}	0.01	0.00	0.00	0.85	0.04	0.01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00		0.00	0.00
Sr	mg/l_{CaCO3}	32.3	0.00	0.00	3834	192	50.8	6.17	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00		0.00	0.00
Fe	mg/l_{CaCO3}	18.8	0.00	0.00	62.2	3.11	18.6	0.10	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00		0.00	0.00
HCO_3	mg/l_{CaCO3}	2797	2.13	2.13	0.00	0.00	2717	35	17	23	24.4	0.70	17.7	1.23	20.2	2.13		21.9	1.05
CO_3	mg/l_{CaCO3}	30.8	0.13	0.13	0.00	0.00	24.4	672	0.00	0.01	11.5	0.01	46	0.04	93	0.13		178	0.03
Cl	mg/l_{CaCO3}	6996	429	429	429	429	6819	6819	6819	6819	6819	136	12098	242	21463	429		38055	206
Br	mg/l_{CaCO3}	8.82	0.54	0.54	0.54	0.54	8.60	8.60	8.60	8.60	8.60	0.17	15.3	0.31	27.1	0.54		48.0	0.26
NO_3	mg/l_{CaCO3}	20.2	1.24	1.24	1.24	1.24	19.7	19.7	19.7	19.7	19.7	0.39	34.9	0.70	61.8	1.24		109.7	0.59
SO_4	mg/l_{CaCO3}	1603	28.62	28.62	46447	2322	1795	1795	1795	1795	1795	8.97	3205	16.03	5724	28.62		10217	13.67
Total SiO_2	mg/l	19.6	2.08	2.08	2.08	2.08	19.1	19.1	19.1	19.1	19.1	0.67	33.7	1.18	59.4	2.08		105	1.01
Total NH_3, mg/l_N	mg/l_N	16.6	21.8	21.8	21.8	21.8	16.7	16.7	16.7	16.7	16.7	10.8	21.4	19.3	23.1	21.8		24.2	14.6
Total Alk, mg/l_{CaCO3}	mg/l_{CaCO3}	2877	2.26	2.26	2.26	2.26	2800	707	707	35.9	35.9	0.72	64	1.27	113	2.26		200	1.09
B, mg/l_B	mg/l_B	12.1	1.28	1.28	1.28	1.28	11.8	11.8	11.8	11.8	11.8	0.41	20.8	0.73	36.7	1.28		64.6	0.62
o-PO_4, mg/l_P	mg/l_P	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00		0.0	0.00
TDS	mg/l	15062	558	558	63463	3683	14979	11537	10726	10733	10784	175	19151	312	34035	558		60442	267
pH		7.84	8.93	8.93	1.32	2.90	7.75	11.09	4.50	6.37	9.50	8.46	10.20	8.70	10.45	8.93		10.70	8.64

System Net Recovery **82.27%**

Table B-2 (page 8 of 10)
Process Chemistry – Alternative 8 – PW/PW Blend – BC 2 + BC 3

PNM – Produced Water Project – SJGS

		Produced Water	BC FW	BC Distillate	BC Brine
Flow Rate	gpm	1160	1160	998.9	161.1
Solids	tpd				
Waters of Moisture	gpm				
Na	mg/l_{CaCO3}	9552	9552	8.55	68748
K	mg/l_{CaCO3}	107.6	107.6		775
Ca	mg/l_{CaCO3}	275	3075		22145
Mg	mg/l_{CaCO3}	1417	1417		10208
Ba	mg/l_{CaCO3}	0.01	0.01		0.05
Sr	mg/l_{CaCO3}	32.3	32.3		233
Fe	mg/l_{CaCO3}	18.8	18.8		135
HCO_3	mg/l_{CaCO3}	2797	0.04		0.92
CO_3	mg/l_{CaCO3}	30.8	0.00		0.00
Cl	mg/l_{CaCO3}	6996	9796	8.55	70503
Br	mg/l_{CaCO3}	8.82	8.82		63.6
NO_3	mg/l_{CaCO3}	20.2	20.2		145.2
SO_4	mg/l_{CaCO3}	1603	4477		32243
Total SiO_2	mg/l	19.6	19.6		141
Total NH_3, mg/l_N	mg/l_N	16.6	16.6		120
Total Alk, mg/l_{CaCO3}	mg/l_{CaCO3}	2877	3.00		21.6
B, mg/l_B	mg/l_B	12.1	12.1		87.2
o-PO_4, mg/l_P	mg/l_P	0.00	0.00		0.00
TDS	mg/l	15062	17500	10.0	125984
pH		7.84	4.50	7.00	5.00

System Net Recovery **86.12%**

Table B-2 (page 9 of 10)
Process Chemistry – Alternative 9 – PW/PW Blend – CRO + BC 3 (Alternative 6 + BC 3)

PNM – Produced Water Project – SJGS

		Alternative 6 3rd Stg Rej	Total Conv RO Perm	BC FW	BC Distillate	BC Brine	Total Recov'd Water
Flow Rate	gpm	327.2	981.5	327.2	173.7	153.5	1155.2
Solids	tpd						
Waters of Moisture	gpm						
Na	$\text{mg/l}_{\text{CaCO3}}$	35745	213	35745	8.55	76199	182
K	$\text{mg/l}_{\text{CaCO3}}$	401	3.01	401		855	2.56
Ca	$\text{mg/l}_{\text{CaCO3}}$	209	0.39	7209		15369	0.33
Mg	$\text{mg/l}_{\text{CaCO3}}$	47.7	0.09	47.7		101.8	0.08
Ba	$\text{mg/l}_{\text{CaCO3}}$	0.01	0.00	0.01		0.01	0.00
Sr	$\text{mg/l}_{\text{CaCO3}}$	24.5	0.05	24.5		52.3	0.04
Fe	$\text{mg/l}_{\text{CaCO3}}$	0.40	0.00	0.40		0.85	0.00
HCO_3	$\text{mg/l}_{\text{CaCO3}}$	80	1.05	0.04		0.27	0.69
CO_3	$\text{mg/l}_{\text{CaCO3}}$	0.12	0.00	0.00		0.00	0.00
Cl	$\text{mg/l}_{\text{CaCO3}}$	26072	196	33072	8.55	70500	168
Br	$\text{mg/l}_{\text{CaCO3}}$	32.9	0.25	32.9		70.1	0.21
NO_3	$\text{mg/l}_{\text{CaCO3}}$	75.1	0.56	75.1		160	0.48
SO_4	$\text{mg/l}_{\text{CaCO3}}$	10409	19.3	10496		22378	16.4
Total SiO_2	mg/l	72.0	0.96	72.0		153	0.81
Total NH_3, mg/l_{N}	mg/l_{N}	63.0	0.05	63.0		134	0.05
Total Alk, $\text{mg/l}_{\text{CaCO3}}$	$\text{mg/l}_{\text{CaCO3}}$	90	19.0	3.00		6.40	16.2
B, mg/l_{B}	mg/l_{B}	44.4	0.59	44.4		94.7	0.50
o-PO_4, mg/l_{P}	mg/l_{P}	0.00	0.00	0.00		0.00	0.00
TDS	mg/l	46052	265	53803	10.0	114697	226
pH		6.97	5.05	4.50	7.00	5.00	5.00

BC Recovery	**53.10%**
System Net Recovery	**87.78%**

Table B-2 (page 10 of 10)
Process Chemistry – Alternative 10 – PW/PW Blend – HERO® + BC 3 (Alternative 7 + BC 3)

PNM – Produced Water Project – SJGS

		Alternative 7 3rd Stg Rej	Total HERO Perm	BC FW	BC Distillate	BC Brine	Total Recov'd Water
Flow Rate	gpm	234.5	1082.7	234.5	171.9	62.6	1254.6
Solids	tpd						
Waters of Moisture	gpm						
Na	mg/l_{CaCO3}	48421	179	48421	8.55	181249	156
K	mg/l_{CaCO3}	586	3.18	586		2192	2.74
Ca	mg/l_{CaCO3}	0.00	0.00	0.00		0.00	0.00
Mg	mg/l_{CaCO3}	0.00	0.00	0.00		0.00	0.00
Ba	mg/l_{CaCO3}	0.00	0.00	0.00		0.00	0.00
Sr	mg/l_{CaCO3}	0.00	0.00	0.00		0.00	0.00
Fe	mg/l_{CaCO3}	0.00	0.00	0.00		0.00	0.00
HCO_3	mg/l_{CaCO3}	21.9	1.05	61		132	0.91
CO_3	mg/l_{CaCO3}	178	0.03	139		618	0.02
Cl	mg/l_{CaCO3}	38055	206	38055	8.55	142440	179
Br	mg/l_{CaCO3}	48.0	0.26	48.0		180	0.22
NO_3	mg/l_{CaCO3}	110	0.59	110		411	0.51
SO_4	mg/l_{CaCO3}	10217	13.7	10217		38250	11.8
Total SiO_2	mg/l	105	1.01	105		392	0.87
Total NH_3, mg/l_N	mg/l_N	24.2	14.6	24.2		90.5	12.6
Total Alk, mg/l_{CaCO3}	mg/l_{CaCO3}	200	1.09	200		750	0.94
B, mg/l_B	mg/l_B	64.6	0.62	64.6		242	0.54
o-PO_4, mg/l_P	mg/l_P	0.00	0.00	0.00		0.00	0.00
TDS	mg/l	60442	267	60130	10.0	225001	231
pH		10.70	8.64	10.70	7.00	11.00	8.63

BC Recovery	**73.29%**
System Net Recovery	**95.33%**

Table B-3
Produced Water Treatment Cost Summary – Preliminary Cost Evaluation

Produced Water Treatment Cost Summary - Preliminary Cost Evaluation

PNM - Produced Water Project - SJGS

	Alt 1	**Alt 2**	**Alt 3**	**Alt 4**	**Alt 5**	**Alt 6**	**Alt 7**	**Alt 8**	**Alt 9**	**Alt 10**
	CRO	HERO	BC	CRO-BC	HERO-BC	CRO	HERO	BC	CRO-BC	HERO-BC
Capital Costs - SJGS Only										
Receiving, Transfer, Distribution	$1,478,000	$1,478,000	$1,478,000	$1,478,000	$1,478,000	$1,478,000	$1,478,000	$1,478,000	$1,478,000	$1,478,000
Pretreatment + CRO	$7,310,000	$0	$0	$7,310,000	$0	$7,670,000	$0	$0	$7,670,000	$0
Pretreatment + HERO	$0	$6,390,000	$0	$0	$6,390,000	$0	$6,700,000	$0	$0	$6,700,000
Refurb BC 2	$0	$0	$4,100,000	$0	$0	$0	$0	$4,100,000	$0	$0
Refurb BC 3	$0	$0	$2,970,000	$2,970,000	$2,970,000	$0	$0	$2,970,000	$2,970,000	$2,970,000
Evap Ponds	$27,610,000	$16,890,000	$14,450,000	$13,970,000	$4,100,000	$22,260,000	$12,770,000	$5,240,000	$4,570,000	$0
Subtotal	$36,398,000	$24,758,000	$22,998,000	$25,728,000	$14,938,000	$31,408,000	$20,948,000	$13,788,000	$16,688,000	$11,148,000
6.125% New Mexico Sales Tax	$2,229,000	$1,516,000	$1,409,000	$1,576,000	$915,000	$1,924,000	$1,283,000	$845,000	$1,022,000	$683,000
5.5% PNM A&G	$2,002,000	$1,362,000	$1,265,000	$1,415,000	$822,000	$1,727,000	$1,152,000	$758,000	$918,000	$613,000
15% Contingency	$5,460,000	$3,714,000	$3,450,000	$3,859,000	$2,241,000	$4,711,000	$3,142,000	$2,068,000	$2,503,000	$1,672,000
Total Install Cost	$46,089,000	$31,350,000	$29,122,000	$32,578,000	$18,916,000	$39,770,000	$26,525,000	$17,459,000	$21,131,000	$14,116,000
Operating Costs - SJGS Only										
Chemicals	$802,000	$168,000	$2,020,000	$1,950,000	$168,000	$652,000	$392,000	$2,378,000	$1,804,000	$392,000
Power	$228,000	$210,000	$1,692,000	$452,000	$424,000	$252,000	$231,000	$1,675,000	$547,000	$522,000
UF/RO Membrane Cleaning	$240,000	$10,000	$0	$240,000	$10,000	$240,000	$10,000	$0	$240,000	$10,000
BC Membrane Cleaning	$0	$0	$52,000	$26,000	$7,000	$0	$0	$52,000	$26,000	$7,000
Labor (same for all)	$499,000	$499,000	$499,000	$499,000	$499,000	$499,000	$499,000	$499,000	$499,000	$499,000
Maintenance (process eqpmt)	$132,000	$118,000	$22,000	$132,000	$118,000	$137,000	$123,000	$22,000	$137,000	$123,000
Maintenance (refurb'd BCs)	$0	$0	$90,000	$45,000	$45,000	$0	$0	$90,000	$45,000	$45,000
Capital Recovery	$4,521,000	$3,075,000	$2,857,000	$3,196,000	$1,856,000	$3,901,000	$2,602,000	$1,713,000	$2,073,000	$1,385,000
Total Operating Cost - SJGS Only	$6,422,000	$4,080,000	$7,232,000	$6,540,000	$3,127,000	$5,681,000	$3,857,000	$6,429,000	$5,371,000	$2,983,000

Note….. The flow basis is approximate and within the likely produced water recovery range.

Table B-4
Produced Water Treatment – Operating and Cost Assumptions – Preliminary Cost Evaluation

Produced Water Treatment - Operating and Cost Assumptions - Preliminary Cost Evaluation

San Juan Generating Station

1.	Lime, $Ca(OH)_2$, $/ton	$86.02	
2.	Coagulant Aide, $/pound	$0.15	
3.	Cationic Polymer, $/pound	$1.00	
4.	Limestone, $/ton	$18.55	Plant cost - delivered to SJGS.
5.	93% Sulfuric Acid, H_2SO_4, $/ton	$90.55	Plant cost - delivered to SJGS.
6.	Sodium Hydroxide, NaOH (dry basis), $/ton	$77.50	Plant cost - delivered to SJGS.
7.	Sodium Hypochlorite, NaOCl, $/pound	$0.80	
8.	Sodium Bisulfite, $NaHSO_3$, $/pound	$0.20	
9.	Anti-Scalant, $/pound	$1.50	
10.	Calcium Chloride, $CaCl_2$, $/ton	$200	
11.	Limestone credit (plant cost for delivered limestone) applied for each ton of $CaCO_3$ generated in pretreatment sludge.		
12.	UF, RO Cleaning Cost	$10,000	
13.	BC Cleaning Cost (per BC)	$26,000	
14.	BC Power, kwh/1,000 gal distillate	78.1	
15.	Power, $/kwh	$0.0400	Includes $0.025/kwh plant power generation cost + $15/Mwh power replacement cost.
16.	Maintenance Worker Time, hours/year	2,080	
17.	Operator Time, hours/year	8,760	
18.	Loaded Labor Cost, $/hour	$46.00	
19.	Maintenance, Pct of Eqpmt Cost	1.50%	
20.	Interest on Capital	7.50%	
21.	Capital Recovery Period, years	20	
22.	Capital Recovery Factor	0.0981	
23.	Evap Pond Cost, $/acre	$171,000	First 30 acres of ponds.
24.	Evap Pond Cost, $/acre	$205,000	Pond area in excess of 30 acres - additional 20% for wastewater piping & remote monitoring.
25.	Receiving, Transfer & Distribution Equipment		
	Receiving Basin	$298,000	3-day basin x 10-feet operating depth and 1,316 gpm inflow (1.74 acre pond).
	Feed Pumps	$272,000	Four 33% capacity, 440 gpm x 50 psi pumps with valves & controls.
	Product Tank	$280,000	One 250,000 gallon tank with valves & controls.
	Product Transfer Pumps	$368,000	Four 33% capacity, 440 gpm x 50 psi pumps (316 impellors) with valves & controls.
	Product Transfer Line	$260,000	5000-feet 8" HDPE line routed to cooling towers with valves & controls.
	Total Plant Improvements	$1,478,000	

Table B-4 (page 2 of 2)
Produced Water Treatment – Operating and Cost Assumptions – Preliminary Cost Evaluation

Produced Water Treatment - Operating and Cost Assumptions - Preliminary Cost Evaluation
PNM - Produced Water Project - SJGS

		1,216 gpm System	1,316 gpm System	
26.	Pretreatment, UF & CRO Equipment	$5,090,000	$5,340,000	Includes control system.
27.	Pretreatment & HERO Equipment	$4,440,000	$4,660,000	
28.	Equipment Installation Factor	0.436	0.436	Cost factor to install pretreatment, UF, RO equipment.
29.	Refurbish BC 2	$4,100,000	Includes demolition & assembly.	
30.	Refurbish BC 3	$2,970,000	Includes demolition & assembly.	
31.	Refurbished BC Valuation	$3,000,000	Value basis used to estimate annual BC maintenance.	

Note….. The flow basis is approximate and within the likely produced water recovery range.

Table B-5
Capital Cost Estimate

Capital Cost Estimate

Collection Center in Bloomfield

Major Processes			Equipment	Installation	Total
O/W Gravity/Coalescing Separator			$220,000	$100,000	$320,000
Gas Flotation Unit			$300,000	$140,000	$440,000
Walnut Shell Filter			$480,000	$220,000	$700,000
Hold Basin					$240,000
PLC/HMI					$50,000
MCCs					$170,000
Tanks					
Receiving Tanks (2)					$600,000
Walnut Shell Filter Feed Tank					$50,000
Backwash Hold Tank					$20,000
Off-Spec Hold Tank					$290,000
Off-Spec Waste Tank					$20,000
Recovered Oil Tank					$20,000
Total - Level Indicators, Misc Valves					$110,000
Tank Insulation					$90,000
Line Insulation					$90,000
Pumps					
Receiving Tank Transfer Pumps			$130,000	$60,000	$190,000
Walnut Shell Filter Feed Pumps			$130,000	$60,000	$190,000
Walnut Shell Filter Recirc Mixer/Pump (included with filter)					
Backwash Transfer Pumps			$10,000	$5,000	$15,000
Off-Spec Transfer Pumps			$10,000	$5,000	$15,000
Pipeline Charge Pumps			$250,000	$110,000	$360,000
Miscellaneous					
Electric Tank Heaters			$30,000	$10,000	$40,000
Total - control valves, manual valves			$290,000	$130,000	$420,000
Transformer/Switchgear					$200,000
Office, Control Room, MCC Room & Shop/Storage Area					$50,000
Add'l Site Grading @	5%	of installed cost			$260,000
General Civil @	5%	of installed cost			$260,000
Freight @	1.5%	of equipment cost			$30,000
				Total	$5,240,000

Table B-6
Operating Cost Estimate

Operating Cost Estimate

Collection Center in Bloomfield

			kwh/day
Gas Flotation Drive Motors		$22,200	1,216
Receiving Tank Transfer Pumps		$12,600	689
Walnut Shell Filter Feed Pumps		$12,600	689
Walnut Shell Filter Recirc Mixer		$900	47
Backwash Transfer Pumps		$300	15
Off-Spec Transfer Pumps		$500	30
Electric Tank Heaters		$29,600	1,622
Misc Power	3% of total	$2,400	129
Total Power		$81,100	4,437

Total Annual Power Cost	$81,000	(rounded)
Offsite Power Cost, $/kwh	$0.050	
Power Demkand, kw	184.9	

Chemicals		Dose mg/l	Usage #/day	Unit Cost $/pound
O/W Media Pack Change-out	$5,000			
Walnut Shell Filter Media	$800			
Emulsion Breaker	$21,700	5	59.6	$1.00
Filter Aide	$21,700	5	59.6	$1.00
Biocide	$42,500			
Total Chemicals + Mat'ls	$91,700			

Total Annual Chemicals + Materials	$92,000	(rounded)

Labor Summary		Burdened Rate $/hr	Staffing hr/yr
Operators	$201,480	$46	4380
Maintenance Techs	$95,680	$46	2080
Clerical	$0	$15	0
Supervisor	$0	$55	0
Annual Labor Cost	$297,160		

Total Annual Labor	$297,000	(rounded)
Equipment Maintenance	1.5%	of Installed Equipment
	$70,000	(rounded)

C

SECTION 5, EMERGING TECHNOLOGY TESTING

CeraMem Test Report

Membrane Filtration of Produced Water

Introduction

Produced water treatment

Produced water cleaning is notoriously challenging for separation processes. Treatment of produced water has employed conventional water filtration equipment such as sand bed and dual media filters, diatomaceous earth pre-coat filters, and cartridge filters. However, these techniques can be expensive, require appreciable space and are prone to operating problems. Because of this, there has been much interest in use of cross-flow membrane filters[i] for treatment of produced waters and similar wastes. Spiral-wound filters would seem an obvious membrane process selection for a given wastewater application due their comparatively low cost and high "turbulence promotion"[ii]. However, a tubular type design may be the best approach when dealing with relatively high levels of insoluble solids or when concentrating the level of solids to a reasonably high level inside the membrane process. In processes with high feed solids, spiral wound or small flow-path tubular (hollow fiber) membranes may experience flow path plugging.

A report[iii] for the Offshore Operators Committee evaluated five alternative cross-flow filtration systems for removal of oil and grease from produced water. None of these systems was judged to be entirely satisfactory for commercial operation, for reasons ranging from inadequate field testing, to serious operational shortcomings. Such shortcomings included the following:

- Polymeric membranes considered in the study required upstream pre-filtration of suspended solids by backwashable dead-ended filters, effectively negating the continuous nature of cross-flow filtration.
- The only ceramic modules evaluated required continuous addition of pretreatment chemicals to ensure proper operation.[iv] This made the process very sensitive to normal fluctuations in water chemistry and flow rate, and left oil-wet solids in the concentrate bleed that were judged to be difficult to treat further.
- Relatively rapid irreversible flux degradation, which can significantly affect operating costs due to the resulting need to remove modules from service and clean them frequently, was problematic for all modules.

Another study[v] has further examined de-oiling of produced water in the field by three different cross-flow membrane modules and found severe long-term fouling in each. The fouling mechanisms identified in these tests included: inorganic solids deposition, oil wetting, scale formation by salts present in the produced water, pore clogging by gelatinous material, and membrane blinding by flocs of ferric oxide.

In the past, CeraMem Corporation has had discussions with engineering staff at an oil company which has had many years of experience with processing of produced water for enhanced oil recovery with ceramic membranes.[vi] Their engineers stated that, for ceramic membranes used for treatment of produced water, the most troublesome foulant that builds on membrane surfaces is a layer of dense amorphous silica, which is formed slowly from soluble silica (used for enhanced oil recovery at the oil well) present in the wastewater. The formation of the layer is affected by the feed chemistry as well as the surface properties of the membrane surface. CeraMem has developed specialized filter technology and cleaning techniques to achieve stable process fluxes in the presence of highly fouling silica containing produced waters in conjunction with the oil company with whom this development work was done.

Membrane technology

Membrane filtration is defined as a pressure-driven, primarily size-based separation technique where one or more soluble or insoluble components are removed from a fluid carrier stream by the membrane. The process is commonly used in industry for the separation or fractionation of multiple types of immiscible or dissolved species present in a liquid or gaseous feed stream (Cheryan, 1998[vii]). Concentration of single-solute feed streams is also a wide spread application of filtration. The convention is to use a feed stream, which flows tangentially across the membrane surface, termed the cross-flow operation mode. In its simplest form, the separation process employs the membrane as a sieve to remove dissolved and suspended species from the feed stream, producing a concentrated, solute-rich retentate stream and a dilute or solute-free permeate or filtrate stream. The filtrate or permeate flux is generally reported as volumetric flux per unit surface area of membrane and has dimensions of velocity ($m.s^{-1}$) or more commonly liters/m^2/hour (lmh) or gallons/ft^2/day (gfd).

Permeate flux decline is a key issue of concern for the filtration process. It is observed that every time a membrane is used in a constant driving pressure operation, the filtration flux rate decreases from the initial value to a lower, steady or slightly declining level. Concentration polarization (initially) and the build-up of rejected solute at the membrane surface, the so-called cake or fouling layer, are responsible for this trend. At some point in time, the permeate flux may become too low for the process to be economically viable. The membrane then either has to be cleaned or, in extreme cases, replaced in order to restore the permeate flux. Flux decline thus reduces the overall efficiency of a filtration process by reducing the filtration rate (i.e. a lower rate of processing and product recovery) and introducing costs of cleaning and/or replacing membranes. This inherent property of membrane filtration often has been seen as the factor limiting wider-spread application of membrane-based separation processes in industry.

Flux maintenance techniques have, however, evolved rapidly over the last three decades and many in-line cleaning techniques have been developed for membrane processes. Typical enhancement techniques employed in industry include (i) fast-flushing or intermittent feed velocity acceleration – which aims to clean the fouling layer off the membrane wall by high wall shear stresses; (ii) back-flushing or intermittent blowback of clean permeate back through the membrane in the opposite direction of regular permeate flow – this aims to lift the fouling layer off the membrane surface and inside the membrane pores, by high velocity of back-flushed flow; (iii) gas-bubble injection on the membrane feed side – which introduces complex and turbulent flow patterns which have been observed to reduce the thickness of the fouling layer on the membrane surface. In-line cleaning technology is crucial as it reduces both manpower and downtime for the membrane process and, over the last 15 years, membrane technology has become the process of choice for tackling a wide range of industrial separation challenges, e.g. wastewater treatment and potable water production.

Aqueous oil solutions

Typically three forms of oil exist in aqueous solutions: (i) free oil, which floats on the solution's surface because of lower density, (ii) soluble oil and (iii) emulsified oil. Generally, thermodynamically stable mineral oil-water emulsoids are of the order 10 nm in size (microemulsiods)[viii]. Most process industry emulsions e.g. food industry, are thermodynamically unstable and require some form of stabilization. The stabilized emulsoids are typically larger (nominally 100-1000 nm) than microemulsoids and size depends on the fluids, emulsifier/stabilizers and process conditions. Emulsified oil is generally the only aqueous-oil component that can be effectively rejected by an ultrafiltration membrane. Soluble oil generally passes through membranes as part of the carrier fluid and could foul the membrane, while free oil would foul the membrane severely by coating and forming complexes at the membrane surface with other compounds. Soluble oil often can lead to turbidity detections in a permeate stream by forming droplets (unstable emulsoids), but can also pass through a membrane completely solubilized in the aqueous phase and show low turbidity[ix]. Often an unstable emulsion, can be severely fouling as it comprises larger, unstable oil emulsoids (generally up to 5 μm in size[viii]). Large, unstable emulsoids have a tendency of coalescing, which causes further increases in size. Once large enough, these unstable emulsoids float upward due to buoyancy and become free oil. Large emulsoids can be emulsified more efficiently by mechanical means, e.g. pumping, which essentially reduces emulsoid-size and increases oil-water interfacial area. Mechanical emulsification must be quickly followed by addition of emulsifiers to stabilize the newly formed interfaces e.g. surfactants, which affect surface chemistry of the solution[x]. If no chemical stabilizers are added, the newly formed emulsoids coalesce once agitation is halted. Having too much oil in an emulsion, can cause thermodynamic instability and cause an emulsion to 'break', forming free oil, which once again would foul a membrane. No emulsion has a zero fouling potential for a membrane, even if stable, but certainly a stable emulsion is less fouling than less stable counterparts.

In the case of produced water, the process chemistry is generally complex and often unique to each well, as the naturally occurring minerals and hydrocarbons as well as the fracturing fluid, cleaning, flushing and de-scaling chemical regimes differ by well. Since major salt water disposal (SWD) units, like McGrath SWD, generally take varying amounts of produced water from its various supply wells and mixes them, the untreated produced water essentially is of dynamic composition. Furthermore, many SWDs, like McGrath, add scale-inhibitor and a variety of other chemicals to the influent to adjust the process chemistry further. Temperature changes and emulsifier dosing during filtration, could thus cause unexpected changes in the produced water chemistry. Effects of temperature and a range of emulsifiers would need to be studied experimentally and statistically on any SWD due to the great variations in process chemistry. Hence, due to complex process chemistry, the oil-water emulsion may not behave exactly as water and exhibit a viscosity decrease with an increase in temperature, which would be advantageous for a membrane filtration process.

An important aspect of produced water chemistry is that it contains suspended solids and oils in solution. Complexes of these two materials may in fact be a large cause of fouling in produced water filtration, as these oils will most likely behave like a freely transferable coating when these complexes reach the membrane surface. In fact the AKZO Macro Porous Polymer (MPP) Extraction (MPPE) process[xi] employs the ability of polymeric structures, similar to microfiltration membranes, to immobilize oil-solids complexes in order to intimately contact the oil with an immobilized liquid extraction phase contained within the MPP structure. These complexes could thus transfer their oily coating to a membrane when in contact with the membrane and coat and foul its active surface. Numerous studies[xii] have investigated means of removing oil from oil-mineral and oil-organic complexes and numerous commercial surfactants used in the oil and gas industry aim at doing just this e.g. treating oil-sand complexes in offshore drilling applications. Clearly these complexes represent a fouling issue for membrane processes. It is unclear if the oil coating can be removed from these oil-solids complexes by forming a stable emulsion around it, but it seems reasonable, assuming that the complexation is reversible.

The crude oil saturation concentration in water is ca. 4 mg/l at ambient conditions with low water salinity[xiii]. Oil solubility decreases as salt content and temperature rises, but increases as the oil-to-water ratio increases[xiv]. The soluble oil concentration represents the lowest permeate oil concentration achievable by an ultrafiltration (UF) based de-oiling process. UF was shown to decrease total oil and grease (TOG) to 2 mg/l in a European study operated at 60°C[xv] and so clearly low permeate TOG is possible. The chemistry of oil-water solutions is not simple and, as such, one could have a case where the soluble oil content can be reduced chemically and moved into the emulsion fraction, and that may lower the achievable concentration. Current U.S. EPA on-shore regulations for regions West of the 98th meridian, require a peak of 35 mg/l TOG concentration for disposal into navigable waters used for agricultural and wildlife applications. This rule is subject to the water being of quality for direct use in agriculture and wildlife applications in times of drought. Regions East of the 98th meridian are not allowed to discharge produced water into navigable waters because these produced water sources are often seen as more contaminating. These regulations may tighten, e.g. the E.U. standard requires < 5 mg/l total hydrocarbons for onshore disposal[xv]. There are generally no regulations for disposal into non-surface waters, such as into disposal wells, and the water quality is generally governed by the well characteristics, with the aim of maximizing well life. Compliance with current on-shore disposal regulations can in many cases be met with standard wound filters, mainly because the relevant produced waters are typically clean.

Wells with higher TOG and solids content in their produced waters, e.g. McGrath, generally find it challenging to generate water quality which can be used for agricultural and wildlife purposes and prefer the down-hole disposal route. Even for down-hole purposes, McGrath uses 1-micron pore-size cartridge filters. However, 1 micron is generally the tightest wound filters available and these are clearly not tight enough to present a perfect barrier to most process emulsoids, which have nominally < 1 µm size. In this case, filtration of the emulsoids is solely by cake filtration as a layer of rejected materials formed on the cartridge filter. It is important to note that cake filtration is sensitive to process upsets and filter hydrodynamics and may not be a reliable method of oil re-injection control. Even for cleaner produced water sites that do use the surface water disposal route, tighter regulations may require tighter membrane filters, such as UF filters or alternative separation processes. Membrane ultrafilters have pore sizes < 10 nm and can present near-perfect barriers to emulsoids, with the only process ramification being that the oil be maintained in emulsion form to attain effective separation. In swapping from wound to membrane filters, Western SWDs with more fouling TSS and TOG can possibly meet the 35-mg/l TOG surface water disposal limit and obtain large disposal savings by offering their water for agricultural and wildlife use. Constructing, operating and maintaining a disposal well is expensive and energy intensive. Most energy consumption stems from the operation of injection pumps.

In general, free oil is most effectively removed by floatation and overflow methods, such as the standard American Petroleum Industry (API) gravity settler, prior to any membrane separation process. This practice is employed at McGrath SWD. The soluble and emulsified oil can then be treated by membrane filtration to a nominally best performance of 4 mg/l of oil in the filtrate or permeate stream - if the membrane filtration is performed in the stable emulsion regime.

Experimental

Test objectives

The overall objective of the program was to evaluate alternative usage of produced water, rather than down-hole well disposal, specifically this project looks at reuse in electric power generation plants, after a reverse osmosis step. The technical objective of this experimental work was to evaluate CeraMem® ceramic membranes for process flux and permeate quality on land-based produced water as the feed stream. Membranes are designed to remove emulsified oil and particulates. Soluble oils, surfactants, and salts will pass through the membrane.

Membrane Types:

CeraMem fabricated three different membrane types for evaluation. These included a nominal 5 nm pore size silica and a nominal 10 nm titania ultrafiltration membrane, as well as a Teflon® coated membrane.

The lab-scale membrane elements had 1.5 ft^2 of membrane area which was located on 60, parallel, 2-mm square channels which run from the inlet end face to the outlet end face. Channel length was 12 inches and the ceramic element was 1 inch in diameter. The elements were installed in stainless steel housings with EPDM o-rings and gaskets.

Description of Test Apparatus

Test System

CeraMem's test system is a stainless steel, lab scale unit sized to work with 30 cm long, 1" OD membranes that were fabricated by CeraMem. The membrane has 2 mm channels and an area of ca. 0.13 m^2. The system is capable of simultaneously testing two membranes-in-series, in cross-flow conditions up to 45 psi trans-membrane pressure (on the first membrane in series) and 15 ft/s cross-flow velocity. The system was automated so as to operate safely in an unmanned manner to maximize run time for the field tests. Photographs of the test system are shown in Figs. C-1 through C-3 below.

Figure C-1 shows a front view of the test system with the cart on which the system is mounted as well as the NEMA 4 electrical enclosure for the power, monitoring and control systems on the right-hand-side of the of the cart. Fig. C-1 clearly identifies the membrane locations and the ca. 3 gallon feed tank as well as the actual slipstream feed point for the test system. The pipe on the left of the picture, entering the feed tank from the top (slipstream transfer line) as shown in the picture, is a 30 psi water supply that feeds the system through a valve, which is controlled by a level controller in the feed tank. This hose was replaced by a 500 psi chemical hose for the process tests and a needle valve was added to the feed line to control the maximum feed flow to the test system.

Figure C-2 focuses on the electrical components of the test system and shows the pump location, the pump starter as well as the control, data logging and power supply unit which was enclosed in a water proof (NEMS 4) unit. The permeate disposal line was run directly into the waste tank (mud pit) at the test site and was at atmospheric pressure. A hose, not the transparent hose shown in the picture, was employed at the test site.

Figure C-3 shows a slightly magnified view of the feed piping and the product sample ports for the test system, as well as safety features built into the system.

Figure C-4 shows a process and instrumentation diagram (P&ID) of the test system and how it interfaced with the test site. The most important operation features shown in Figure C-4 include the back-flush tank (T-2) which is used to automatically flush its product content (500 ml maximum) back through the permeate line for automated flux maintenance. Secondly, the back-flush line, when the back-flush pressure is set to zero, also serves as an automated relaxation system. Relaxation is operating with cross-flow of the feed, but with the product lines closed. This enables the build-up at the filter surface to be removed by the cross-flow effect while there is no tangential flow through the filter.

The test system was designed to operate under controlled permeate flow, with adjustable trans-membrane pressure (TMP). Permeate flow was controlled by differential pressure flow controllers (Kates' Flow Controllers) and the TMP was logged from the permeate-side pressure transmitters (Cole-Parmer Instrument Co) via a National Instruments data logger. Permeate flow rate was determined by turbine-type flow transmitters, FT1 and FT2 (Titan Flow Meters, UK), while the feed flow rate was measured with a Great Plains Industries turbine flow indicatortransmitter. Back-flushing and relaxation cycles were controlled by an Omron relay-timer switch which opened and closed SV2, SV3, SV4 and SV5 at pre-set cycles.

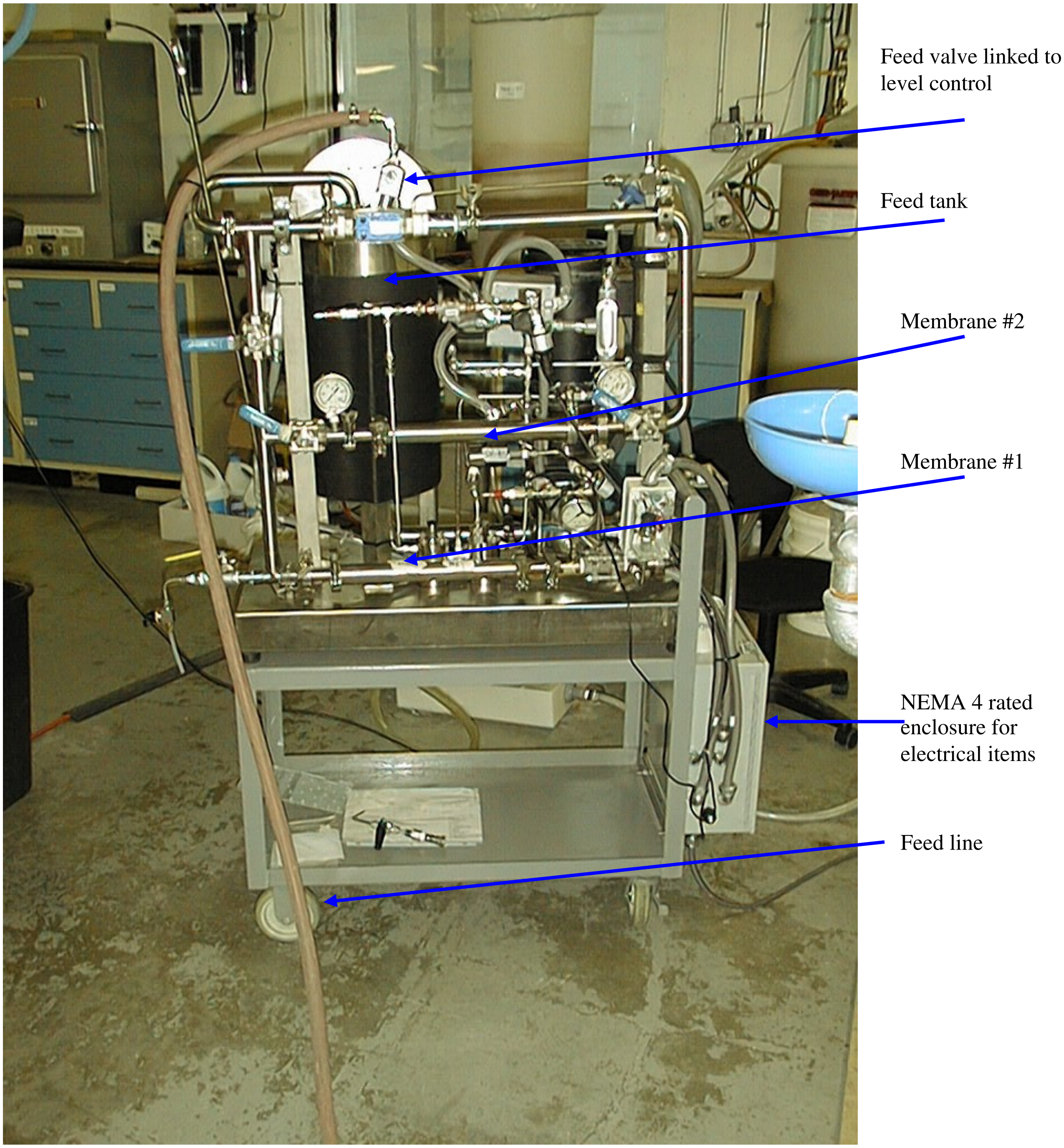


Figure C-1
Front view of the full height of the test system

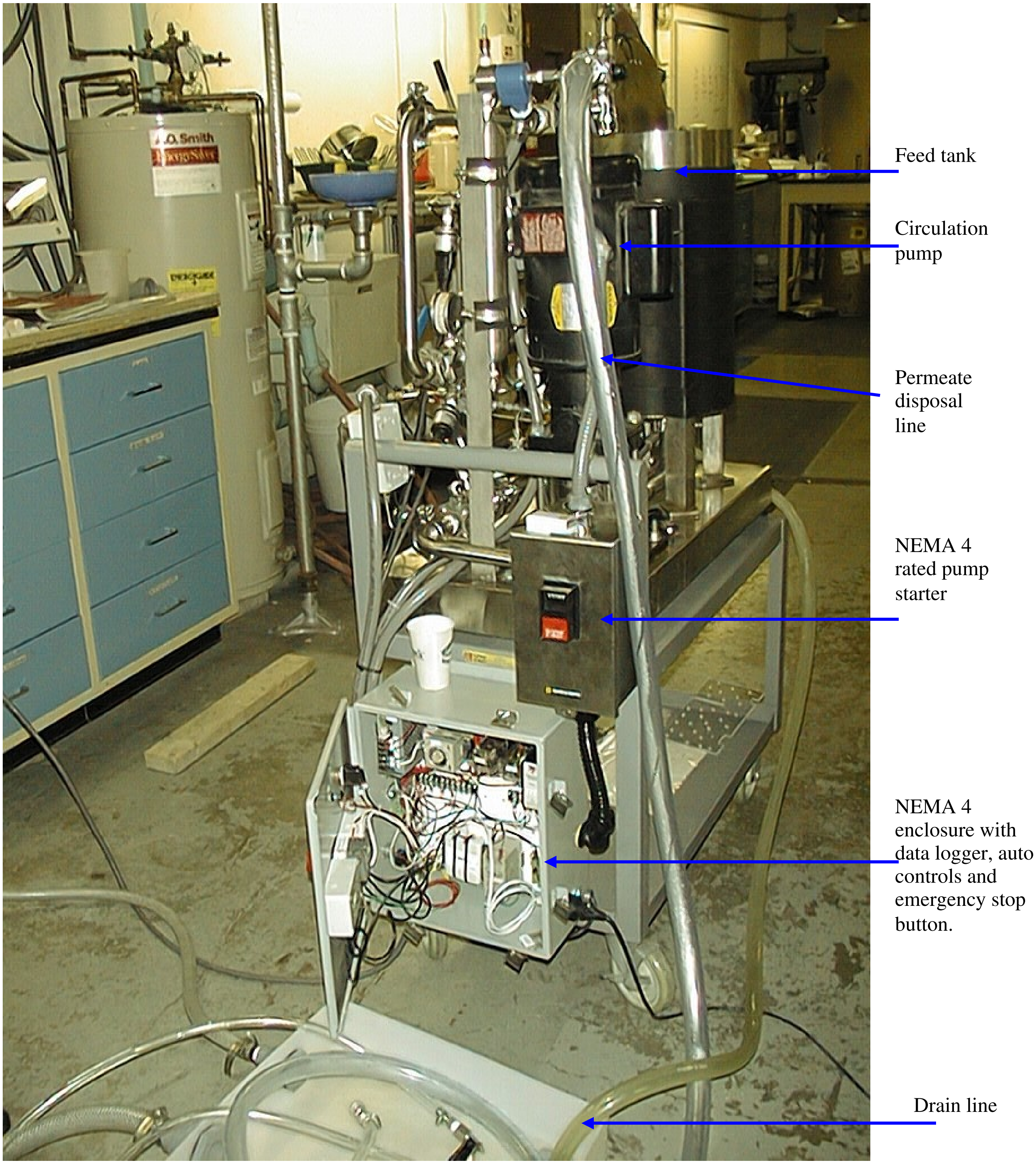
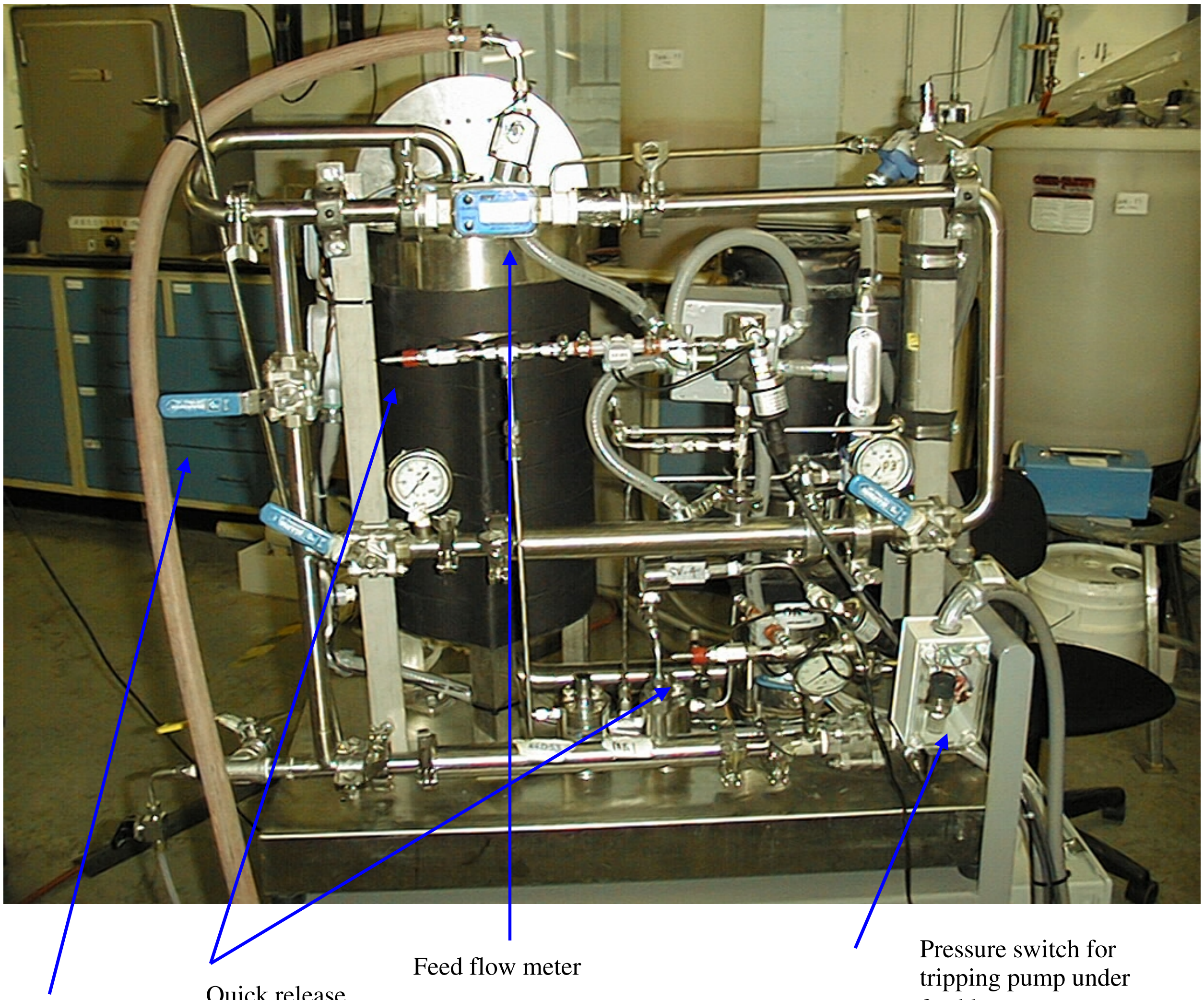


Figure C-2
Side view of the test system



Feed flow meter

Pressure switch for
tripping pump under
feed loss

Quick release
sample ports

Bypass valve for
Membrane #2

Figure C-3
Front view of the test system, focusing on the membrane process.

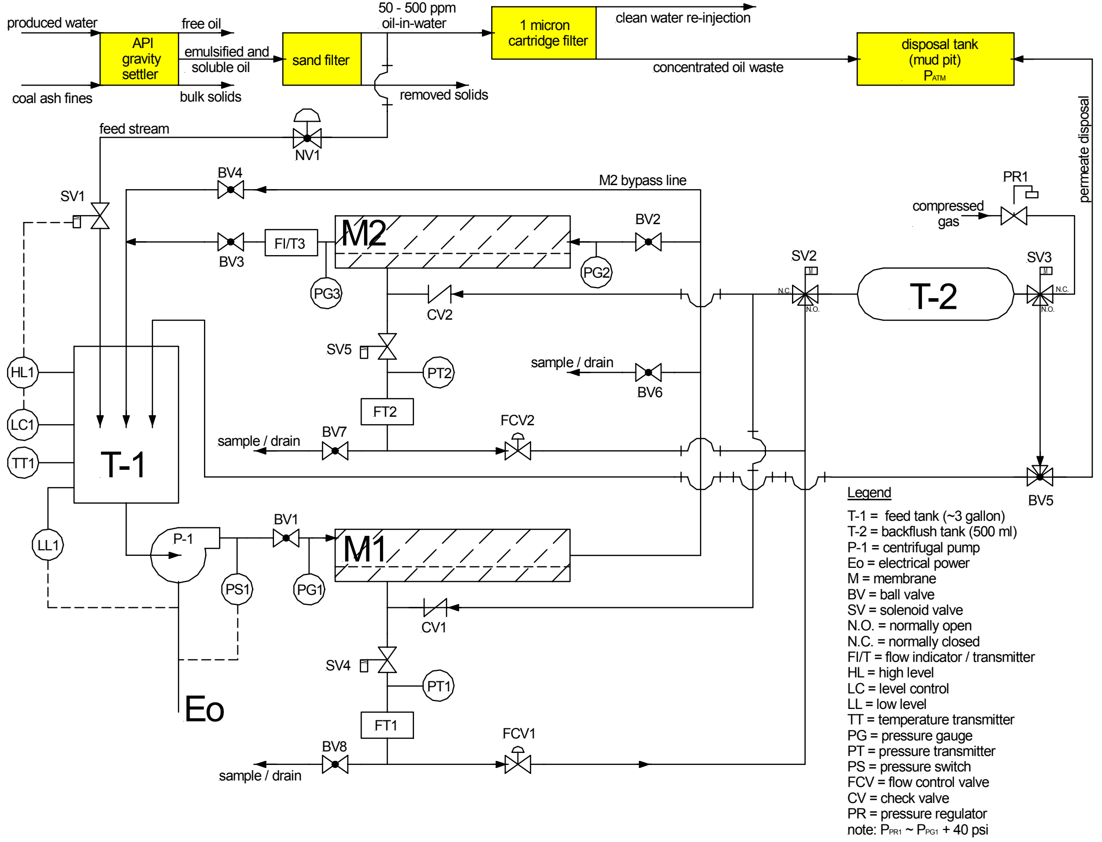


Figure C-4
P&ID of the test system highlighting interfacing with the test site

Tests performed

Two main types of tests were performed: feed batch concentration (BCT) and constant-feed concentration tests (differential tests). For all tests, an existing oil/water gravity separator and sand filters at the test site were employed prior to the membrane test system to reduce the free oil and solids concentrations prior to entering the membrane system (Fig.C-4). The advantage of this process arrangement was that the membrane system concentrate could be returned to the oil/water separator, where free oil could be skimmed off continuously and the soluble oil stream then taken back to the membrane system for further processing. This greatly simplifies the overall produced water treatment process in that the concentrated waste to be disposed of is greatly diminished, and the free oil product can be exploited. In the field tests performed, no recycle to the free oil separator (API) was performed.

For the performed cross-flow filtration studies, variables investigated and their quantitative ranges are shown in Table C-1. Membrane chemical cleaning was performed manually, both inside the test system as well as by soaking membranes in cleaning solutions, which included detergents (surfactants), acids, and bleach. Chemical cleaning was an important process aspect to investigate, as knowing frequency and cost of materials for cleaning is useful to determine process operating costs.

Table C-1
Process Variables for Cross-flow UF

Variable	Range
Cross-flow Velocity	9 to 16 ft/s
Trans-membrane Pressure	15 to 60 psi
Process Temperature	Ambient to 75°C

Permeate quality was assessed largely by turbidity measurements as well as some oil and grease (EPA 1664) and total suspended solids (EPA 160.2) measurements. Feed and concentrate samples were analyzed at various times throughout the test period.

Field test site

All tests were performed at Burlington Resources' McGrath SWD site from 22 through 30 June 2005. Burlington manages in excess of 1,000 gas wells in the San Juan Basin region of North Western New Mexico and many of these are coal bed methane (CBM) wells, which produce a fairly clean produced water, largely free from hydrocarbons with some coal dust present. Burlington's other wells are on oil-bearing formations and so produced waters from these wells tend to contain substantial amounts of hydrocarbons and less coal dust fines. In many wells, the produced waters may be a combination of CBM and hydrocarbon-rich formation waters. CBM waters are generally processed and injected on unmanned SWDs as these waters often require less stringent filtration and filters have a longer lifespan than that of filters treating oilier produced waters. McGrath SWD is Burlington's largest manned disposal facility and is specifically assigned the worst produced waters and the main focus of the staff is to continuously monitor filter performance and replace spent filters.

Currently McGrath SWD uses the process layout shown in Fig. C-4 (top section of the figure) for produced water treatment prior to re-injection – just before the 1 µm cartridge (wound) filters, there are 5 µm cartridge filters. This process essentially comprises an API gravity separator, followed by a sand filter and then cartridge filters, followed by re-injection. According to a contracted (Sierra Chemicals) dosing chemist who attends to McGrath (and a number of other Burlington and non-Burlington SWDs), the produced waters entering McGrath are some of the worst (highest solids and oil content) in the San Juan Basin. He said the flows through some of the CBM SWDs are similar to McGrath at ca. 6,000 bbl/day and are much cleaner (lower solids and oil content).

Test Results and Discussion

Seven successive test runs were performed from the 22nd through 30th June 2005. The total run time for these tests was 122 hours of which more than 80% comprised batch concentration tests, clearly pointing to the exploratory nature of the study. Table C-2 shows the feed-side operating conditions. Refer to the Appendix for a summary the entire field test data set. All tests were performed at constant permeate flux, controlled by mechanical flow control valves. Recorded pressure variations are thus the process performance indicator. However, the minimum pressure differential for the flow controllers to work was 10 psid and hence the loss of flow control at low permeate pressure was observed for some tests. In these cases, the process was changed to constant TMP, variable flux mode.

Table C-2
Operating conditions for each of the 7 test runs (all at 65°C)

Run #	Run time	$P_{feed,M1}$	$P_{feed,M2}$	Xflow	Maintenance	Add Chemicals
1	7.2 hrs.	46.3 psig	39 psig	3.0 m/s	None	No
2	9.2 hrs.	56 psig	45.5 psig	3.1 m/s	None	No
3	21.5 hrs.	48.5 psig	26.5 psig	5.1 m/s	Back-flush	Acid
3b	22 hrs.	47 psig	23 psig	4.8 m/s	Back-flush	No
4	23.3 hrs.	46.5 psig	24 psig	4.8 m/s	Back-flush	No
5	10.1 hrs.	47 psig	25 psig	4.9 m/s	Back-flush	No
6	24 hrs.	45.5 psig	24 psig	4.9 m/s	Relaxation	Acid + Surfactant

Run 1

Run 1 was performed at a moderate cross-flow velocity (3 m/s; 9.85 fps) as a test to see if the process could operate reliably at moderate shear and lower energy input. Figure C-5 shows the flux and TMP plot at various times over the run.

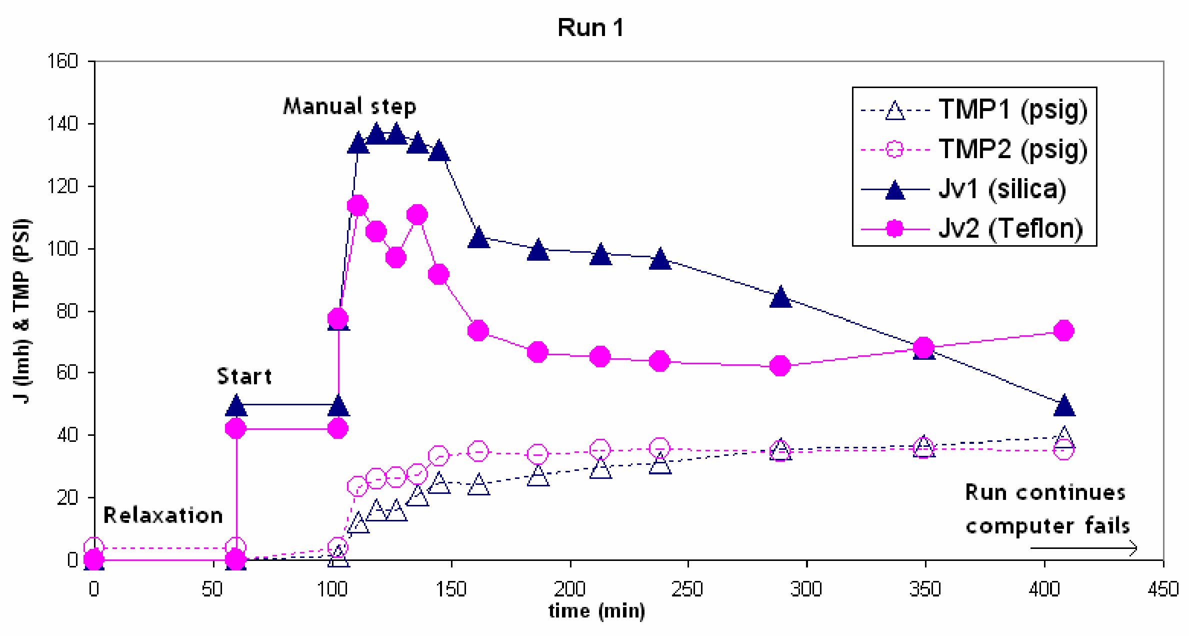


Figure C-5
Flux and TMP relationship over time for Run 1

The first hour of Run 1 simply re-circulated the feed without permeate production – relaxation – in order to warm the feed fluid up to the equilibrium circulation temperature. No heat control was installed in the test system and hence the feed heated up to the equilibrium temperature, which was a balance of heat due to friction from the cross-flow velocity and the heat loss through the stainless steel tube and tank walls. For all the tests, the equilibrium temperature (at relaxation) varied between 60 and 70°C and once permeation was started, the rate of permeation adjusted the temperature accordingly, but anywhere from 50 to 60°C was observed as an equilibrium batch concentration temperature. For pure water, the viscosity and hence membrane hydraulic resistance effect will be reduced as temperature increases. This will lead to increased filtration flux. Produced water, as mentioned before, exhibits complex chemistry and may not behave exactly like water. Future studies should elucidate the effect of lower temperature on the filtration operation for comparison to the present results. Full-scale processes in feed-and-bleed, stages-in-series operation may not heat up to 60°C and a temperature of 40-50°C may be more realistic. Future tests should perhaps consider temperature control. Emulsions are generally also de-stabilized by higher temperature.

Run 1 was isothermal and some interesting trends are shown in Fig. C-5. The batch concentration process flux was started at t = 60 min, with fluxes on each membrane at approximately 45 liters, per meter squared of membrane surface area, per hour (lmh). At these fluxes, the TMP remained fairly low and steady and for 40 min., indicating a relative stable operation. At 100 min into the test, the flux rate was manually increased by a factor of 2.5-to-3, which was the last operator-applied process change for this test. At that point, the flux for membrane 1 (M1) was then manually set at 134 lmh and that for membrane 2 (M2 – Teflon®) at 114 lmh at t = 110 min. The TMP increased to overcome hydraulic resistance at the higher permeation rate (scales with membrane pure water permeability) and concentration polarization (CP) or boundary layer (BL) thickness increased as a result. The TMP adjustment from CP and hydraulic resistance changes essentially occurs instantaneously. Membrane fouling, which includes membrane pore blockage and active surface blinding, is also increased with flux rate increases, but this is a slower, continuous process. Fouling continually increases filtration resistance and so continuously causes a demand for a higher TMP to maintain constant flux, until in-line process maintenance or cleaning is employed to restore membrane activity. Slower fouling is of course desired to reduce the need for membrane cleaning, which reduces process operational costs.

Twenty minutes after the high fluxes were set, the membranes reached their maximum TMP due to fouling rate increases and so flow control could no longer maintain flux. The flux for each membrane subsequently fell by 35 lmh at t = 162 min, only 50 minutes after stepping up the flux – giving a 25-30% drop in performance due to fouling in less than one hour. This is severe fouling and can be related in part to the high process flux. However, a second source of fouling is a build-up of free oil passed through the pre-treatment system into an unstable emulsion, due to the lack of emulsifying agent. The feed to the test system comes from an API and gravity settling tanks and so it would most likely contain demulsifiers (to improve API and settling performance), rather than emulsifier, which would not stabilize any free oil. Due to the turbulence in the small feed tank volume and short residence time, free oil could not float up (no oil layer was seen in most tests) and so it is pulled into the pump inlet and gets mechanically emulsified into an unstable emulsion (no emulsifiers), which fouls the membrane. In some batch concentration tests performed in this work, the membranes fouled severely at the outset and then showed signs of performance recovery with time. However, the fouling effects from start-up often lingered and suppressed membrane performance throughout that run and only chemical cleaning, not back-flush, could recover the flux.

Looking at the flux for M2, the Teflon® membrane in Fig. C-5, it is clear that once flux was stepped up at t ~ 100 min, the unstable emulsoid was brought into intimate contact with the membrane and blinded the membrane, probably in conjunction with minerals in the feed stream. Later, as the emulsion became more concentrated at t ~ 300 min, the performance on M2 improved. This clearly meant that the fouling layer became less resistant as no process alterations were made at that point. A feed change could be responsible, but this is unlikely since the same quiescent clarifier tank system was used throughout the run. M2 was CeraMem's non-standard Teflon® coated membrane, which most likely exhibited lower surface energy and a higher tendency to release foulants than M1, which was CeraMem's standard nominally 5nm pore size silica membrane. This could be linked to the recovery observed in M2 and not M1 (silica membrane). Run 1 ran longer than the time that the data logger recorded due to computer failure, but after ca. 10 hours of batch concentrating, it was observed that both M1 and M2 exhibited low flux. It is suspected that the increased concentration of mechanically (poorly) emulsified free oil in the retentate led to excessive membrane fouling. Poorly emulsified oil combined with the solids present in the retentate, as mentioned earlier, would have severely blinded the membranes.

A mixed permeate (from the combined permeate pipe) total oil and grease (TOG) sample was taken at t ~ 60 min. and 26.3 mg/l of oil was detected. This TOG value was below on-shore regulatory limits, but was larger than a typical saturated soluble oil concentration of 4 mg/l. This is a strong indication that a non-stable emulsion may have been present at this point in the system. A mixed permeate turbidity sample (t ~ 60 min.) of 12.6 NTU was relatively poor, once again pointing to emulsoids or droplet of oil present in permeate. Turbidity is generally indicative of suspended particulate or colloidal matter. A typical feed stream to a reverse osmosis (RO) plant would have a turbidity of < 1 NTU and so the product from Run 1 was not at that standard.

Run 2

Run 2 aimed to duplicate Run 1: no back-flushing or chemical process adjustments and the use of a moderate cross-flow velocity (3 m/s). Like Run 1, the start-up procedure also included circulating the oil-water solution to allow the temperature to increase and apply shear stress to the solution, in order to attempt to stabilize the emulsion. As ahown in Fig C-6, flux was started at ca. 50 lmh for each membrane and this seemed to be the maximum flux that M2 (silica membrane) could produce. Flux from M2 could not be increased after about seven min. of operation when membrane flux was increased manually and slowly over a period of about one hour, until t ~ 75 min. TMP for M2 seemed to reach its maximum at t ~ 32 min. and so fouling was very fast for this membrane and this must be attributed to the inlet stream or feed condition at the test site, as well as operating conditions.

The inlet TOG and TSS concentrations were variable at McGrath and four samples of the feed stream taken during the entire test period were analyzed for TOG and TSS. Average and standard deviation values were: TOG = 57.5 mg/l ± 102.4% and TSS = 127.4 mg/l ± 50.4%. The feed stream sample taken with Run 2 had a TSS = 133 mg/l and TOG = 22.8 mg/l. This

TSS is at the average of the variable feed TSS range, while the TOG was 40% of the average value. Run 5 had the highest feed TOG sample of 159 mg/l and both membranes showed signs of fouling at start-up, but it was not irrecoverable, as was the case of M2 for Run 2. It is clear that in order for a larger amount of oil to pass through the API in Run 5, there was either more free-oil passing through the API (unlikely because demulsifiers were added to the API) or there was more surfactant present in the produced water in order to produce a larger TOG emulsion that entered the filtration stage. For this reason, one can deduce that not only was TSS and TOG changing, but that process chemistry was also changing across runs. Clearly process control on an envisaged membrane process for treating the produced water will need to use on-line TOG monitoring, such as an infrared method.

At t = 75 min., M1 exhibited a flux of 90 lmh and M2 a flux of 40 lmh, but M2 showed signs of excessive fouling and rapid flux decline at this point. No more operational adjustments occurred after t = 75 min.

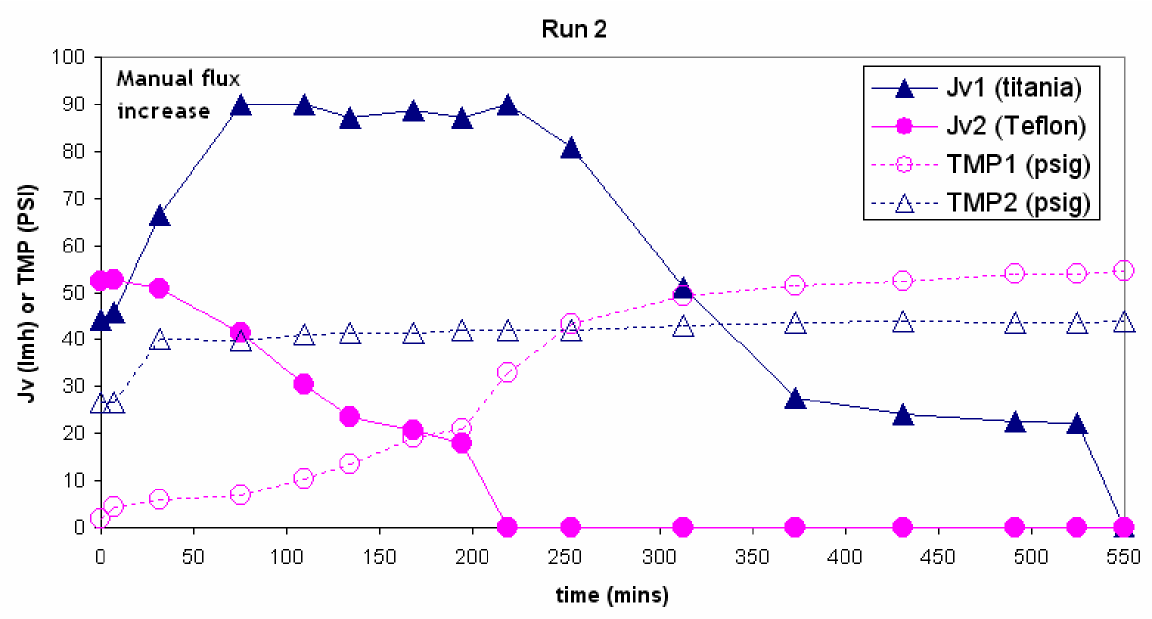


Figure C-6
Flux and TMP relationship over time for Run 2

From t = 75 to 219, M2 showed a steady flux decline to zero, while M1 maintained flux at about 90 lmh, its setpoint, but showed a steady TMP increase. Fouling was severe during this 2.5 hour period until t = 219 min. To develop an idea of the fouling potential of the retentate stream, its oil and solids concentration need to be estimated at this point. Based on a dead volume of 6 liters in the test system and the average flux of the membranes, it was determined that at t = 219 min, the volume concentration factor (VCF) of the batch process was 17. Note that concentration factor (CF) can be determined as: $CF = C/C_o \sim 1 + VCF$, if one assumes that rejection of oil and solids is high.

At the end of Run 2, a mixed permeate (M1 + M2) sample was analyzed and had a TOG = 40.7 mg/l (just above the maximum onshore regulatory peak level of 35 mg/l). Assuming, that the average permeate TOG = $(TOG_{feed} + TOG_{permeate,end})/2 = (22.8 + 40.7)/2 = 32$ mg/l, and that the feed TOG was constant at 22.8 mg/l, the retentate is determined to have TOG = 179 mg/l at VCF = 17. Retentate TSS could not be estimated as no permeate TSS values were known. However, the retentate TSS at the end of Run 2 was analytically determined as 587 mg/l and so the retentate TSS at t = 219 min. was between that of the feed and final concentrate levels, i.e. between 133 and 587 mg/l. Clearly a TOG = 179 mg/l was larger than the soluble limit and so an unstable mechanical emulsion, coupled with a presence of TSS meant that the entire 219 min. period from start-up was one where fouling potential for the process was large. The Teflon® membrane (M2) experienced more severe fouling than the titania membrane (M1). Titania, like silica (metal oxides in general) is hydrophilic and oleophobic whereas Teflon® is generally hydrophobic and oleophobic. CeraMem's Teflon® membrane comprised an outer structure of Teflon® cast over a metal oxide undercoat. The Teflon® membrane may thus not behave as pure Teflon®, but as a hybrid material with properties leaning strongly towards that of Teflon®. Hence understanding the fouling properties of the feed-membrane combination is not simple.

During the entire period from start-up to t = 450 min., the McGrath cartridge and sand filters experienced a severe loss of performance. The McGrath process treated the same feed stream as CeraMem's membranes in a similar batch concentration mode, so a direct comparison could be drawn. The operators stated that on seemingly random days, a layer or cloud of unsettled oil and/or dirt would arrive at the suction inlet of the unstirred settling tanks that hold the filtration feed at McGrath SWD, and adversely affect filtration. In this particular case, after the cloud passed through the McGrath filters, the filters experienced a low flow alarm and the control system automatically cut the feed pump and halted filtration at a corresponding Run 2 time of t = 450 min. The source of the 'cloud' needs to be elucidated in future work in order to avoid it. CeraMem's ceramic membrane M1 (silica membrane) managed to continue the batch concentration process through this highly fouling event, which was indicative of improved fouling management of a cross-flow membrane filtration process, using ceramic membranes of often lower fouling propensity.

At the end of the run a concentrate TSS and TOG sample was taken, along with a mixed permeate TOG and turbidity. These results are shown in Table 3 below along with the feed properties associated with Run 2. A summary of all results is shown in the Appendix.

Table C-3
Analytical results at the end of Run 2.

Stream	**TOG (mg/l)**	**TSS (mg/l)**	**pH**
Feed	22.8	133	6.84
Concentrate	294	587	7.57
Permeate	40.7	Turbidity = 0.6	7.57

It is important to note that the turbidity of the permeate product was low - a typical RO feed will demand turbidity < 1. This indicates that a low proportion of particulate solids and colloidal matter was present in the permeate, which is typical for UF. However, at the end of Run 2, the VCF ~ 25 and so the ratio of retentate-to-feed TSS of 4.4 is much lower than one expects when the permeate was essentially solids free. This result can partly be explained by the fact that the feed stream was highly variable and so perhaps 133 mg/l was at the high end of the spectrum.

Furthermore, since membrane fouling was severe, some solids were of course contained in the fouling layer of the membranes. Also, settled solids could not easily be stirred into the concentrate and included into the retentate TSS analysis. When the test system feed tank was drained, there was a layer of solid deposit observed on the tank base. Future work should investigate continuous solids deposit removal from the feed loop.

Both in Run 1 and 2, no back-flushing was employed and the cross-flow velocity was moderate. The flux decline observed throughout these two batch concentration runs was fairly strongly linked to the start-up fouling of the runs. The performance of the membranes was much worse than would be the case if tests employed emulsification of the oil concentrate. However, if one could operate without dosing of emulsifiers e.g. use the constituents of the feed to generate a stable emulsion, the process would be cheaper both from cost of emulsifier perspective and for subsequent demulsification required as part of the full-scale feed-and-bleed system. Thus emulsifier dosing, which in itself requires some study, will be avoided until necessary.

Run 3

In Run 3 both back-flush and a greater cross-flow velocity were applied in order to improve process performance. These two operation modes both improve mass transfer of the membranes during operation. The back-flush duration was 10 sec. at 20 min intervals. Essentially the entire 500 ml back-flush tank contents would be blown back through the membranes every twenty minutes and this was accounted for in determining the net flux (total permeate flow rate less back-flushed permeate). Figure C-7 plots net flux versus time for Run 3. Run 3 was started up with ca. 50 lmh gross process flux for each membrane.

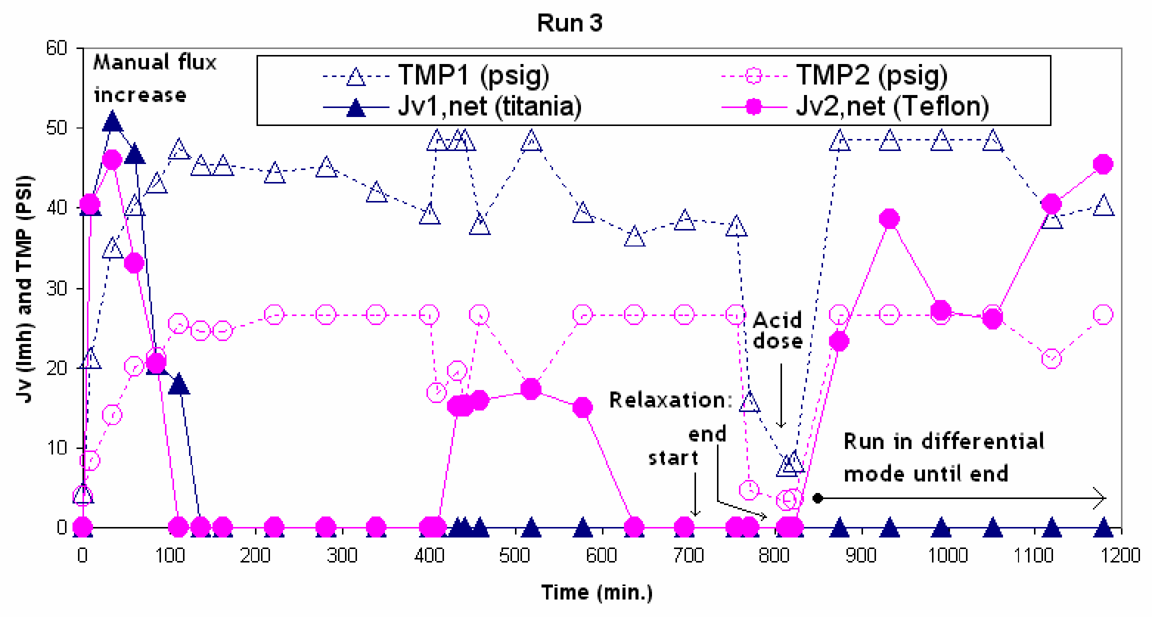


Figure C-7
Net flux and TMP relationship over time for Run 3

Flux was started at 40 lmh (net) and the process was then left undisturbed until t = 730 min. Essentially start-up fouling was excessive and the process flux dropped within two hours (t ~ 130 min.) to zero on both membranes. M2 (Teflon® membrane) flux recovered at 400 < t (min.) < 640, but then again dropped to zero. Flux from M1 (silica membrane) never recovered. At t = 730, the membrane permeation was stopped and both membranes were run in relaxation mode for 90 min. When flux was resumed in *differential* mode at t = 813 min, M2 showed some recovery, but M1 showed no recovery. At t = 820 min. ca. 10 ml of 66% H_2SO_4 was dosed into the feed tank and so the pH was dropped from the feed pH of 7.12 to a pH of 3 and this was maintained while the process was in differential mode. After acid dosing, M2 showed a more marked flux improvement over 820 < t (min.) < 1180, but M1 was still performing poorly showing zero flux. At t = 850 min, M1 was run in relaxation mode for ten more minutes and flux was resumed in *differential mode* at t = 860 min., but M1 produced a zero net flux until t = 1180 at which point Run 3 was stopped.

From about the start of Run 3, McGrath SWD started treating water that was significantly different to that for Runs 1 and 2 in that it had a much lower TOG and TSS. This water was comprised largely of the settling tank bottoms water (settled out waste) from one of Burlington Resources' CBM re-injection SWDs. Essentially the CBM SWDs, as mentioned before, operate in an unmanned manner because CBM water generally has very low TOG and TSS. The CBM re-injection process is simple: take produced water, put it into a settling tank, send the floating hydrocarbons to a storage tank, keep the settled solids in the bottom of the tank, and simply filter and re-inject the middle portion of the settling tank. At some point, the CBM settling tank is ridded of its tank bottoms and these waters with concentrated solids and low hydrocarbon content is what McGrath was treating during Run 3, Run 3b and Run 4. Of course McGrath mixes this water with other waters and so the water is always a mixture of different waters and never easily defined. At the end of Run 3, for comparison, the product from the McGrath cartridge filters exhibited a TSS = 83 mg/l and a TOG = 42 mg/l. This was fairly bad quality water by cartridge filter standards and was indicative of the high TSS of the produced water treated by the test system during Run 3. A concentrate sample, from the end of Run 3 had TSS = 283 mg/l and a TOG = 9.44 mg/l. Although Run 3 exhibited a low VCF due to fouling issues, this TOG value was very low and in fact this result contradicts a mixed permeate TOG = 64 mg/l (turbidity = 6.2 NTU) taken only three hours earlier. A mixed permeate sample from t = 100 min. also showed a TSS = 43.3 mg/l (turbidity = 2.61 NTU) and TOG = 48.3 mg/l. This confirmed that at the end of Run 3 permeate and hence concentrate TOG was in excess of 9.44 mg/l. Acid dosing between the TOG = 64 and TOG = 9.44 mg/l samples may have altered the efficiency of the hexane extraction method (EPA 1664) of TOG determination e.g. acid dosing could have altered water chemistry and produced non-hexane extractable oil-solids complexes to form. The concentrate TOG should have been > 64 mg/l, e.g., that of the mixed permeate sample.

The large cartridge filter product TSS = 83 mg/l was indicative of a large feed stream TSS, which was in keeping with the presence of CBM tank bottoms. The permeate analytical results discussed above indicated poor performance of the ceramic membrane filters and are indicative of relatively high feed TSS coupled with poor emulsion stability, a highly fouling combination, present in Run 3. Again influent variability on multi-influent disposal wells such as McGrath requires on-line TOG and possibly TSS monitoring and feedback to dosing control equipment for success for a full-scale membrane-based produced water treatment system.

Run 3b

Run 3b was a repeat of Run 3 except that no chemical (acid) dosing was employed in this case (M1 = silica, M2 = Teflon® membrane). Like Run 3, fouling was severe and immediate at start-up (Figure C-8). Run 3b was simply set up and left undisturbed from t = 0 to t = 875 min.

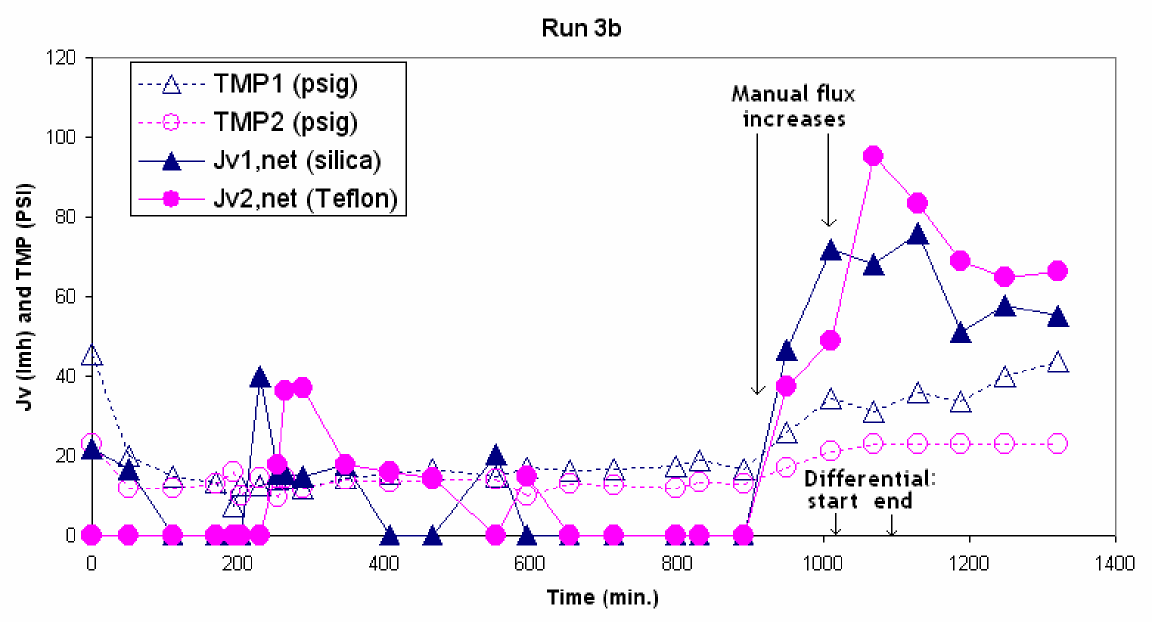


Figure C-8
Net flux and TMP relationship over time for Run 3b

For Run 3b and later runs, it seemed that the permeate-side pressure transmitters were not always reliable and acid damage (from dosing in Run 3) was suspected. It is however clear from the previous results that the system was normally running at maximum TMP in order to achieve flux control during the tests. It can be seen in Fig. C-8 that although the flux was not stable from t = 0 to 875 min., it certainly was not zero for M1 throughout this period, as was the case for Run 3. At t = 875 min, flux was manually increased and permeate turbidity measurements taken: turbidity M1 = 3.55 NTU and M2 = 5.87 NTU. These turbidities were indicative of a large solids and/or colloidal content in the permeate. Oil colloids in the permeate could have been present as a mixed permeate sample gave TOG = 41.1 mg/l. So again it seemed like when TOG was large in the permeate, permeate TSS was large as well, and so it clearly points to the presence of oil-coated solids and unstable emulsoids in the concentrate. This assertion was confirmed by the analysis of the concentrate at the end of Run 3b, which gave TSS = 950 mg/l and TOG = 78.4 mg/l. This TOG was certainly in a super-saturated solution range, without emulsifier added. The TSS was large and thus these yielded a highly fouling combination. It is interesting to compare these results with the feed stream (sand filter output), which had a TOG = 17 mg/l and TSS = 78 mg/l. This low TOG, which agrees with the higher fouling potential of the feed, was indicative of the presence of the CBM tank bottoms at McGrath SWD.

From t = 875 to 1000 min, membrane flux was manually adjusted a number of times to maximize it. For t = 1010 to 1095 min, the process was switched to differential operation mode and during that period, where the feed was no longer concentrated, the process flux remained either stable or improved on each membrane. At t = 1095 the batch concentration mode was resumed and the flux on membrane M1 was manually increased. From t = 1095 to the end of Run 3b, the process was left undisturbed, and once again, flux decline was obvious under the batch concentration mode over t = 1095 to 1321, when Run 3b ended. Permeate turbidities at the end of Run 3b were: M1 = 3.7 NTU and M2 = 2.3 NTU. This M2 turbidity value was 60% lower than its corresponding value from t = 875, while turbidity was essentially the same for M1.

Overall, Run 3b operated at lower flux for the first 14 hours compared to Run 3. When flux was increased after that start-up period, the membranes responded positively, unlike in Run 3, where even chemical dosing could not recover flux. This slow start-principle is important for preservation of membrane performance during start-up with an unstable emulsion, but may be less important if the emulsion was chemically stabilized.

Run 4

Run 4 was operated like Run 3b, but the back-flush frequency was doubled at the same net back-flush flow rate: 5 sec. back-flush every 10 min. Once again, no chemical dosing was employed and the process was simply started-up at ca. 50 lmh and left to run undisturbed from t = 0 to 881 min. Observation of the flux performance during this overnight run revealed that, as was the case in Run 3b, the flux dropped at essentially non-maximum TMP. This flux was often below the reliable range of the flow meters on the permeate lines (~15 lmh) and registered as zero permeate flow, when it was just low. The poor permeate flow control can be attributed to the fact that the air temperature dropped overnight, causing the concentrate and permeate temperatures to drop, causing permeate viscosity to increase. This increase in viscosity caused the flow control valves to reduce the permeate flow rate to keep a constant pressure drop across itself. This seemed to have worked in the favor of Run 3b and Run 4 (seemingly slightly cooler evenings) in that a relatively large VCF could be obtained at this lower flux without the membranes being completely fouled by the morning. So that in the morning, just after sunrise, when the permeate flow was manually increased, the membranes responded positively in these two runs. This was exactly what happened at t = 881 min. when permeate flow was manually increased for Run 4 (Figure C-9).

Clearly for Run 4, slow start-up was effective in attaining concentration with the unstable emulsion without completely losing performance of the membranes. At t = 881 min, permeate sample turbidities were: M1 (silica) = 17.2 NTU and M2 (Teflon®) = 17.8 NTU. These were poor and the product appeared visually murky, implying that colloidal oil was present. No analytical samples were collected to verify the cause of this large turbidity. At t = 881 min., assuming an average flux of 10 lmh for each membrane, the VCF (based on 6 liters dead volume) was ca. 6.4. Flux was then manually increased at various points from t = 881 to 916 min. Using the measured feed (sand filter output) TOG for Run 4, TOG = 30.3 mg/l, the concentrate TOG was estimated as approximately 223 mg/l. This was a reasonable value, given that the measured TOG at the end of Run 4 was 315 mg/l. A TOG = 223 mg/l was expected to be above the soluble level and so a fair amount of unstable mechanically emulsified oil was present in the retentate. This observation implied that the fouling propensity of the retentate was large. This was not shown to be the case as the flux was maintained for both membranes from t = 881 to 1,200 min. without any chemical augmentation or any other process changes. In the case of M2, the flux was maintained at a fairly large average value of 65 lmh (net), while batch concentrating for more than five more hours. Clearly these results showed that operating with an increasingly more concentrated TOG somehow improved the filtration process stability, as seen in the flux increase in both membranes over t = 881 to 1,115 min. However, as concentration continued, eventually fouling did have an effect and the fluxes went through a maximum at t = 1,115 min. and thereafter declined over the next two hours until at t = 1,248 min. M1 dropped to zero net flux, and at t = 1,317 min. the M2 flux was ca. 35 lmh (net).

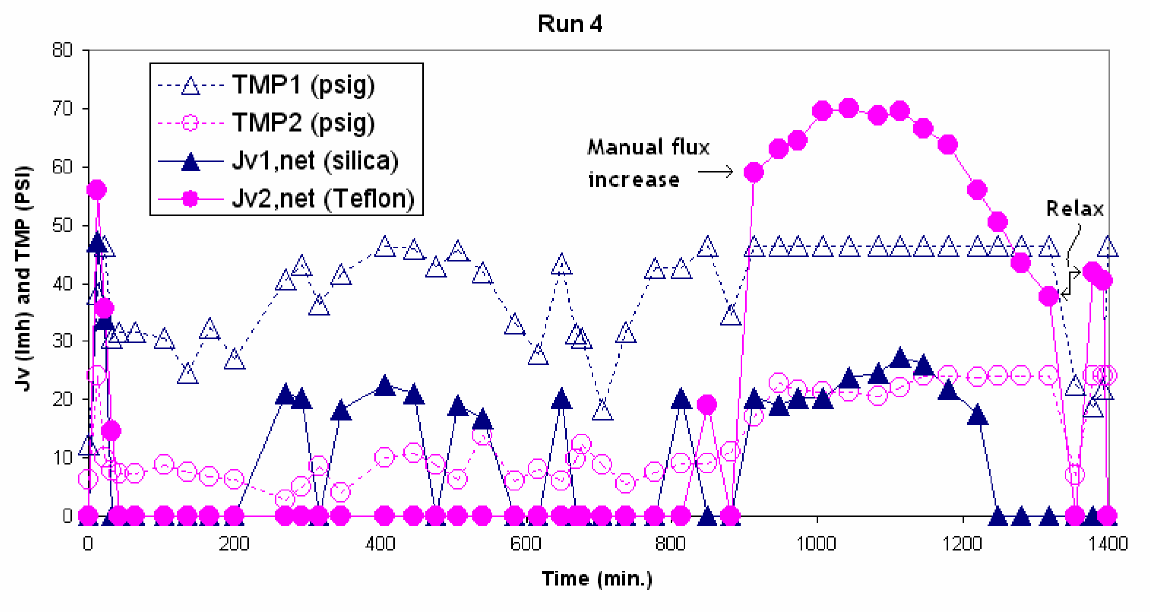


Figure C-9
Net flux and TMP relationship over time for Run 4

For t = 1,341 to 1,371 min. the system was set into relaxation mode, but once flux was started at 1,371 min., M2 showed flux recovery, but M1 was still at zero flux, until the end of Run 4 at t = 1397 min. Note that for Run 4, the feed stream (sand filter output) was still low in TOG (= 30.3 mg/l). The TSS = 66.7 mg/l was relatively low too.

Run 4 showed that low start-up flux enabled the VCF to build-up, which enabled concentration, without irreparable fouling damage to the membranes. Furthermore, once the higher concentrate TOG was attained, the membranes seemed to operate fairly stable, for at least five hours during subsequent batch concentration. This was unlike the quick fouling observation (M1) during the low TOG BCT operation of the start-up of Run 3.

Run 5

Run 5 was operated like Run 4 and Run 3b, but the back-flush frequency was doubled over that of Run 4 and quadrupled over that of Run 3b, at the same net back-flush flow rate: 2 sec. back-flush every 5 min. One distinction of Run 5 from runs before it was that the feed stream (sand filter output) exhibited an increased TSS and TOG at the start point: TOG = 159 mg/l and TSS = 230 mg/l. These were the largest TSS and TOG of the four feed samples taken during the 9-day field test and clearly indicated that the CBM tank bottoms were no longer present in McGrath SWD's tanks. As mentioned earlier, the higher feed TOG meant that more emulsifiers were present in the produced water and allowed a larger emulsified TOG portion to pass through the API. This emulsion, however, would only enable a limited amount of emulsification of incoming free oil to the test system feed tank. The mechanism of emulsification, under increasing oil-to-surfactant ratio is by increasing the average emulsoid size and maintaining a constant net interfacial area. However, as the VCF increased and the emulsoids grew to unstable sizes, these emulsoids would break into a stable and unstable portion. The unstable portion would again lead to membrane fouling. As a result Run 5 ran relatively poorly, as shown in Figure C-10. The permeate turbidities were measured just after start-up as M1 (silica membrane) = 1.33 NTU and M2 (Teflon® membrane) = 3.10 NTU. These values were still not within the desired range of < 1 NTU as a reverse osmosis feed, but the tests performed here were by no means optimized.

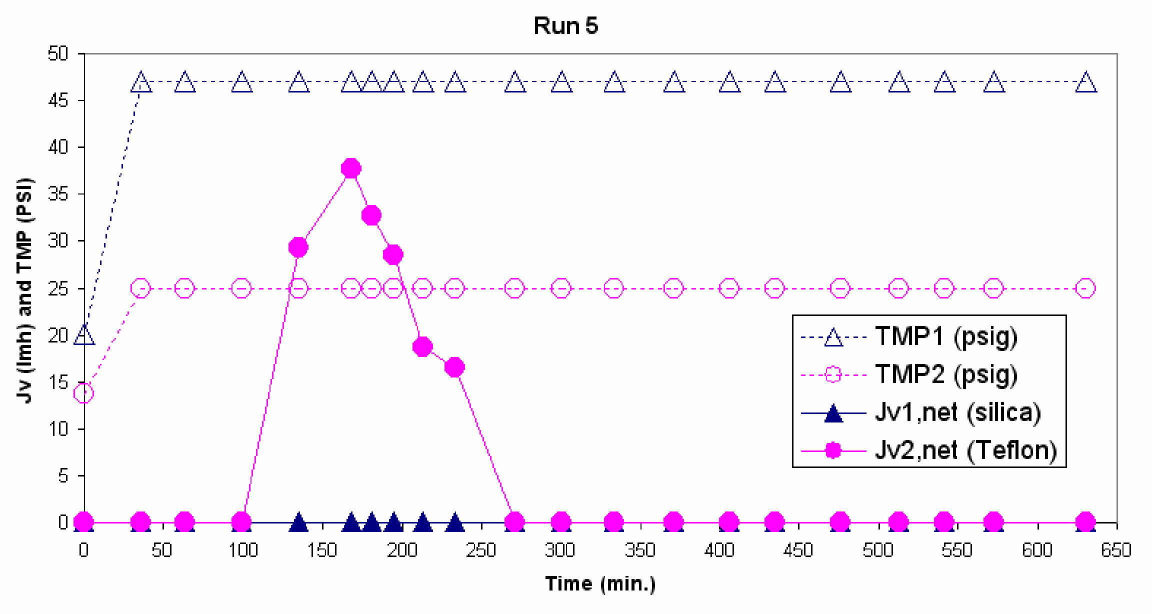


Figure C-10
Net flux and TMP relationship over time for Run 5

As mentioned before, flux was not zero at all the times when it was shown to be zero in Fig. C-10. The permeate flow was simply lower than the flow meter range at some instances where the flux was logged as zero. These low fluxes required for successful start-up were below the values predicted during the design phase of the field test equipment. As a result, the VCF calculated from the data seen in Fig. C-4 was perhaps under-estimated and gave a VCF = 1 at the end of Run 5. A more accurate VCF can be determined by simply taking the average concentration factors (CF) based on the concentrate TSS and TOG at the end of Run 5: TOG = 540 mg/l and TSS = 500 mg/l. The concentration factors were thus 3.4 based on TOG and 2.2 based on TSS, an average of 2.8 (VCF ~ 1.8), indicating that slow permeation was taking place throughout Run 5. Estimating a net flux of 15 lmh for M2 during the period t = 100 to 300 min. and using VCF = 1.8, it can be determined that the average net flux through the membranes throughout the rest of Run 5 was 2 lmh. This was much less than the minimum flow detection limit of ca. 15 – 20 lmh.

Runs 1 and 2 showed that fouling was severe and so in-line flux maintenance was employed for Runs 3 through 5. However, even with flux maintenance enabling longer runs, the natural feed chemistry did not lend itself to forming a self-supported stable emulsion and the need for chemical augmentation of the process became clear.

Run 6

The membranes were cleaned after Run 5, with both soap and acid solutions and rinsed with distilled water. However, M1 (silica membrane) still showed signs of residual fouling, while M2 (Teflon® membrane) was recovered, and so M1 was replaced with an unused membrane of the same characteristics and material (silica). For this final run, the process system was also mechanically altered to increase the maximum TMP by removing SV4 and SV5 (Figure C-4). This decision stemmed by the fact that low TMP range on especially M2 (< 25 psi) limited the test range. Removing SV4 and SV5 meant that the system back-flush mechanism was inactivated but could still be used for relaxation. The results from previous runs did not indicate that back-flushing significantly improved membrane operation, so this was not expected to cause a major loss of performance.

Run 6 was operated with a cross-flow of 4.6 m/s, similar to Runs 3 through 5, but on this occasion, the system was operated with automatic relaxation rather than back-flush: membranes being relaxed for 30 sec. every 10 min. Flux was started at ca. 50 lmh. The system was run in relaxation mode for ca. 30 min. and the equilibrium temperature was 64^{o}C just before flux was started at 50 lmh (Figure C-11). Note that as temperature dropped overnight, the flux rate will drop, hence to ensure a measurable overnight flux, the system was started at 50 lmh. The system was left undisturbed overnight until t = 670 min.

At start-up the permeate turbidities were M1 = 0.87 NTU and M2 = 0.72 NTU. These were relatively good and hinted at high TOG with emulsifiers (as per Run 5), which could, for the first few minutes, incorporate any free oil into the stable emulsion and so prevent colloidal oil from causing permeate turbidity. One weakness of Run 6 was that there was no analytical laboratory work performed on the feed and concentrate during this run and so this observation could not be confirmed.

Slow start-up enabled a slow BCT to take place overnight and M2 ran continuously overnight without seeing failure. From the flow rate data in Fig. C-11, at t = 670 min. the VCF was ca. 6.8 and the permeate turbidities were: M1 = 2.55 and M2 = 2.84 NTU. At t = 670 min. a mixed permeate sample showed TSS of 16.1 mg/l. This was the lower of the permeate TSS values measured in this program (the other from Run 3 was 43.3 mg/l).

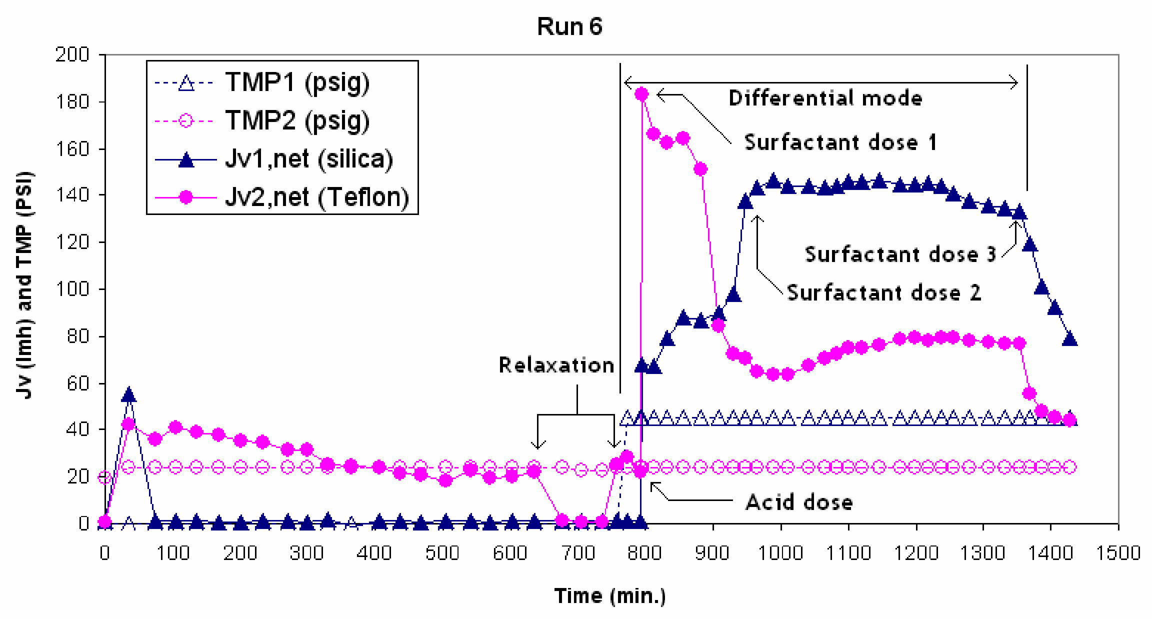


Figure C-11
Net flux and TMP relationship over time for Run 6

For t = 670 to 757 min. the process was put into relaxation mode and at t = 757 min., the process was started up in *differential mode* (all other conditions being the same, except that temperature increased by ca. 10°C) and for five minutes thereafter M2 was recovering to its pre-relaxation level, while M1 was not recovering at all. At t = 760 min. 5 ml of H_2SO_4 was dosed into the system. The process fluxes remained essentially the same, but TMP1 changed to its maximum value – acid cleaning may have been at work, but TMP measurements were unreliable and maximum TMP was assumed. At t = 773 min. ca. 10 ml of citric acid with surfactant (undisclosed surfactant and acid concentration supplied by Sierra Chemicals, operators of the McGrath SWD chemicals dosing team) was added into the 6 liter feed tank. Nothing noticeable changed by the acid dosing and flux of M1 was still poor and that of M2 fairly low. At this point the concentrate solution was acidic and since it was in differential mode, it remained acidic.

At t = 791 min. ca. 10 ml of Dawn dish detergent was dosed into the feed tank to improve emulsion stability. This was the first surfactant dosing of the field tests, and, as can be seen in Fig. C-11, the flux on both membranes suddenly increased at this point. The flux of M1 went from 1.4 to 68 lmh and that for M2 went from 22 to 183 lmh in less than one minute. Ten minutes later, at t = 801 min, the turbidity on M1 was reduced to 1.37 NTU and the flux remained large on both membranes. At t = 841 min, the system fluxes were still large and M2 had increased from 68 to ca. 85 lmh. At this point, the turbidities on both membranes were low: M1 = 0.58 NTU and M2 = 0.67 NTU. These turbidities were of quality that could possibly be used for reverse osmosis, if chemical analyses and SDI were acceptable. It was clear that dosing acid plus soap, improved process fluxes and turbidities. In general, dosing improved membrane performance.

The exact mechanism of this improvement needs to be determined, but it is clear that acid alone did not improve the process flux. One question is raised: would soap alone have been equally effective or do both a low pH (or even the acid type: citric and H_2S0_4) and surfactant (soap) need to be present. It is known that ionic surfactants can be affected by pH and so this is an avenue to consider in future. The mechanism of flux and separation performance enhancement must have a strong relationship with fouling reduction. By all previous arguments, the dosed surfactant coupled with the mechanical emulsifying action of the pump must have improved concentrate emulsion stability. The addition of Dawn (surfactant and ethanol) stabilized the oil-water interface and so chemically stabilized the unstable mechanical emulsion in the concentrate solution. This reduces the concentrate's fouling potential and the rate of deposition of unstable emulsoids onto the membrane surface. The stable emulsion effectively reduces the free oil concentration and this established a concentration driving force that drives oily deposits from the membrane surface into the bulk. This oil back-transport is sped up by the high shear from cross-flow. This back-transported oil is also mechanically emulsified and chemically stabilized once in solution – the stabilization (mixing) process is also convectively driven and hence is rapid. In so doing, the mechanical pumping and shear, combined with the stabilization action of the surfactant resulted in the sudden increase in flux and separation performance at t = 791 min., when the surfactant was dosed. The cross-flow velocity for Run 6 was 5m/s, but future work should evaluate membrane performance at reduced velocities so as to reduce pumping costs.

The surfactant cleaning effect was observed to be faster in the Teflon® membrane (M2). The surface chemistry difference between the two membranes were important, but the slow release of its oil foulants by the silica membrane (M1) strongly pointed toward reversible pore fouling rather than only reversible surface fouling – conversely surface fouling seemed to be the sole reversible fouling mechanism for the Teflon® membrane (M2). Reversible surface oil fouling would be easily removed by surfactant coupled with the high shear operation (5 m/s crossflow), as was the case for M2. Reversible pore plugging with oil-covered-solids, was suspected to be the reversible fouling mechanism for M1 (silica). As the innermost oil-coated-solid is released by surfactant removing the oil matrix inside which the solid is lodged, the next particle can be released and so forth, until the pore is unblocked. The reason suggested for reduced oil-coated-solids plugging for the Teflon® membrane is that this membrane is probably less oleophobic than the silica membrane. So when oil-coated solids enter the Teflon® pore, the oil is transferred to the Teflon® pore wall, the effective particle size is reduced and the solid is released. The strongly oleophobic silica membrane will not accept the oil film from these oil-coated solids and hence the particles remained lodged in the pores, only released by surfactants.

At t = 908 min., 2 hours after differential operation and dosing were initiated, flux from M1 equaled that for M2 = 90 lmh. Flux from M1 had dropped over these two hours from 180 to 90 lmh, while that from M1 had increased from 67 to 90 lmh. The difference in performance supports the hypothesis that the release of oily foulants from M2 was faster than that of M1 and essentially complete at the point of dosing. However, the high flux on M2 was counter-productive and unsustainable as it re-delivered foulants to the surface of M2 at a rate proportional to the fluxes. The slow increase in M1 supported the hypothesis that the pores in the oleophobic M2 were still continuously ridding itself of oil and solids over this period.

One interesting similarity between the silica and Teflon® membranes was that the addition of surfactant improved process flux such that *flux permeability* at the instant of surfactant dosing (t ~ 791 min.) for *both* membranes was 30 % of its *native, pure water* permeability, measured before any process testing was performed. The flux at the instant of dosing was 21 lmh/bar for M1 (native water flux of 68 lmh/bar) and for M2 it was 112 lmh/bar (native flux was 382 lmh/bar). The normalized performance was the same for the two membranes, but in the end, the high absolute flux of M2 curtailed its performance by re-depositing foulants and so the equilibrium flux was much lower than 182 lmh. The fact that normalized recovery was similar at the point of dosing surfactant (t = 791 min.) supports the theory of a sudden reversible membrane surface fouling layer removal – hence reversible surface fouling alone, reduced the flux performance by 30%. The rest of the recovery process from M1 is thus purely reversible pore unplugging. It is clear that surfactant dosing is an area to focus on for process optimization.

At t = 933 min., flux for M1 = 97 lmh and M2 = 72 lmh and turbidities were: M1 = 0.54 NTU and M2 = 0.71 NTU. Once again, the turbidities were in the range required for RO processes, notwithstanding the chemical make-up of the permeate. Visual observation of the concentrate tank showed that the level of foam had decreased and so it seems that the unassimilated surfactant levels have dropped. For this to have occurred in a differential filtration process requires that the surfactant must be participating in an ongoing process inside the process that 'consumes' it or binds into some form, e.g. forming a more stable oil-water emulsion – this again supports the pore unplugging mechanism of M1. When the free surfactant level dropped, the enhancement on M1 slowed. And at t = 945 min, ca. 10 ml more surfactant was dosed into solution and M1 showed further flux increases, while that of M2 was not altered – so M1 was still experiencing pore unplugging. It seemed as if the emulsion was stabilizing further, but the added stability was not affecting M2 in a sudden manner any longer. M2 showed signs of a general flux decrease to a minimum flux of ca. 63 lmh. M1 showed a large increase from 97 to 143 lmh in the five minutes after this second dose of surfactant. M1 clearly still responded positively to increases in emulsion stability and the difference in surface chemistry between M1 (silica membrane) and the emulsion and M2 (Teflon® membrane) and the emulsion is distinctive and presents another area for further exploration.

The differential process was left undisturbed from the time of the second surfactant dose (t = 945 min.) to t = 1,365 min, a period of 7 hours. At t = 1,200 min., permeate turbities were still within the RO range: M1 = 0.37 NTU and M2 = 0.69 NTU and fluxes were steady for M1 at ca. 145 lmh, while that for M2 had increased from 63 to 80 lmh and seemed to be going through a maximum flux at this time. From t = 1,200 to 1,354 min. the process fluxes dropped slightly to: M1 = 133 lmh and M2 = 77 lmh. The higher net flux of the hydrophilic and oleophobic silica membrane, M2, agreed with the previously discussed theoretical prediction: although more oleophobic membranes fouled more readily (reversible pore blocking and surface blinding) in the presence of an unstable oil emulsion, it operated much better in a stable oil emulsion.

At t ~ 1,365 min, the process was switched back to the *batch concentration mode* and ca. 10 ml of surfactant was dosed into the concentrate at this point. A number of process changes occurred as the change to a BCT took place: (i) permeate was now removed and not recycled and so the acidity and free surfactant concentration of the retentate was reduced; (ii) process temperature (ca. 65°C) dropped as hot permeate was removed and replaced by ambient (ca. 25°C) feed and so the concentrate viscosity was increased and (iii) the oil-to-surfactant ratio increased (after a while) and led to emulsion instability. These factors clearly led to a less favorable filtration process and hence there was a sudden and persistent decline in the flux levels over the BCT period t = 1,365 to 1,430 min. as shown in Fig. C-11. At the end of the test, turbidity was slightly higher at: M1 = 1.39 NTU and M2 = 1.59 NTU.

Summary discussion

Overall, Runs 1 through 5 were mainly exploratory tests where observations of the system performance under a number of batch concentration conditions were made. These shed light on what the important processes that drove the filtration steps were. From the results of these studies, it was clear that at start-up, the membranes operated right near the water flux. However, flux deteriorated sharply in the first one or two hours after start-up and worsened, at first, as the oil concentration increased while in an unstable emulsion regime. Large TSS aggravated this effect, as it was suspected that unstable emulsoids formed oil-solids complexes, which fouled the membranes severely and also plugged the pores of more oleophobic membranes.

In Run 6, surfactant (and possibly also acid) dosing improved emulsion stability and also flux stability and showed that stable and relatively large flux could be maintained over a 7-hour period, t = 950 – 1365 min. Hence surfactant dosing was required for process improvement and this is an important area of future study. Run 6 highlighted that different membrane materials strongly affected the surface chemistry and hence equilibrium flux attained with that membrane for a set process condition. Membrane material selection is vital and should be carefully considered for each processes' operating regime. For example, the two different membranes tested in Run 6 behaved completely differently when the emulsion was stable (second half of Run 6) versus when the emulsion was unstable (first half of Run 6).

What was clear from the results of Run 6 was that permeate quality and flux rate were inextricably linked and that, when the one was good, so too was the other. This has been CeraMem's experience with a number of oil-emulsion filtrations: if the process chemistry is correct, the flux rate and permeate quality will be good; if not, it will fail in both those areas (all-or-nothing type process).

Process Economics

CeraMem conducted a preliminary economic evaluation of a proposed commercial filtration process. In order to develop process costs, an operating regime had to be developed using the field test data. This regime would enable estimation of the cleaning and dosing chemicals costs, dosing and cleaning frequency as well as other in-line process maintenance techniques (back-flush, relaxation, etc) to be employed. The operating regime will also give the pumping rate and pressure (TMP) requirements, at a desired flux rate. This data enables estimation of process capital and operating costs and hence the economic feasibility of a produced water reverse-osmosis pre-treatment process based on CeraMem's membranes. Due to the fairly short-term nature of the test data, membrane life-time and hence replacement costs could not be estimated from it and was estimated from similar long-term membrane installations performed by CeraMem. Operating cost estimates also included capital recovery and maintenance costs. The economic analysis also included pre-treatment such as an API and sand filter bank prior to the ultrafiltration plant.

Suggested scaled-up process design

Steady state membrane filtration processes that operate at a steady, but higher-than-feed concentration and that require some control of retentate concentration require a feed and bleed design. In this particular case, the bleed stream is returned to the API separator where it is demulsified (by dilution, quiescence and possibly by adding demulsifiers) and its oil content recovered (Figure C-12). Large-scale processes often operate as staged units with each stage having the ability to operate at independent concentrations generally in a series arrangement so as to be able to concentrate more in each successive stage. However, in this program, the test results made it very clear that the process was sensitive to concentrate TOG and that there was a small range of operation of TOG that would enable successful filtration. For this reason all stages would be operated at the same concentrate TOG level, with stages in a parallel arrangement. CeraMem has experience with design of such staged equipment in a recent oil emulsion filtration system deployed for industrial service for fall of 2005. Using that design, five full-scale CeraMem membranes comprise a single stage as per the arrangement shown in Figure C-13. The required number of stages-in-parallel will then be determined for the system including over-design for off-line cleaning of membranes.

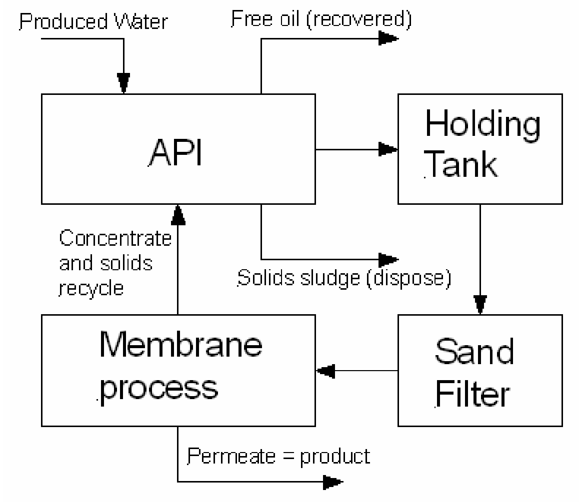


Figure C-12
Proposed feed and bleed process schematic

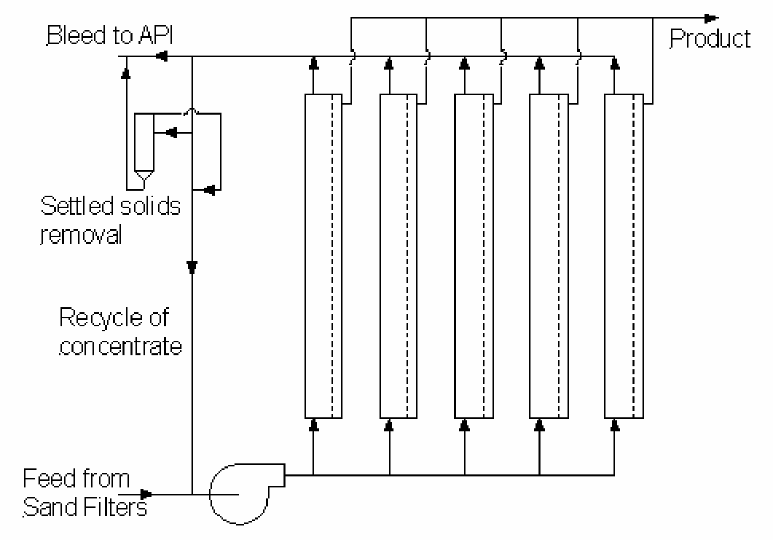


Figure C-13
One stage of the full-scale feed-and-bleed stages-in-parallel process design

For conservative scale-up it is crucial to select a most likely TMP, back-flush rate, cross-flow velocity and chemical cleaning frequency from the range associated with reliable, albeit short-term test results. The selected design basis for the full-scale system, based on the test runs, is shown in Table C-4. This design basis was conservative as the feeling was that a flux of 200 lmh could be achieved in an optimized process.

Economic viability of the membrane process

All process economics calculations used the operating parameters listed in Table C-4 and inputs of Table C-5. CeraMem's projected commercial full-scale membrane (area 38 m^2, length 60", diameter 200 mm), larger than that shown in Fig. C-14, was used for economics calculations. Outputs from the process economics calculations are shown in Table C-6.

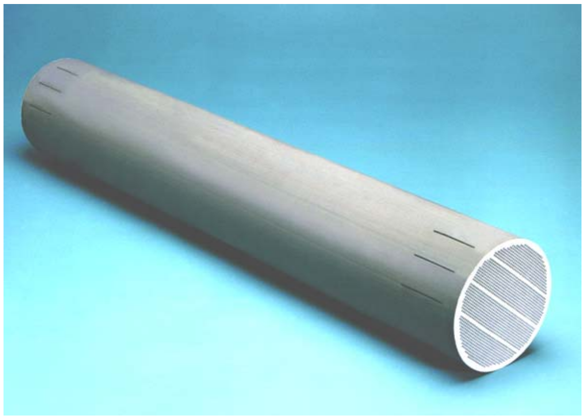


Figure C-14
CeraMem's 10.7 m2 commercial membrane module (5.66" x 34")

Table C-4
Operating parameters for envisaged scaled-up process

Parameter	Design Value	Comment
TMP_{max}	2.5 bar	Larger average TMP was 2 bar (M1)
Cross-flow	3 m/s	3 m/s (Run 1 and 2) was felt to be sufficient
Surfactant dose	Continuous	Use 4:1 TOG:Emulsifier concentration
Chemical clean	Monthly	Clean only the out-of-service 20% of stages
Back-flush	None	Unclear that back-flush was valuable from tests
Relaxation	Intermittent	30 sec. every 10 min. (Run 6)
Net flux	120 lmh	Conservative (see Run 6) – target is > 150 lmh

Table C-5
Economic parameters employed in calculating process viability

Parameter	Value	Comment
Design Flow	53,000 bbl/day	Desired design flow rate
New membrane cost	250 $/m^2	Includes housing
Replacement membrane cost	250 $/m^2	Includes housing
Membrane life	15 years	Estimate from CeraMem data
Membrane support system cost	2x membrane cost	Including pumps and controls
Process installation cost	0.45x system cost	Membrane process installation
Cost of capital	7% p.a. for 20 yrs	Amortized
Capacity over-design	20%	For cleaning cycle and overload
Membranes per stage	5	Booster + feed pump per stage
Power cost	7 c/kWhr	For booster and feed pumps
Pump efficiency	70%	For power calculation
Emulsifier (surfactant) cost	$14 per pure gallon	Bulk cost from Sierra Chemical
Cleaning chemicals composition	Citric acid + surfactant	5% citric acid + 20ppm surfactant
Cleaning chemicals cost	$0.075/liter	Citric acid $1/lb; surfactant above
Cleaning chemicals weekly load	50*lumen volume	On 20% over-design membranes
Burdened labor cost	$50/hour	DOE: experienced technician
Maintenance	3% of capital	Annual maintenance cost
Labor required (most routine items can be automated, but not membrane replacement)	10 min/stage daily on 6day/week routine (separate maintenance cost); 1/month, 1 hr/membrane chemical clean; 2 hrs/membrane every 15 years for membrane replacement.	

As shown in Table C-6, the specific annualized cost per barrel of treated water (membrane plus pre-treatment contribution) is \$0.051/bbl. The pre-treatment to the ultrafiltration (API and sand filter bank) is included in the capital cost estimate, while the operating costs for pre-treatment are included in the annual maintenance cost. A specific cost of \$0.051/bbl (incl. pre-treatment) would be lower than the typical cost for third party water disposal which ranged from \$0.10/bbl – \$5.00/bbl as taken from a study of produced water disposal in the Rocky Mountain region[xvi]. From these literature rates, it is clear that cleaning ones own produced water for down-hole disposal for \$0.051/bbl can be cheaper than outsourcing, notwithstanding risks associated with designing and operating the well and treatment unit.

Table C-6
Process cost estimates based on assumed inputs in Tables 4 and 5.

Capital cost	Total Stages	Cost of capital	Membrane Replacement	Pumping Cost	Total p.a. Cost	Specific Cost
\$3.93M	19	\$371K p.a.	\$60K p.a.	\$122K p.a.	\$932K p.a.	\$0.051/bbl

The current project aims to reuse the treated produced water, after UF and RO, at power stations. The best results obtained in this work were those of Run 6. Run 6 produced good flux levels and permeate turbidities, for a steady differential test of longer than 6 hours, of below 1 NTU and TSS of below 16.1 mg/l. The true TSS was expected to be lower than 16.1 mg/l because this analytical sample was taken before the filtration performance showed a marked improvement upon surfactant dosing. Not knowing the exact chemical make-up of the permeate and the exact requirements of a RO plant that would be installed to upgrade the water for use in a power plant makes it difficult to estimate a cost for post treatment. However, rough estimates of RO process costs were determined by over-estimating operating costs as 50% of the capital cost (p.a.) for cases 1 and 2 in Table C-7. Case 3 was used for comparison and reasonability check for the rough estimations of case 1 and 2.

Table C-7
Specific cost for RO water treatment

#	Capital and scale	Specific cost	Comments and source
1	\$47K – 0.024 MGD	\$0.12/bbl	Bottled water pre-packaged RO unit[xvii] with 5µm pre-treatment.
2	\$3M – 1 MGD	\$0.17/bbl	Novartis AG contact lense water recycling plant[xviii] (high grade product required)
3	Capital cost not known – 0.32 MGD	\$0.07/bbl	Dow Chemical surface water treatment economic study incl. 5µm pre-treatment[xix]

Assuming a conservative value of \$0.2/bbl for RO, the total produced water treatment cost for feed to the power plant cooling line would be ca. \$0.251/bbl. This is an overestimate of the specific treatment costs, but it is nevertheless reasonable when considering that water can be re-sold locally in Farmington, NM for well drilling for up to \$1/bbl[xx]. Government subsidies for such a large-scale water recycling process could reduce costs considerably. The environmental conditions may also become such that, especially in the San Juan Basin region, water could become scarcer and hence a price of ca. \$0.3/bbl could be highly competitive if lack-of-water could halt power production plants from operating. Process optimization and long term testing of the UF step for RO pretreatment, could provide more accurate costs and the hope its that further testing would improve process economics. For example, if a stable flux of 150 lmh (as observed in Run 6 for M1) is employed in calculations, then conservative water treatment costs would be: $0.042/bbl for UF (see Fig. C-15) and hence ca. $0.242/bbl for UF + RO (including pre-treatment).

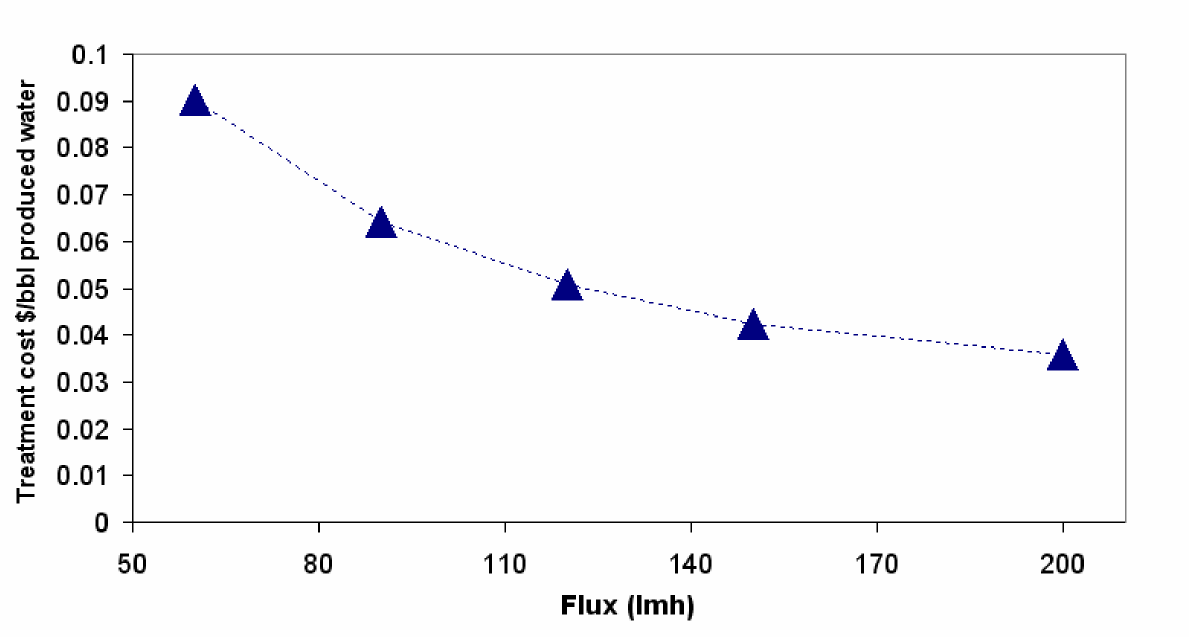


Figure C-15
Scaling of the specific cost of produced water ultrafiltration (including pre-treatment) with the process flux.

Conclusions and recommendations

The field tests performed in this work were largely preliminary and exploratory in nature: more than 80% of all the tests were batch runs. Some interesting observations were made through 122 hours of process testing performed over a 9-day period at McGrath SWD, operated by Burlington Resources in Farmington NM. The results indicated that the cost of pre-treating (UF) the challenging and highly variable produced water feed stream ranged from $0.09/bbl to $0.04/bbl for a UF flux of 60 to 150 lmh respectively. Long-term, feed and bleed tests would be required to refine these cost estimates.

The tests performed in this work concentrated mainly on determining process feasibility: high enough flux levels and reasonable permeate turbities. A next step would be to perform feed and bleed trials of a longer-term nature, with concentrate temperature control, to determine optimized operating parameters for observing long-term membrane fouling trends. This will enable determination of a more accurate chemical cleaning regime and the labor and chemicals cost associated with this regime. Future tests should also investigate emulsifier-dosing economics, based on tests with a range of emulsifiers. Note that dilute concentrate emulsions offer lower fouling potential to membranes, but in a closed loop process (no water wastage) would mean that the feed supply rate would be larger. This may require a larger API to get the same demulsification residence time or could require chemical demulsification, and so an economic optimum exists regarding concentrate TOG. A mechanism of fouling and fouling reversal was proposed in this work. This mechanism can serve as a useful guide to optimize membrane surface properties (membrane materials selection) for future work.

Finally, in order to develop the applicability of the UF treated water for feed to an RO plant, more targeted separation performance analysis, e.g. silt density index (SDI), total dissolved solids (TDS) as well as component analysis needs to be performed along with the usual indicators of filtration performance, used in this work.

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[i] Membrane cross-flow filtration is a process that can use microfiltration (MF) membranes, pore size ca. 0.1-to-1 μm, or ultrafiltration (UF) membranes, pore size ca. < 0.1 μm.

[ii] L. Comb, "Wastewater Recovery Using Reverse Osmosis".

http://www.osmonics.com/products/Page724.htm

[iii] "Cross-flow Membrane Separation Systems Study," Offshore Operators Committee, report by Paragon Engineering Services, PES Project No. 90421, December, 1990.

[iv] A. Chen et al., "Removal of Oil, Grease, and Suspended Solids from Produced Water with Ceramic Crossflow Microfiltration," *SPE Prod. Eng.*, 131-136 (1991).

[v] H. Norris and M. Quattrini, "A Field Study on the Use of Membranes to Remove Oil from Produced Water," *Proc. 1994 AIChE Spring Natl. Mtg.*, April, 1994.

[vi] Personal Communication, 1994-95, Ian Rubenstein, Exxon Chemical Co.

[vii] Cheryan, M, 1998, Ultrafiltration and Microfiltration Handbook, Technomic Publishing Company Incorporated, Lancaster, Pennsylvania.

[viii] J. D. Van Hamme et al., "Recent Advances in Petroleum Microbiology" Microbiology And Molecular Biology Reviews, Vol. 67, No. 4: 503–549 (2003)

[ix] J. M. Benito et al., "Design and construction of a modular pilot plant for the treatment of oil-containing wastewaters", Desalination vol. 147: 5-10 (2002)

[x] E. Dickinson, "Interfacial interactions and the stability of oil-in-water emulsions", Pure & App. Chern., Vol. 64, No. 11: 1721-1724 (1992).

[xi] H.M. Pars and D.Th. Meijer, "Removal of dissolved hydrocarbons from production water by Macro Porous Polymer Extraction (MPPE)", 1998 SPE Int. Conf. on Health, Safety and Environment in Oil and Gas Exploration and Production held in Caracas, Venezuela, 7–10 June 1998. Also on-line at: http://www.environmental-expert.com/articles/article1057/article1057.htm

[xii] I.B. Ivshina1 et al., "Oil desorption from mineral and organic materials using biosurfactant complexes produced by Rhodococcus species." World J. Microbio. Biotech., vol. 14 (5): 711-717 (1998).

[xiii] K. Urum and T. Pekdemir, "Evaluation of biosurfactants for crude oil contaminated soil washing." Chemosphere vol. 57: 1139–1150 (2004)

[xiv] Data from Robust Summary: http://www.petroleumhpv.org/Product_Categories/Crude_Oil/crude_oil.htm

[xv] T. Bilstad and E. Espedal, "Membrane separation of produced water" Water Sci. Technol. Vol. 34 (9): 239-246 (1996).

[xvi] Tom Hayes and Dan Arthur, "Overview of the emerging produced water treatment technologies" 11th Annual International Petroleum Environment Conference, Albuquerque, NM, Oct 12-15 2004.

[xvii] http://www.aquatechnology.net/24000GPDspecification.html

[xviii] http://www.bsr.org/CSRResources/IssueBriefDetail.cfm?DocumentID=49620

[xix] http://www.dow.com/liquidseps/design/ix_ro.htm

[xx] Personal conversation with Jim Jones, the McGrath SWD Process Specialist for Burlington Resources.

Appendix: Results summary for all field tests

Date	**Time**	**pH**	**TSS**	**TOG**	**NTU (M1)**	**NTU (M2)**	**NTU Mix**	**Mode**	**sample origin**	**M1**	**M2**	**Q (GPM)**	**Temp (°C)**	**Run number with comments**
22-Jun	0	feed pH						BCT	Concentrate	Silica	Teflon	10.4	67	run1 – start
23-Jun	720	feed pH		26.3			12.6	BCT	Permeate	Silica	Teflon			run1
23-Jun	0	feed pH						BCT	Concentrate	Titania	Teflon	10.61		run2 – start
24-Jun	544	7.57	587	294				BCT	Concentrate	Titania	Teflon			run2
24-Jun	544	6.84	133	22.8				n/a	SandfilterOut	Titania	Teflon			run2
24-Jun	544	feed pH		40.7			0.6	BCT	Permeate	Titania	Teflon			run2
24-Jun	0	acidic						BCT	Concentrate	Titania	Teflon	17.4	64	run3 – start: acidic due to acid rinse before hand
24-Jun	125	2.11	43.3	48.3			2.61	BCT	Permeate	Titania	Teflon			run3
25-Jun	725	feed pH		64			6.2	BCT	Permeate	Titania	Teflon			run3
25-Jun	1078	3					3.1	**DIFF**	Permeate	Titania	Teflon			run3 – dosed acid at this point
25-Jun	1078	3	283	**9.44**				**DIFF**	Concentrate	Titania	Teflon			run3
25-Jun	1166	7.12	83.3	42.3				n/a	CartridgeOut	Titania	Teflon			run3 – cartridge filter sample NTU = 24.1
25-Jun	0	feed pH						BCT	Concentrate	Silica	Teflon	16.4	70	run 3b – start
26-Jun	929	feed pH		41.1	3.55	5.87		BCT	Permeate	Silica	Teflon			run 3b
26-Jun	999	7.16	80	17.7				n/a	SandfilterOut	Silica	Teflon			run 3b
26-Jun	1325	7.79			3.7	2.3		BCT	Permeate	Silica	Teflon			run 3b
26-Jun	1325	7.79	950	78.4				BCT	Concentrate	Silica	Teflon			run 3b
26-Jun	0	feed pH						BCT	Concentrate	Silica	Teflon	16.5	60	run4 – start
27-Jun	785	7.51	66.7	30.3				n/a	SandfilterOut	Silica	Teflon			run4
27-Jun	785	7.51			17.2	17.8		BCT	Permeate	Silica	Teflon			run4 – permeate is murky.
27-Jun	1270	8.19	1350	315				BCT	Concentrate	Silica	Teflon			run4 – total Fe content: 99.2 mg/l
28-Jun	0	feed pH	230	159				n/a	SandfilterOut	Silica	Teflon			run5 – just prior to start-up
28-Jun	0	feed pH						BCT	Concentrate	Silica	Teflon	16.7	68.7	run5 – start
28-Jun	5	feed pH			1.33	3.1		BCT	Permeate	Silica	Teflon			run5
29-Jun	670	feed pH	500	540				BCT	Concentrate	Silica	Teflon			run5
29-Jun	0	feed pH						BCT	Concentrate	Silica	Teflon	16.9	70	run6 – start
29-Jun	5	feed pH			0.87	0.72		BCT	Permeate	Silica	Teflon			run6
30-Jun	720	7.94	16.1		2.55	2.84		**DIFF**	Permeate	Silica	Teflon			run6
30-Jun	795	acidic				1.37		**DIFF**	Permeate	Silica	Teflon			run6 – acid dosed
30-Jun	835	acidic			0.58	0.67		**DIFF**	Permeate	Silica	Teflon			run6 – surfactant dose - flux increases rapidly
30-Jun	925	acidic			0.54	0.71		**DIFF**	Permeate	Silica	Teflon			run6
30-Jun	1195	acidic			0.37	0.69		**DIFF**	Permeate	Silica	Teflon			run6 – more surfactant - M1 flux increases
30-Jun	1430	acidic			1.39	1.59		BCT	Permeate	Silica	Teflon			run6

D

SECTION 7, COST/BENEFIT ANALYSIS

Table D-1
Life-of-Project Produced Water Resource Summary

Life-of-Project Produced Water Resource Summary - Scenario 1

San Juan Generating Station

Project Year	Prod Water Volume Increase Over Prev Year	Backflow Volume Decline Over Prev Year	Tri-City BPD	Fairway BPD	Close-in BPD	Potential Produced Water BPD	Potential Backflow Water BPD	BHP Mine Water BPD	Prax Air CT Blowdown BPD	Purge Water BPD	Potential Water BPD	Potential Water AF/yr
2006	2.0%	0.0%	3,016	22,598	13,681	39,294	10,000	1,700	300	3,429	54,723	2,575
2007	2.0%	0.0%	3,076	23,050	13,955	40,080	10,000	1,700	300	3,429	55,509	2,611
2008	2.0%	0.0%	3,138	23,511	14,234	40,882	10,000	1,700	300	3,429	56,311	2,649
2009	1.9%	-2.0%	3,197	23,957	14,504	41,659	9,800	1,700	300	3,429	56,887	2,676
2010	1.5%	-2.0%	3,245	24,317	14,722	42,284	9,604	1,700	300	3,429	57,316	2,696
2011	1.0%	-2.0%	3,278	24,560	14,869	42,706	9,412	1,700	300	3,429	57,547	2,707
2012	0.5%	-2.0%	3,294	24,683	14,943	42,920	9,224	1,700	300	3,429	57,572	2,709
2013	0.0%	-2.0%	3,294	24,683	14,943	42,920	9,039	1,700	300	3,429	57,388	2,700
2014	-0.5%	-2.0%	3,278	24,559	14,868	42,705	8,858	1,700	300	3,429	56,992	2,681
2015	-1.0%	-2.0%	3,245	24,314	14,720	42,278	8,681	1,700	300	3,429	56,388	2,653
2016	-1.5%	-2.0%	3,196	23,949	14,499	41,644	8,508	1,700	300	3,429	55,580	2,615
2017	-1.9%	-2.0%	3,137	23,506	14,231	40,874	8,337	1,700	300	3,429	54,640	2,571
2018	-2.0%	-2.0%	3,075	23,040	13,949	40,064	8,171	1,700	300	3,429	53,664	2,525
2019	-2.0%	-2.0%	3,014	22,580	13,670	39,263	8,007	1,700	300	3,429	52,699	2,479
2020	-2.0%	-2.0%	2,953	22,128	13,397	38,478	7,847	1,700	300	3,429	51,754	2,435
2021	-2.0%	-2.0%	2,894	21,685	13,129	37,708	7,690	1,700	300	3,429	50,827	2,391
2022	-2.0%	-2.0%	2,836	21,252	12,866	36,954	7,536	1,700	300	3,429	49,919	2,348
2023	-2.0%	-2.0%	2,780	20,827	12,609	36,215	7,386	1,700	300	3,429	49,029	2,307
2024	-2.0%	-2.0%	2,724	20,410	12,357	35,491	7,238	1,700	300	3,429	48,157	2,266
2025	-2.0%	-2.0%	2,670	20,002	12,109	34,781	7,093	1,700	300	3,429	47,303	2,225

Table D-1 (page 2 of 3)
Life-of-Project Produced Water Resource Summary

Life-of-Project Produced Water Resource Summary - Scenario 2

San Juan Generating Station

Project Year	Prod Water Volume Increase Over Prev Year	Backflow Volume Decline Over Prev Year	Tri-City BPD	Fairway BPD	Close-in BPD	Potential Produced Water BPD	Potential Backflow Water BPD	BHP Mine Water BPD	Prax Air CT Blowdown BPD	Purge Water BPD	Potential Water BPD	Potential Water AF/yr
2006	2.00%	0.0%	3,016	22,598	13,681	39,294	10,000	1,700	300	3,429	54,723	2,575
2007	2.00%	0.0%	3,076	23,050	13,955	40,080	10,000	1,700	300	3,429	55,509	2,611
2008	2.00%	0.0%	3,138	23,511	14,234	40,882	10,000	1,700	300	3,429	56,311	2,649
2009	1.80%	-2.0%	3,194	23,934	14,490	41,618	9,800	1,700	300	3,429	56,846	2,674
2010	1.30%	-2.0%	3,236	24,245	14,678	42,159	9,604	1,700	300	3,429	57,191	2,691
2011	0.65%	-2.0%	3,257	24,402	14,774	42,433	9,412	1,700	300	3,429	57,273	2,694
2012	0.00%	-2.0%	3,257	24,402	14,774	42,433	9,224	1,700	300	3,429	57,085	2,686
2013	-0.65%	-2.0%	3,236	24,244	14,678	42,157	9,039	1,700	300	3,429	56,625	2,664
2014	-1.30%	-2.0%	3,194	23,929	14,487	41,609	8,858	1,700	300	3,429	55,896	2,630
2015	-2.00%	-2.0%	3,130	23,450	14,197	40,777	8,681	1,700	300	3,429	54,887	2,582
2016	-2.80%	-2.0%	3,042	22,793	13,800	39,635	8,508	1,700	300	3,429	53,571	2,520
2017	-3.40%	-2.0%	2,939	22,019	13,330	38,288	8,337	1,700	300	3,429	52,054	2,449
2018	-3.80%	-2.0%	2,827	21,182	12,824	36,833	8,171	1,700	300	3,429	50,432	2,373
2019	-4.00%	-2.0%	2,714	20,335	12,311	35,359	8,007	1,700	300	3,429	48,795	2,296
2020	-4.00%	-2.0%	2,605	19,521	11,818	33,945	7,847	1,700	300	3,429	47,221	2,222
2021	-4.00%	-2.0%	2,501	18,740	11,346	32,587	7,690	1,700	300	3,429	45,706	2,150
2022	-4.00%	-2.0%	2,401	17,991	10,892	31,284	7,536	1,700	300	3,429	44,249	2,082
2023	-4.00%	-2.0%	2,305	17,271	10,456	30,032	7,386	1,700	300	3,429	42,847	2,016
2024	-4.00%	-2.0%	2,213	16,580	10,038	28,831	7,238	1,700	300	3,429	41,498	1,952
2025	-4.00%	-2.0%	2,124	15,917	9,636	27,678	7,093	1,700	300	3,429	40,200	1,891

Table D-1 (page 3 of 3)
Life-of-Project Produced Water Resource Summary

Life-of-Project Produced Water Resource Summary - Scenario 3

San Juan Generating Station

Project Year	Prod Water Volume Increase Over Prev Year	Backflow Volume Decline Over Prev Year	Tri-City BPD	Fairway BPD	Close-in BPD	Potential Produced Water BPD	Potential Backflow Water BPD	BHP Mine Water BPD	Prax Air CT Blowdown BPD	Purge Water BPD	Potential Water BPD	Potential Water AF/yr
2006	2.00%	0.0%	3,016	22,598	13,681	39,294	10,000	1,700	300	3,429	54,723	2,575
2007	2.00%	0.0%	3,076	23,050	13,955	40,080	10,000	1,700	300	3,429	55,509	2,611
2008	2.00%	0.0%	3,138	23,511	14,234	40,882	10,000	1,700	300	3,429	56,311	2,649
2009	1.70%	-2.0%	3,191	23,910	14,476	41,577	9,800	1,700	300	3,429	56,806	2,672
2010	0.95%	-2.0%	3,221	24,137	14,613	41,972	9,604	1,700	300	3,429	57,005	2,682
2011	0.00%	-2.0%	3,221	24,137	14,613	41,972	9,412	1,700	300	3,429	56,812	2,673
2012	-0.90%	-2.0%	3,192	23,920	14,482	41,594	9,224	1,700	300	3,429	56,246	2,646
2013	-1.80%	-2.0%	3,135	23,490	14,221	40,845	9,039	1,700	300	3,429	55,313	2,602
2014	-2.70%	-2.0%	3,050	22,855	13,837	39,743	8,858	1,700	300	3,429	54,030	2,542
2015	-3.60%	-2.0%	2,941	22,033	13,339	38,312	8,681	1,700	300	3,429	52,422	2,466
2016	-4.55%	-2.0%	2,807	21,030	12,732	36,569	8,508	1,700	300	3,429	50,505	2,376
2017	-5.40%	-2.0%	2,655	19,894	12,044	34,594	8,337	1,700	300	3,429	48,360	2,275
2018	-5.80%	-2.0%	2,501	18,741	11,346	32,588	8,171	1,700	300	3,429	46,187	2,173
2019	-6.00%	-2.0%	2,351	17,616	10,665	30,632	8,007	1,700	300	3,429	44,068	2,073
2020	-6.00%	-2.0%	2,210	16,559	10,025	28,794	7,847	1,700	300	3,429	42,070	1,979
2021	-6.00%	-2.0%	2,077	15,566	9,424	27,067	7,690	1,700	300	3,429	40,186	1,891
2022	-6.00%	-2.0%	1,953	14,632	8,858	25,443	7,536	1,700	300	3,429	38,408	1,807
2023	-6.00%	-2.0%	1,836	13,754	8,327	23,916	7,386	1,700	300	3,429	36,730	1,728
2024	-6.00%	-2.0%	1,725	12,929	7,827	22,481	7,238	1,700	300	3,429	35,148	1,654
2025	-6.00%	-2.0%	1,622	12,153	7,358	21,132	7,093	1,700	300	3,429	33,654	1,583

Table D-2
Life-of-Project Recoverable Water by Case

Life-of-Project Recoverable Water by Case - Scenario 1

San Juan Generating Station

		Recoverable Water					*Fraction of Recoverable Water*				
	Potential	50%	60%	70%	80%	90%	50%	60%	70%	80%	90%
Project	Water	Case 1	Case 2	Case 3	Case 4	Case 5	Case 1	Case 2	Case 3	Case 4	Case 5
Year	AF/yr	AF/yr	AF/yr	AF/yr	AF/yr	AF/yr	Fraction	Fraction	Fraction	Fraction	Fraction
2006	2,575	789	923	1,057	1,191	1,325	30.6%	35.9%	41.1%	46.3%	51.5%
2007	2,611	1,033	1,229	1,424	1,737	2,081	39.6%	47.0%	54.5%	66.5%	79.7%
2008	2,649	1,225	1,468	1,712	2,013	2,319	46.2%	55.4%	64.6%	76.0%	87.5%
2009	2,676	1,303	1,566	1,830	2,118	2,396	48.7%	58.5%	68.4%	79.1%	89.5%
2010	2,696	1,339	1,611	1,883	2,156	2,428	49.7%	59.7%	69.8%	79.9%	90.0%
2011	2,707	1,343	1,616	1,890	2,163	2,436	49.6%	59.7%	69.8%	79.9%	90.0%
2012	2,709	1,343	1,617	1,890	2,163	2,437	49.6%	59.7%	69.8%	79.9%	90.0%
2013	2,700	1,339	1,611	1,884	2,156	2,428	49.6%	59.7%	69.8%	79.9%	89.9%
2014	2,681	1,331	1,601	1,871	2,141	2,411	49.6%	59.7%	69.8%	79.9%	89.9%
2015	2,653	1,318	1,585	1,852	2,119	2,386	49.7%	59.7%	69.8%	79.9%	89.9%
2016	2,615	1,301	1,564	1,826	2,089	2,352	49.8%	59.8%	69.8%	79.9%	89.9%
2017	2,571	1,281	1,539	1,797	2,055	2,312	49.8%	59.9%	69.9%	79.9%	90.0%
2018	2,525	1,261	1,514	1,766	2,019	2,271	49.9%	60.0%	70.0%	80.0%	90.0%
2019	2,479	1,241	1,488	1,736	1,984	2,231	50.1%	60.0%	70.0%	80.0%	90.0%
2020	2,435	1,221	1,464	1,706	1,949	2,192	50.2%	60.1%	70.1%	80.1%	90.0%
2021	2,391	1,202	1,440	1,677	1,915	2,153	50.3%	60.2%	70.1%	80.1%	90.0%
2022	2,348	1,183	1,416	1,649	1,882	2,115	50.4%	60.3%	70.2%	80.1%	90.1%
2023	2,307	1,164	1,393	1,621	1,849	2,078	50.5%	60.4%	70.3%	80.2%	90.1%
2024	2,266	1,146	1,370	1,594	1,818	2,041	50.6%	60.5%	70.3%	80.2%	90.1%
2025	2,225	1,128	1,348	1,567	1,786	2,006	50.7%	60.6%	70.4%	80.3%	90.1%
						Average Recovery - Starting at Year 5	50.00%	60.00%	70.00%	80.00%	90.00%

Table D-2 (page 2 of 3)
Life-of-Project Recoverable Water by Case

Life-of-Project Recoverable Water by Case - Scenario 2

San Juan Generating Station

		Recoverable Water					*Fraction of Recoverable Water*				
Project Year	Potential Water AF/yr	50% Case 1 BPD	60% Case 2 BPD	70% Case 3 BPD	80% Case 4 BPD	90% Case 5 BPD	50% Case 1 Fraction	60% Case 2 Fraction	70% Case 3 Fraction	80% Case 4 Fraction	90% Case 5 Fraction
2006	2,575	782	917	1,051	1,186	1,321	30.4%	35.6%	40.8%	46.1%	51.3%
2007	2,611	1,022	1,219	1,415	1,728	2,072	39.1%	46.7%	54.2%	66.2%	79.4%
2008	2,649	1,212	1,456	1,701	2,003	2,311	45.7%	55.0%	64.2%	75.6%	87.2%
2009	2,674	1,289	1,553	1,818	2,107	2,387	48.2%	58.1%	68.0%	78.8%	89.3%
2010	2,691	1,323	1,597	1,870	2,144	2,417	49.2%	59.3%	69.5%	79.7%	89.8%
2011	2,694	1,325	1,598	1,872	2,146	2,420	49.2%	59.3%	69.5%	79.6%	89.8%
2012	2,686	1,320	1,593	1,866	2,138	2,411	49.2%	59.3%	69.5%	79.6%	89.8%
2013	2,664	1,311	1,581	1,851	2,121	2,392	49.2%	59.3%	69.5%	79.6%	89.8%
2014	2,630	1,296	1,562	1,828	2,095	2,361	49.3%	59.4%	69.5%	79.7%	89.8%
2015	2,582	1,275	1,536	1,797	2,058	2,319	49.4%	59.5%	69.6%	79.7%	89.8%
2016	2,520	1,248	1,502	1,756	2,010	2,264	49.5%	59.6%	69.7%	79.8%	89.8%
2017	2,449	1,217	1,463	1,709	1,955	2,201	49.7%	59.7%	69.8%	79.8%	89.9%
2018	2,373	1,184	1,421	1,659	1,896	2,134	49.9%	59.9%	69.9%	79.9%	89.9%
2019	2,296	1,150	1,379	1,608	1,837	2,066	50.1%	60.1%	70.1%	80.0%	90.0%
2020	2,222	1,118	1,339	1,560	1,780	2,001	50.3%	60.3%	70.2%	80.1%	90.1%
2021	2,150	1,087	1,300	1,513	1,726	1,938	50.5%	60.4%	70.4%	80.2%	90.1%
2022	2,082	1,057	1,262	1,468	1,673	1,878	50.8%	60.6%	70.5%	80.4%	90.2%
2023	2,016	1,028	1,226	1,424	1,622	1,820	51.0%	60.8%	70.7%	80.5%	90.3%
2024	1,952	1,001	1,192	1,382	1,573	1,764	51.3%	61.0%	70.8%	80.6%	90.4%
2025	1,891	974	1,158	1,342	1,526	1,710	51.5%	61.2%	71.0%	80.7%	90.4%
		Average Recovery - Starting at Year 5					50.00%	60.00%	70.00%	80.00%	90.00%

Table D-2 (page 3 of 3)
Life-of-Project Recoverable Water by Case

Life-of-Project Recoverable Water by Case - Scenario 3

San Juan Generating Station

		Recoverable Water					*Fraction of Recoverable Water*				
Project Year	Potential Water AF/yr	50% Case 1 AF/yr	60% Case 2 AF/yr	70% Case 3 AF/yr	80% Case 4 AF/yr	90% Case 5 AF/yr	50% Case 1 Fraction	60% Case 2 Fraction	70% Case 3 Fraction	80% Case 4 Fraction	90% Case 5 Fraction
2006	2,575	771	907	1,043	1,179	1,315	30.0%	35.2%	40.5%	45.8%	51.1%
2007	2,611	1,007	1,204	1,402	1,715	2,060	38.5%	46.1%	53.7%	65.7%	78.9%
2008	2,649	1,193	1,440	1,687	1,990	2,299	45.0%	54.3%	63.7%	75.1%	86.8%
2009	2,672	1,268	1,535	1,802	2,092	2,375	47.5%	57.4%	67.4%	78.3%	88.9%
2010	2,682	1,301	1,577	1,852	2,127	2,402	48.5%	58.8%	69.1%	79.3%	89.6%
2011	2,673	1,297	1,571	1,845	2,119	2,394	48.5%	58.8%	69.0%	79.3%	89.6%
2012	2,646	1,285	1,556	1,828	2,098	2,370	48.6%	58.8%	69.1%	79.3%	89.5%
2013	2,602	1,266	1,533	1,799	2,065	2,331	48.7%	58.9%	69.1%	79.3%	89.6%
2014	2,542	1,241	1,500	1,759	2,018	2,278	48.8%	59.0%	69.2%	79.4%	89.6%
2015	2,466	1,208	1,459	1,710	1,960	2,211	49.0%	59.2%	69.3%	79.5%	89.7%
2016	2,376	1,170	1,410	1,651	1,891	2,132	49.2%	59.4%	69.5%	79.6%	89.7%
2017	2,275	1,127	1,356	1,585	1,814	2,044	49.5%	59.6%	69.7%	79.7%	89.8%
2018	2,173	1,083	1,301	1,519	1,736	1,954	49.8%	59.9%	69.9%	79.9%	89.9%
2019	2,073	1,041	1,247	1,454	1,660	1,867	50.2%	60.2%	70.1%	80.1%	90.1%
2020	1,979	1,000	1,197	1,393	1,589	1,785	50.5%	60.5%	70.4%	80.3%	90.2%
2021	1,891	963	1,149	1,335	1,521	1,707	50.9%	60.8%	70.6%	80.4%	90.3%
2022	1,807	927	1,104	1,280	1,457	1,634	51.3%	61.1%	70.9%	80.6%	90.4%
2023	1,728	893	1,061	1,229	1,397	1,565	51.7%	61.4%	71.1%	80.8%	90.5%
2024	1,654	861	1,021	1,181	1,340	1,499	52.1%	61.7%	71.4%	81.0%	90.7%
2025	1,583	831	983	1,135	1,286	1,438	52.5%	62.1%	71.7%	81.2%	90.8%
Average Recovery - Starting at Year 5							50.00%	60.00%	70.00%	80.00%	90.00%

Table D-3
Volume Revenue Elements

Volume Revenue Elements - Scenario 1 (2% Decline) - Case 1 (50% Recovery)

Project Year	Produced Water BPD	BackFlow BPD	Collection Center BPD	Total BR Inj Water BPD	Total BR BF Water BPD	Other BF Water BPD	Other Producer via SWD BPD	Other Producer Satellite BPD
2006	5,936	2,318	8,253	5,092	3,162	0	0	0
2007	8,453	3,275	11,729	5,790	3,828	0	106	2,006
2008	10,282	4,017	14,299	5,614	3,966	51	490	4,178
2009	11,395	4,450	15,844	5,148	3,883	566	906	5,341
2010	11,872	4,635	16,507	4,597	3,656	979	1,309	5,965
2011	11,991	4,542	16,533	4,149	3,457	1,086	1,631	6,211
2012	12,051	4,451	16,502	3,803	3,292	1,159	1,880	6,367
2013	12,051	4,362	16,413	3,546	3,183	1,179	2,067	6,438
2014	11,990	4,275	16,265	3,396	3,110	1,165	2,191	6,403
2015	11,870	4,190	16,060	3,251	3,013	1,177	2,301	6,318
2016	11,692	4,106	15,798	3,098	2,906	1,200	2,381	6,214
2017	11,476	4,024	15,500	2,988	2,825	1,199	2,428	6,061
2018	11,249	3,943	15,192	2,893	2,757	1,186	2,457	5,899
2019	11,024	3,864	14,888	2,800	2,690	1,174	2,500	5,724
2020	10,803	3,787	14,590	2,712	2,623	1,164	2,525	5,567
2021	10,587	3,711	14,299	2,626	2,557	1,154	2,547	5,413
2022	10,376	3,637	14,013	2,543	2,493	1,144	2,561	5,271
2023	10,168	3,564	13,732	2,462	2,430	1,134	2,574	5,132
2024	9,965	3,493	13,458	2,384	2,368	1,125	2,593	4,988
2025	9,765	3,423	13,189	2,308	2,308	1,115	2,610	4,847

Table D-3 (page 2 of 15)
Volume Revenue Elements

Volume Revenue Elements - Scenario 1 (2% Decline) - Case 2 (60% Recovery)

Project Year	Produced Water BPD	BackFlow BPD	Collection Center BPD	Total BR Inj Water BPD	Total BR BF Water BPD	Other BF Water BPD	Other Producer via SWD BPD	Other Producer Satellite BPD
2006	7,427	2,900	10,326	6,370	3,956	0	0	0
2007	10,576	4,098	14,674	7,244	4,789	0	132	2,509
2008	12,865	5,026	17,890	7,024	4,962	63	613	5,227
2009	14,256	5,567	19,823	6,441	4,859	708	1,133	6,682
2010	14,853	5,799	20,652	5,752	4,574	1,225	1,638	7,463
2011	15,002	5,683	20,685	5,190	4,325	1,358	2,041	7,771
2012	15,077	5,569	20,646	4,759	4,119	1,450	2,353	7,966
2013	15,077	5,458	20,535	4,437	3,982	1,476	2,585	8,054
2014	15,001	5,349	20,350	4,249	3,891	1,458	2,742	8,010
2015	14,851	5,242	20,093	4,067	3,769	1,473	2,879	7,905
2016	14,629	5,137	19,766	3,876	3,635	1,502	2,979	7,774
2017	14,358	5,034	19,392	3,738	3,534	1,500	3,037	7,583
2018	14,074	4,934	19,007	3,619	3,449	1,484	3,074	7,381
2019	13,792	4,835	18,627	3,503	3,366	1,469	3,128	7,161
2020	13,516	4,738	18,255	3,393	3,282	1,456	3,159	6,965
2021	13,246	4,643	17,890	3,286	3,199	1,444	3,187	6,773
2022	12,981	4,551	17,532	3,182	3,119	1,432	3,204	6,595
2023	12,722	4,460	17,181	3,081	3,040	1,419	3,220	6,421
2024	12,467	4,370	16,837	2,983	2,963	1,408	3,244	6,241
2025	12,218	4,283	16,501	2,888	2,888	1,395	3,266	6,065

Table D-3 (page 3 of 15)
Volume Revenue Elements

Volume Revenue Elements - Scenario 1 (2% Decline) - Case 3 (70% Recovery)

Project Year	Produced Water BPD	BackFlow BPD	Collection Center BPD	Total BR Inj Water BPD	Total BR BF Water BPD	Other BF Water BPD	Other Producer via SWD BPD	Other Producer Satellite BPD
2006	8,917	3,482	12,399	7,649	4,750	0	0	0
2007	12,699	4,921	17,620	8,698	5,750	0	159	3,013
2008	15,447	6,035	21,482	8,434	5,959	76	736	6,276
2009	17,118	6,684	23,802	7,733	5,834	851	1,361	8,024
2010	17,835	6,963	24,798	6,906	5,493	1,470	1,967	8,961
2011	18,013	6,824	24,837	6,232	5,193	1,631	2,450	9,330
2012	18,103	6,687	24,790	5,714	4,946	1,741	2,825	9,565
2013	18,103	6,554	24,657	5,328	4,782	1,772	3,104	9,671
2014	18,013	6,422	24,435	5,102	4,672	1,750	3,292	9,618
2015	17,832	6,294	24,126	4,883	4,526	1,768	3,457	9,492
2016	17,565	6,168	23,733	4,654	4,365	1,803	3,576	9,335
2017	17,240	6,045	23,285	4,488	4,244	1,801	3,647	9,105
2018	16,899	5,924	22,823	4,345	4,142	1,782	3,691	8,863
2019	16,561	5,805	22,366	4,206	4,041	1,764	3,756	8,599
2020	16,229	5,689	21,919	4,074	3,941	1,749	3,793	8,363
2021	15,905	5,575	21,480	3,945	3,841	1,734	3,827	8,132
2022	15,587	5,464	21,051	3,820	3,745	1,719	3,848	7,919
2023	15,275	5,355	20,630	3,699	3,650	1,704	3,866	7,710
2024	14,970	5,248	20,217	3,582	3,557	1,690	3,895	7,493
2025	14,670	5,143	19,813	3,467	3,467	1,675	3,921	7,282

Table D-3 (page 4 of 15)
Volume Revenue Elements

Volume Revenue Elements - Scenario 1 (2% Decline) - Case 4 (80% Recovery)

Project Year	Produced Water BPD	BackFlow BPD	Collection Center BPD	Total BR Inj Water BPD	Total BR BF Water BPD	Other BF Water BPD	Other Producer via SWD BPD	Other Producer Satellite BPD
2006	10,409	4,064	14,473	8,929	5,545	0	0	0
2007	16,134	5,744	21,878	10,800	7,140	0	197	3,741
2008	19,257	7,044	26,302	10,327	7,296	0	911	7,768
2009	20,506	7,803	28,309	9,198	6,939	864	1,640	9,669
2010	20,819	8,128	28,947	8,062	6,412	1,716	2,296	10,461
2011	21,027	7,965	28,992	7,275	6,061	1,904	2,860	10,891
2012	21,132	7,806	28,938	6,670	5,774	2,032	3,297	11,165
2013	21,132	7,650	28,782	6,219	5,582	2,068	3,624	11,289
2014	21,026	7,497	28,523	5,956	5,454	2,043	3,843	11,228
2015	20,816	7,347	28,163	5,701	5,283	2,064	4,036	11,080
2016	20,504	7,200	27,704	5,432	5,095	2,105	4,175	10,897
2017	20,125	7,056	27,181	5,239	4,954	2,102	4,257	10,628
2018	19,726	6,915	26,641	5,072	4,835	2,080	4,308	10,345
2019	19,332	6,777	26,108	4,910	4,717	2,059	4,384	10,037
2020	18,945	6,641	25,586	4,755	4,600	2,041	4,427	9,763
2021	18,566	6,508	25,074	4,606	4,484	2,025	4,467	9,493
2022	18,195	6,378	24,573	4,460	4,371	2,007	4,491	9,244
2023	17,831	6,251	24,081	4,318	4,261	1,989	4,513	9,000
2024	17,474	6,126	23,600	4,181	4,153	1,973	4,546	8,747
2025	17,125	6,003	23,128	4,047	4,047	1,956	4,577	8,500

Table D-3 (page 5 of 15)
Volume Revenue Elements

Volume Revenue Elements - Scenario 1 (2% Decline) - Case 5 (90% Recovery)

Project Year	Produced Water BPD	BackFlow BPD	Collection Center BPD	Total BR Inj Water BPD	Total BR BF Water BPD	Other BF Water BPD	Other Producer via SWD BPD	Other Producer Satellite BPD
2006	11,900	4,646	16,546	10,207	6,339	0	0	0
2007	19,754	7,062	26,816	13,237	8,752	0	241	4,586
2008	22,729	8,502	31,231	12,262	8,663	0	1,082	9,224
2009	23,562	9,060	32,622	10,599	7,996	1,064	1,880	11,084
2010	23,800	9,292	33,092	9,216	7,330	1,962	2,625	11,959
2011	24,038	9,106	33,144	8,317	6,929	2,177	3,270	12,451
2012	24,158	8,924	33,082	7,625	6,601	2,323	3,770	12,764
2013	24,158	8,746	32,904	7,110	6,381	2,365	4,143	12,906
2014	24,037	8,571	32,608	6,809	6,235	2,336	4,393	12,836
2015	23,797	8,399	32,196	6,517	6,040	2,360	4,614	12,666
2016	23,440	8,231	31,671	6,210	5,825	2,406	4,773	12,457
2017	23,007	8,067	31,073	5,989	5,663	2,404	4,867	12,150
2018	22,551	7,905	30,456	5,799	5,527	2,378	4,925	11,827
2019	22,100	7,747	29,847	5,613	5,393	2,354	5,012	11,475
2020	21,658	7,592	29,250	5,436	5,259	2,334	5,061	11,161
2021	21,225	7,440	28,665	5,265	5,126	2,314	5,107	10,853
2022	20,800	7,292	28,092	5,098	4,997	2,294	5,135	10,567
2023	20,384	7,146	27,530	4,936	4,871	2,274	5,160	10,288
2024	19,977	7,003	26,979	4,780	4,747	2,255	5,197	10,000
2025	19,577	6,863	26,440	4,627	4,627	2,236	5,233	9,718

Table D-3 (page 6 of 15)
Volume Revenue Elements

Volume Revenue Elements - Scenario 2 (4% Decline) - Case 1 (50% Recovery)

Project Year	Produced Water BPD	BackFlow BPD	Collection Center BPD	Total BR Inj Water BPD	Total BR BF Water BPD	Other BF Water BPD	Other Producer via SWD BPD	Other Producer Satellite BPD
2006	5,855	2,286	8,141	5,022	3,119	0	0	0
2007	8,339	3,231	11,569	5,711	3,776	0	104	1,978
2008	10,143	3,962	14,105	5,538	3,912	50	483	4,121
2009	11,240	4,389	15,629	5,078	3,831	558	893	5,268
2010	11,711	4,572	16,283	4,535	3,607	965	1,292	5,884
2011	11,787	4,481	16,267	4,082	3,401	1,080	1,603	6,102
2012	11,787	4,391	16,178	3,729	3,228	1,163	1,837	6,221
2013	11,710	4,303	16,013	3,460	3,105	1,198	2,005	6,245
2014	11,558	4,217	15,775	3,294	3,016	1,201	2,107	6,157
2015	11,327	4,133	15,459	3,129	2,900	1,233	2,189	6,009
2016	11,009	4,050	15,060	2,953	2,770	1,280	2,232	5,825
2017	10,635	3,969	14,604	2,815	2,662	1,307	2,237	5,584
2018	10,231	3,890	14,121	2,689	2,563	1,327	2,217	5,325
2019	9,822	3,812	13,634	2,564	2,463	1,348	2,206	5,051
2020	9,429	3,736	13,165	2,447	2,367	1,369	2,178	4,804
2021	9,052	3,661	12,713	2,335	2,273	1,388	2,149	4,567
2022	8,690	3,588	12,277	2,228	2,184	1,404	2,113	4,349
2023	8,342	3,516	11,858	2,126	2,098	1,418	2,076	4,140
2024	8,008	3,446	11,454	2,029	2,015	1,430	2,045	3,934
2025	7,688	3,377	11,065	1,936	1,936	1,440	2,013	3,739

Table D-3 (page 7 of 15)
Volume Revenue Elements

Volume Revenue Elements - Scenario 2 (4% Decline) - Case 2 (60% Recovery)

Project Year	Produced Water BPD	BackFlow BPD	Collection Center BPD	Total BR Inj Water BPD	Total BR BF Water BPD	Other BF Water BPD	Other Producer via SWD BPD	Other Producer Satellite BPD
2006	7,355	2,872	10,226	6,309	3,918	0	0	0
2007	10,474	4,058	14,533	7,174	4,743	0	131	2,485
2008	12,740	4,977	17,718	6,956	4,915	63	607	5,177
2009	14,118	5,513	19,632	6,378	4,812	702	1,122	6,618
2010	14,710	5,743	20,453	5,696	4,530	1,213	1,622	7,391
2011	14,805	5,628	20,434	5,127	4,272	1,356	2,013	7,665
2012	14,805	5,516	20,321	4,684	4,054	1,461	2,308	7,814
2013	14,709	5,405	20,114	4,346	3,901	1,504	2,518	7,845
2014	14,518	5,297	19,815	4,137	3,789	1,509	2,647	7,734
2015	14,228	5,191	19,419	3,931	3,643	1,548	2,749	7,548
2016	13,829	5,087	18,917	3,709	3,479	1,608	2,803	7,317
2017	13,359	4,986	18,345	3,536	3,343	1,642	2,809	7,014
2018	12,851	4,886	17,737	3,377	3,219	1,667	2,785	6,689
2019	12,337	4,788	17,126	3,221	3,094	1,694	2,771	6,345
2020	11,844	4,692	16,536	3,073	2,973	1,720	2,736	6,034
2021	11,370	4,599	15,969	2,933	2,856	1,743	2,700	5,737
2022	10,915	4,507	15,422	2,799	2,743	1,763	2,654	5,462
2023	10,479	4,416	14,895	2,671	2,636	1,781	2,608	5,200
2024	10,060	4,328	14,388	2,549	2,532	1,796	2,569	4,942
2025	9,657	4,242	13,899	2,432	2,432	1,809	2,529	4,696

Table D-3 (page 8 of 15)
Volume Revenue Elements

Volume Revenue Elements - Scenario 2 (4% Decline) - Case 3 (70% Recovery)

Project Year	Produced Water BPD	BackFlow BPD	Collection Center BPD	Total BR Inj Water BPD	Total BR BF Water BPD	Other BF Water BPD	Other Producer via SWD BPD	Other Producer Satellite BPD
2006	8,855	3,457	12,312	7,595	4,717	0	0	0
2007	12,610	4,886	17,496	8,637	5,710	0	157	2,992
2008	15,338	5,992	21,330	8,375	5,917	76	731	6,232
2009	16,997	6,637	23,635	7,679	5,793	845	1,351	7,967
2010	17,709	6,914	24,623	6,858	5,454	1,460	1,953	8,898
2011	17,824	6,776	24,600	6,173	5,143	1,633	2,423	9,228
2012	17,824	6,640	24,465	5,639	4,881	1,759	2,778	9,407
2013	17,708	6,507	24,216	5,232	4,696	1,811	3,032	9,444
2014	17,478	6,377	23,855	4,981	4,561	1,816	3,187	9,310
2015	17,129	6,250	23,378	4,732	4,386	1,864	3,310	9,087
2016	16,649	6,125	22,774	4,465	4,189	1,936	3,375	8,809
2017	16,083	6,002	22,085	4,257	4,025	1,977	3,382	8,444
2018	15,472	5,882	21,354	4,066	3,875	2,007	3,353	8,053
2019	14,853	5,765	20,617	3,877	3,725	2,039	3,337	7,639
2020	14,259	5,649	19,908	3,700	3,579	2,070	3,294	7,265
2021	13,688	5,536	19,225	3,531	3,438	2,098	3,250	6,907
2022	13,141	5,426	18,566	3,370	3,303	2,123	3,195	6,576
2023	12,615	5,317	17,932	3,215	3,173	2,144	3,140	6,260
2024	12,111	5,211	17,321	3,069	3,048	2,163	3,092	5,950
2025	11,626	5,106	16,733	2,928	2,928	2,178	3,044	5,654

Table D-3 (page 9 of 15)
Volume Revenue Elements

Volume Revenue Elements - Scenario 2 (4% Decline) - Case 4 (80% Recovery)

Project Year	Produced Water BPD	BackFlow BPD	Collection Center BPD	Total BR Inj Water BPD	Total BR BF Water BPD	Other BF Water BPD	Other Producer via SWD BPD	Other Producer Satellite BPD
2006	10,353	4,042	14,395	8,880	5,515	0	0	0
2007	16,047	5,713	21,760	10,742	7,102	0	196	3,721
2008	19,153	7,006	26,159	10,271	7,256	0	906	7,726
2009	20,395	7,761	28,156	9,148	6,901	860	1,631	9,617
2010	20,706	8,084	28,790	8,018	6,377	1,707	2,284	10,404
2011	20,841	7,922	28,763	7,217	6,013	1,909	2,834	10,789
2012	20,841	7,764	28,604	6,593	5,707	2,057	3,249	10,999
2013	20,705	7,609	28,314	6,118	5,491	2,118	3,545	11,043
2014	20,436	7,456	27,892	5,824	5,333	2,123	3,726	10,886
2015	20,027	7,307	27,335	5,533	5,128	2,180	3,870	10,624
2016	19,466	7,161	26,628	5,221	4,897	2,264	3,946	10,299
2017	18,805	7,018	25,823	4,977	4,706	2,312	3,955	9,873
2018	18,090	6,878	24,968	4,754	4,531	2,347	3,921	9,415
2019	17,366	6,740	24,106	4,534	4,356	2,384	3,901	8,932
2020	16,672	6,605	23,277	4,326	4,185	2,421	3,852	8,494
2021	16,005	6,473	22,478	4,129	4,020	2,454	3,800	8,076
2022	15,365	6,344	21,708	3,940	3,862	2,482	3,736	7,689
2023	14,750	6,217	20,967	3,759	3,710	2,507	3,671	7,320
2024	14,160	6,092	20,253	3,588	3,564	2,529	3,616	6,956
2025	13,594	5,971	19,564	3,424	3,424	2,547	3,559	6,610

Table D-3 (page 10 of 15)
Volume Revenue Elements

Volume Revenue Elements - Scenario 2 (4% Decline) - Case 5 (90% Recovery)

Project Year	Produced Water BPD	BackFlow BPD	Collection Center BPD	Total BR Inj Water BPD	Total BR BF Water BPD	Other BF Water BPD	Other Producer via SWD BPD	Other Producer Satellite BPD
2006	11,853	4,628	16,480	10,167	6,314	0	0	0
2007	19,675	7,034	26,709	13,185	8,717	0	240	4,567
2008	22,639	8,468	31,107	12,213	8,628	0	1,078	9,187
2009	23,468	9,024	32,492	10,557	7,964	1,060	1,872	11,039
2010	23,705	9,255	32,960	9,179	7,301	1,954	2,615	11,911
2011	23,859	9,070	32,929	8,263	6,885	2,185	3,244	12,352
2012	23,859	8,889	32,748	7,548	6,534	2,355	3,719	12,593
2013	23,704	8,711	32,415	7,004	6,286	2,424	4,058	12,642
2014	23,396	8,537	31,933	6,668	6,106	2,431	4,266	12,463
2015	22,928	8,366	31,294	6,334	5,870	2,495	4,431	12,163
2016	22,286	8,198	30,485	5,977	5,607	2,592	4,518	11,791
2017	21,529	8,035	29,563	5,698	5,388	2,647	4,527	11,303
2018	20,710	7,874	28,584	5,442	5,187	2,686	4,489	10,779
2019	19,882	7,716	27,598	5,190	4,987	2,730	4,466	10,225
2020	19,087	7,562	26,649	4,953	4,791	2,771	4,410	9,724
2021	18,323	7,411	25,734	4,727	4,602	2,809	4,351	9,246
2022	17,590	7,263	24,853	4,510	4,421	2,841	4,277	8,803
2023	16,887	7,117	24,004	4,304	4,247	2,870	4,203	8,380
2024	16,211	6,975	23,186	4,108	4,080	2,895	4,139	7,964
2025	15,563	6,835	22,398	3,920	3,920	2,916	4,075	7,568

Table D-3 (page 11 of 15)
Volume Revenue Elements

Volume Revenue Elements - Scenario 3 (6% Decline) - Case 1 (50% Recovery)

Project Year	Produced Water BPD	BackFlow BPD	Collection Center BPD	Total BR Inj Water BPD	Total BR BF Water BPD	Other BF Water BPD	Other Producer via SWD BPD	Other Producer Satellite BPD
2006	5,741	2,242	7,983	4,925	3,058	0	0	0
2007	8,176	3,168	11,344	5,600	3,702	0	102	1,940
2008	9,945	3,885	13,830	5,430	3,836	49	474	4,041
2009	11,021	4,304	15,325	4,979	3,756	548	876	5,166
2010	11,483	4,483	15,966	4,446	3,536	947	1,267	5,770
2011	11,483	4,393	15,876	3,984	3,319	1,074	1,560	5,939
2012	11,379	4,305	15,685	3,615	3,129	1,176	1,770	5,994
2013	11,174	4,219	15,394	3,326	2,985	1,234	1,907	5,941
2014	10,873	4,135	15,008	3,134	2,869	1,266	1,973	5,766
2015	10,481	4,052	14,534	2,942	2,726	1,326	2,013	5,526
2016	10,004	3,971	13,976	2,740	2,570	1,401	2,012	5,252
2017	9,464	3,892	13,356	2,574	2,434	1,458	1,970	4,919
2018	8,915	3,814	12,729	2,424	2,310	1,504	1,909	4,583
2019	8,380	3,738	12,118	2,279	2,190	1,548	1,855	4,247
2020	7,877	3,663	11,540	2,145	2,075	1,588	1,789	3,944
2021	7,405	3,590	10,994	2,019	1,966	1,624	1,723	3,662
2022	6,961	3,518	10,478	1,902	1,864	1,654	1,654	3,405
2023	6,543	3,448	9,990	1,791	1,768	1,680	1,587	3,165
2024	6,150	3,379	9,529	1,688	1,677	1,702	1,526	2,936
2025	5,781	3,311	9,092	1,591	1,591	1,720	1,467	2,724

Table D-3 (page 12 of 15)
Volume Revenue Elements

Volume Revenue Elements - Scenario 3 (6% Decline) - Case 2 (60% Recovery)

Project Year	Produced Water BPD	BackFlow BPD	Collection Center BPD	Total BR Inj Water BPD	Total BR BF Water BPD	Other BF Water BPD	Other Producer via SWD BPD	Other Producer Satellite BPD
2006	7,252	2,832	10,084	6,221	3,863	0	0	0
2007	10,328	4,002	14,330	7,074	4,677	0	129	2,450
2008	12,563	4,908	17,471	6,859	4,846	62	599	5,105
2009	13,922	5,436	19,358	6,290	4,745	692	1,107	6,526
2010	14,505	5,663	20,168	5,617	4,467	1,196	1,600	7,288
2011	14,505	5,550	20,055	5,032	4,193	1,357	1,970	7,502
2012	14,374	5,439	19,813	4,567	3,953	1,486	2,236	7,572
2013	14,116	5,330	19,446	4,202	3,771	1,559	2,409	7,505
2014	13,735	5,223	18,958	3,958	3,625	1,599	2,493	7,283
2015	13,240	5,119	18,359	3,716	3,444	1,675	2,543	6,981
2016	12,638	5,017	17,654	3,462	3,247	1,770	2,542	6,634
2017	11,955	4,916	16,871	3,252	3,075	1,841	2,489	6,214
2018	11,262	4,818	16,080	3,062	2,918	1,900	2,411	5,789
2019	10,586	4,722	15,308	2,879	2,766	1,956	2,343	5,364
2020	9,951	4,627	14,578	2,709	2,621	2,006	2,259	4,982
2021	9,354	4,535	13,888	2,551	2,484	2,051	2,177	4,626
2022	8,793	4,444	13,237	2,402	2,355	2,089	2,090	4,301
2023	8,265	4,355	12,620	2,263	2,233	2,122	2,005	3,998
2024	7,769	4,268	12,037	2,133	2,118	2,150	1,928	3,709
2025	7,303	4,183	11,486	2,010	2,010	2,173	1,853	3,440

Table D-3 (page 13 of 15)
Volume Revenue Elements

Volume Revenue Elements - Scenario 3 (6% Decline) - Case 3 (70% Recovery)

Project Year	Produced Water BPD	BackFlow BPD	Collection Center BPD	Total BR Inj Water BPD	Total BR BF Water BPD	Other BF Water BPD	Other Producer via SWD BPD	Other Producer Satellite BPD
2006	8,764	3,422	12,185	7,517	4,668	0	0	0
2007	12,481	4,836	17,316	8,548	5,651	0	156	2,961
2008	15,181	5,931	21,111	8,289	5,856	75	724	6,168
2009	16,823	6,569	23,392	7,600	5,733	836	1,337	7,885
2010	17,527	6,843	24,370	6,787	5,398	1,445	1,933	8,807
2011	17,527	6,706	24,233	6,081	5,066	1,640	2,381	9,066
2012	17,370	6,572	23,942	5,518	4,777	1,795	2,702	9,149
2013	17,057	6,441	23,498	5,077	4,557	1,884	2,911	9,069
2014	16,596	6,312	22,908	4,783	4,380	1,932	3,012	8,801
2015	15,999	6,186	22,184	4,490	4,162	2,024	3,073	8,436
2016	15,271	6,062	21,333	4,183	3,924	2,138	3,071	8,017
2017	14,446	5,941	20,387	3,930	3,716	2,225	3,008	7,509
2018	13,608	5,822	19,430	3,700	3,526	2,296	2,913	6,996
2019	12,792	5,705	18,497	3,479	3,342	2,363	2,831	6,482
2020	12,024	5,591	17,616	3,274	3,167	2,424	2,730	6,020
2021	11,303	5,479	16,782	3,083	3,001	2,478	2,631	5,590
2022	10,625	5,370	15,995	2,903	2,845	2,525	2,525	5,197
2023	9,987	5,262	15,250	2,734	2,698	2,564	2,423	4,831
2024	9,388	5,157	14,545	2,577	2,559	2,598	2,329	4,482
2025	8,825	5,054	13,879	2,429	2,429	2,625	2,239	4,157

Table D-3 (page 14 of 15)
Volume Revenue Elements

Volume Revenue Elements - Scenario 3 (6% Decline) - Case 4 (80% Recovery)

Project Year	Produced Water BPD	BackFlow BPD	Collection Center BPD	Total BR Inj Water BPD	Total BR BF Water BPD	Other BF Water BPD	Other Producer via SWD BPD	Other Producer Satellite BPD
2006	10,274	4,011	14,285	8,812	5,472	0	0	0
2007	15,924	5,669	21,593	10,659	7,047	0	194	3,692
2008	19,006	6,952	25,959	10,192	7,200	0	899	7,667
2009	20,239	7,701	27,940	9,078	6,848	853	1,618	9,543
2010	20,547	8,022	28,569	7,957	6,328	1,694	2,266	10,324
2011	20,547	7,862	28,409	7,129	5,939	1,922	2,791	10,628
2012	20,362	7,704	28,067	6,469	5,600	2,104	3,168	10,726
2013	19,996	7,550	27,546	5,952	5,342	2,208	3,413	10,631
2014	19,456	7,399	26,855	5,607	5,135	2,265	3,531	10,317
2015	18,755	7,251	26,007	5,264	4,879	2,373	3,602	9,889
2016	17,902	7,106	25,008	4,904	4,600	2,507	3,601	9,398
2017	16,935	6,964	23,899	4,607	4,356	2,608	3,526	8,803
2018	15,953	6,825	22,778	4,337	4,134	2,691	3,415	8,201
2019	14,996	6,688	21,684	4,078	3,918	2,770	3,319	7,599
2020	14,096	6,555	20,651	3,838	3,713	2,842	3,201	7,058
2021	13,250	6,423	19,674	3,614	3,518	2,905	3,084	6,553
2022	12,455	6,295	18,750	3,403	3,336	2,959	2,960	6,092
2023	11,708	6,169	17,877	3,205	3,163	3,006	2,840	5,663
2024	11,006	6,046	17,051	3,021	3,000	3,045	2,731	5,254
2025	10,345	5,925	16,270	2,847	2,847	3,078	2,624	4,874

Table D-3 (page 15 of 15)
Volume Revenue Elements

Volume Revenue Elements - Scenario 3 (6% Decline) - Case 5 (90% Recovery)

Project Year	Produced Water BPD	BackFlow BPD	Collection Center BPD	Total BR Inj Water BPD	Total BR BF Water BPD	Other BF Water BPD	Other Producer via SWD BPD	Other Producer Satellite BPD
2006	11,785	4,601	16,386	10,108	6,277	0	0	0
2007	19,563	6,994	26,556	13,109	8,667	0	239	4,541
2008	22,509	8,420	30,929	12,143	8,579	0	1,072	9,135
2009	23,334	8,972	32,306	10,496	7,918	1,054	1,861	10,976
2010	23,570	9,202	32,772	9,127	7,259	1,943	2,600	11,843
2011	23,570	9,018	32,588	8,177	6,813	2,205	3,202	12,191
2012	23,357	8,838	32,195	7,420	6,424	2,414	3,634	12,303
2013	22,937	8,661	31,598	6,827	6,128	2,533	3,915	12,195
2014	22,318	8,488	30,805	6,432	5,890	2,598	4,051	11,835
2015	21,514	8,318	29,832	6,038	5,596	2,722	4,132	11,344
2016	20,535	8,152	28,687	5,625	5,276	2,875	4,130	10,780
2017	19,426	7,988	27,415	5,284	4,996	2,992	4,045	10,098
2018	18,300	7,829	26,128	4,975	4,742	3,087	3,918	9,407
2019	17,202	7,672	24,874	4,678	4,494	3,178	3,807	8,717
2020	16,170	7,519	23,688	4,402	4,259	3,260	3,671	8,096
2021	15,199	7,368	22,568	4,145	4,036	3,333	3,537	7,517
2022	14,287	7,221	21,508	3,903	3,826	3,395	3,396	6,988
2023	13,430	7,077	20,507	3,677	3,629	3,448	3,258	6,496
2024	12,624	6,935	19,559	3,465	3,442	3,493	3,132	6,027
2025	11,867	6,796	18,663	3,266	3,266	3,530	3,010	5,591

Table D-4
PNM Life-of-Project Escalated Operating Costs

PNM Life-of-Project Escalated Operating Costs

Scenario 3 - 6% Declination

Project Year	70% Recovery	80% Recovery	90% Recovery	75% Recovery	85% Recovery
2006	$5,430,140	$5,567,506	$5,704,989	$5,498,823	$5,636,248
2007	$5,616,688	$5,869,172	$6,152,937	$5,742,930	$6,011,055
2008	$5,761,852	$5,996,388	$6,235,877	$5,879,120	$6,116,133
2009	$5,875,885	$6,097,808	$6,312,560	$5,986,847	$6,205,184
2010	$5,966,113	$6,178,661	$6,391,388	$6,072,387	$6,285,024
2011	$6,010,871	$6,226,552	$6,442,416	$6,118,711	$6,334,484
2012	$6,047,726	$6,265,060	$6,482,578	$6,156,393	$6,373,819
2013	$6,076,352	$6,293,796	$6,511,425	$6,185,074	$6,402,610
2014	$6,096,659	$6,312,653	$6,528,831	$6,204,656	$6,420,742
2015	$6,108,807	$6,321,815	$6,535,004	$6,215,311	$6,428,409
2016	$6,112,712	$6,321,179	$6,529,824	$6,216,946	$6,425,502
2017	$6,110,009	$6,312,661	$6,515,484	$6,211,335	$6,414,072
2018	$6,105,642	$6,302,049	$6,498,623	$6,203,845	$6,400,336
2019	$6,101,780	$6,291,885	$6,482,152	$6,196,833	$6,387,018
2020	$6,100,300	$6,284,365	$6,468,586	$6,192,333	$6,376,476
2021	$6,101,162	$6,279,438	$6,457,865	$6,190,300	$6,368,651
2022	$6,104,329	$6,277,055	$6,449,928	$6,190,692	$6,363,491
2023	$6,109,765	$6,277,172	$6,444,722	$6,193,469	$6,360,947
2024	$6,117,441	$6,279,749	$6,442,196	$6,198,595	$6,360,973
2025	$6,127,327	$6,284,749	$6,442,304	$6,206,038	$6,363,526

Table D-5
Life-of-Project Annual Tax Credit

Life-of-Project Annual Tax Credit

	Scenario 3 - 6% Declination		
Project Year	70% Recovery	80% Recovery	90% Recovery
2006	$881,843	$1,017,567	$1,153,405
2007	$1,240,881	$1,553,386	$1,898,486
2008	$1,525,251	$1,828,608	$2,137,802
2009	$1,640,317	$1,931,034	$2,213,254
2010	$1,690,508	$1,965,559	$2,240,844
2011	$1,684,083	$1,958,026	$2,232,203
2012	$1,666,309	$1,937,191	$2,208,302
2013	$1,637,392	$1,903,292	$2,169,417
2014	$1,597,840	$1,856,925	$796,286
2015	$1,548,446	$1,098,412	$0
2016	$1,489,679	$0	$0
2017	$447,453	$0	$0
2018	$0	$0	$0
2019	$0	$0	$0
2020	$0	$0	$0
2021	$0	$0	$0
2022	$0	$0	$0
2023	$0	$0	$0
2024	$0	$0	$0
2025	$0	$0	$0

Table D-6
Life-of-Project Revenue Stream

Life-of-Project Revenue Stream (1)

	Scenario 3 - 6% Declination		
Project Year	70% Recovery	80% Recovery	90% Recovery
2006	$1,680,598	$1,970,153	$2,259,954
2007	$3,217,471	$4,069,988	$5,024,625
2008	$4,841,439	$5,945,112	$7,037,242
2009	$5,664,409	$6,815,281	$7,837,869
2010	$7,500,137	$8,569,176	$9,639,122
2011	$7,812,698	$8,931,282	$10,050,815
2012	$8,026,772	$9,177,849	$10,329,902
2013	$8,144,156	$9,310,982	$10,478,797
2014	$8,157,499	$9,322,062	$10,487,613
2015	$8,096,220	$9,245,575	$10,395,906
2016	$7,960,088	$9,081,249	$10,203,362
2017	$7,751,276	$8,831,629	$9,912,900
2018	$7,516,855	$8,551,896	$9,587,814
2019	$7,281,402	$8,270,857	$9,261,151
2020	$7,053,804	$7,998,929	$8,944,856
2021	$6,836,796	$7,739,317	$8,642,604
2022	$6,629,199	$7,490,638	$8,352,808
2023	$6,431,706	$7,253,700	$8,076,392
2024	$6,244,846	$7,029,122	$7,814,064
2025	$6,067,451	$6,815,535	$7,564,254

Note…..

1. Annual revenue adjusted for backflow charges.

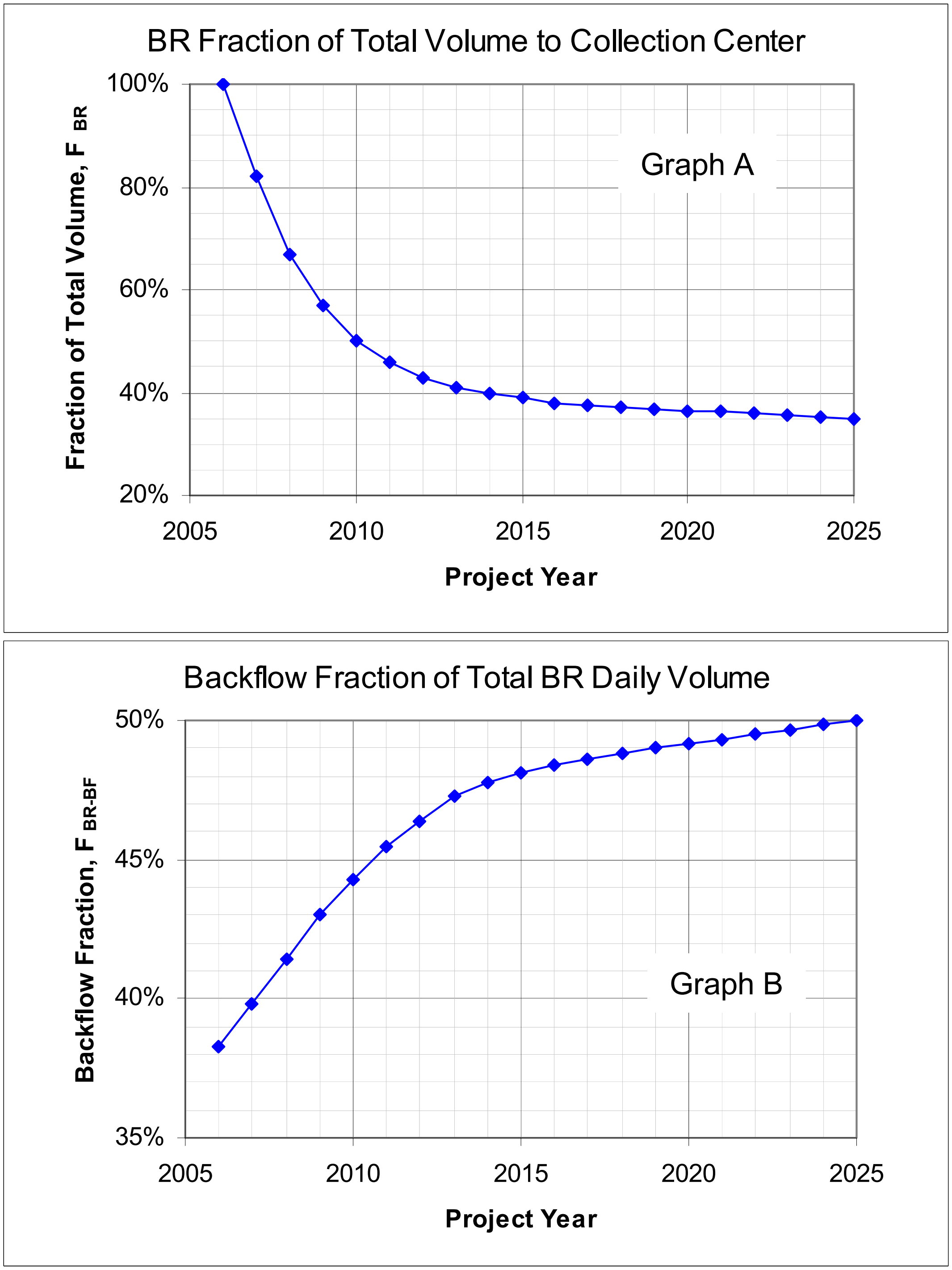
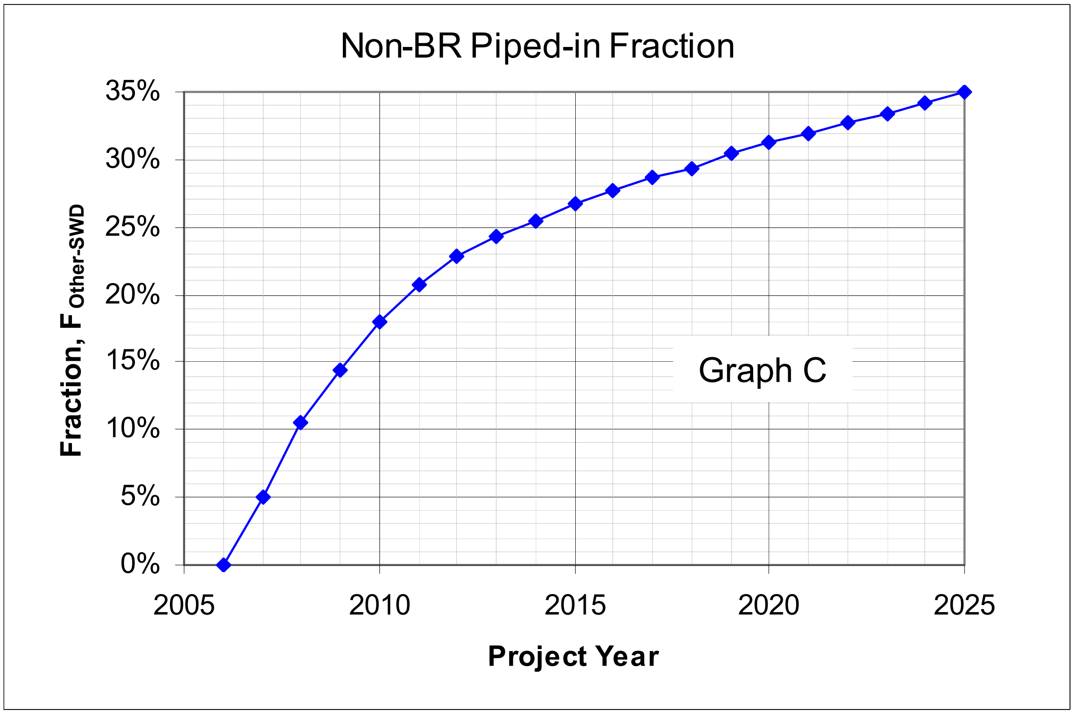


Figure D-1
Collection Center Volume Relationships



Collection Center Parameters

$V_{CC\text{-}BR} = V_{CC} \times F_{BR}$

$V_{CC\text{-}BR} = V_{CC\text{-}BR} \times F_{BR\text{-}BF}$

$V_{CC\text{-}Other\text{-}BF} = V_{CC\text{-}BF} - V_{CC\text{-}BR\text{-}BF}$

$V_{CC\text{-}BR\text{-}PW} = V_{CC\text{-}BR} - V_{CC\text{-}BR\text{-}BF}$

$V_{CC\text{-}Other} = V_{CC} - V_{CC\text{-}BR} - V_{CC\text{-}BF}$

$V_{CC\text{-}Other\text{-}SWD} = V_{CC\text{-}Other} \times F_{Other\text{-}SWD}$

$V_{CC\text{-}Other\text{-}Satellite} = V_{CC\text{-}Other} - V_{CC\text{-}Other\text{-}SWD}$

Where:	V_{CC}	Total Volume, BPD
	F_{BR}	BR Fraction (Graph A)
	$V_{CC\text{-}BF}$	SWD Backflow to Collection Center, BPD
	$V_{CC\text{-}BR}$	BR Volume, BPD
	$F_{BR\text{-}BF}$	BR Backflow Fraction (Graph B)
	$V_{CC\text{-}BR\text{-}BF}$	BR Backflow Volume, BPD
	$V_{CC\text{-}Other\text{-}BF}$	Other Producer Backflow Volume, BPD
	$V_{CC\text{-}BR\text{-}PW}$	BR Produced Water Volume, BPD
	$V_{CC\text{-}Other}$	Other Producer Volume (non-Backflow), BPD
	$F_{Other\text{-}SWD}$	Fraction Other Producers vis SWD to Pipeline (Graph C)
	$V_{CC\text{-}Other\text{-}SWD}$	Other Produced Delivery via SWD to Pipeline
	$V_{CC\text{-}Other\text{-}Satellite}$	Other Producer Delivery via Satellite Station

Figure D-1 (page 2 of 2)
Collection Center Volume Relationships

E

SECTION 9, APPLICABILITY TO OTHER REGIONS IN THE US

Table E-1
Capital and Operating Cost Assumptions

Produced Water Project

Chemical Costs…..

93% $Ca(OH)_2$, $/ton	$86	
98% Na_2CO_3, $/ton	$100	
93% H_2SO_4, $/ton	$91	
50% NaOH, $/ton	$78	Dry basis cost
Other Chemical	15%	of major chemicals

Reactor Clarifier, BC, Crystallizer Op Assumptions.....

RC Sludge Moisture Content	30%	
Crystallizer Solids Moisture Content	50%	
On-Site Sludge & Solids Disposal, $/ton	$25	
HERO Final Reject TDS, mg/l	60,000	or 90% recovery if less than 60,000 mg/l
HERO Operating pH	10.0	
BC Operating pH	10.5	
Excess WAC H_2SO_4	20%	
BC Brine Total Solids, mg/l	225,000	

RO/VCE/Crystallizer cleanings……

	Annual Freq	Cost per Cleaning	
RO	1	$10,000	
VCE	0.66	$30,000	
Crystallizer	1.5	$30,000	
HERO membrane replacement	-----	$180,000	40,000 BPD basis

Equipment Power Requirements…..

HERO System, kwh/kgal	7.0	Feedwater basis - includes 5% allowance for misc power
BC, kwh/kgal	78.1	Distillate basis - includes 2% allowance for misc power
Crystallizer, kwh/kgal	303.7	Feedwater basis - includes 2% allowance for misc power
Power Cost, $/kwh	$0.050	

Labor assumptions…..

Fully Burdened Labor Costs, $/hour	$50
Full Time Coverage, hours/year	8,760

	<40,000 BPD	<80,000 BPD	<100,000 BPD
Operators…..			
HERO/VCE, hours/year	6,240	8,320	10,400
Crystallizer, hours/year	2,080	2,080	2,080
De-Oiling & Pipeline, hours/year	1,040	1,040	1,040
Maintenance & Instrument Techs…..			
HERO/VCE, De-Oiling & Pipeline, hours/year	2,600	2,600	2,600
Crystallizer, hours/year	1,040	1,040	1,040

De-Oiling System…..

Tank Insulation	Yes	
Tank Heaters	Yes	
Off-Spec Water Fraction	0.2%	of daily in-flow
Off-Spec Water Hauling Cost, $/bbl	$1.00	
Off-Spec Water Disposal Cost, $/bbl	$6.50	
Credit Taken for Recovered Oil	None	

Pipeline…..

Unit Pipeline Cost, $/inch-Dfoot	$7.50
Pipeline Material	HDPE
Pipeline Max Operating Pressure, psi	150
Pipeline Pump Stations	0
Route Type	City/Open Country
Terrain Type	Flat

Evaporation Ponds…..

Evap Pond Installed Cost, $/acre	$200,000	
Annual Avg Evap Rate, gpm/acre	2.0	Equivalent to ~40" net evaporation per year

Installation Cost Factor.....	45%	of process equipment equipment costs
De-Oiling, HERO Eqpmt Maintenance Cost	2.0%	of equipment costs
Evap Pond Maintenance Cost	0.5%	of evaporation pond cost
Pipeline Maintenance Cost	1.5%	of installed cost
Capital Cost Contingency	25%	of equipment costs