# THE DESIGN OF A ZERO-EFFLUENT DISCHARGE SYSTEM FOR WESTCOAST ENERGY INC.'S FORT NELSON GAS PLANT 

by

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#### Abstract

This project was initiated by Westcoast Energy Inc. (Westcoast) to examine the feasibility of transforming their Fort Nelson Gas Plant (FNGP) into a zero-effluent discharge (ZED) facility. Water flow and water chemistry data were collected at the plant. The resulting data set was used to identify water leaks in the existing distribution network, as well as to identify methods of optimizing water use at this facility. Designs for implementing a ZED protocol at the FNGP were then developed and subsequently evaluated with a computer simulator.

A total of 18 ZED models were constructed using reverse osmosis (RO) membranes, nanofilters and other ZED technology. The final disposal mechanism in 10 of the 18 scenarios was a brine concentrator - spray dryer (BCS) assembly, which reduced all remaining wastewater into a solid waste. In the other 8 ZED designs, final concentrates were disposed of in a deep well. The best deep well configuration was a 1 -stage RO filter; the necessary equipment, excluding the deep well, would cost an estimated $\$ 101700$ US dollars. The best BCS scenarios were a 1 -stage RO and a BCS-only models; they would cost around $\$ 1.61$ and $\$ 1.79$ million US dollars, respectively. A step by step approach by which the FNGP can be transformed into a ZED facility is detailed in Chapter 8.0 of this report.

While it is possible to implement a ZED program at the FNGP, there are consequences to this course of action which need to be considered. The most important is final waste management. Regardless of whether a deep well or a BCS unit is used, either system will have to be built and operated in such a way that the final waste products produced by the ZED treatment train do not migrate off-site, as this action would violate the ZED principle. Once a ZED program is initiated, wastewater and/or waste solids produced at the FNGP will no longer disappear with the FNR; they will remain on-site indefinitely.


## TABLE OF CONTENTS

Abstract ..... ii
List of tables ..... vi
List of figures ..... viii
Acknowledgements ..... ix
1.0 INTRODUCTION .....  1
2.0 BACKGROUND
2.1 Gas Plant Operations
2.1.1 General ..... 7
2.1.2 Fort Nelson Gas Plant ..... 7
2.2 FNGP Water System
2.2.1 Liquid phase ..... 8
2.2.2 Vapour phase ..... 11
3.0 AVAILABLE RECYCLING TECHNOLOGY
3.1 Distillation ..... 18
3.2 Ion-exchange ..... 18
3.3 Membrane Filtration
3.3.1 Ion selective separation ..... 19
3.3.2 Water permeable membranes ..... 21
3.3.3 Reverse osmosis vs. electrodialysis ..... 23
3.3.4 RO membrane selection ..... 32
3.4 Brine Disposal ..... 33
3.5 Conclusion ..... 35
4.0 METHODOLOGY
4.1 Analysis of the Existing Water Distribution System
4.1.1 Flow data ..... 42
4.1.2 Chemical data ..... 42
4.2 Selecting the Appropriate Recycling Technology for the FNGP ..... 44
4.3 The Computer Simulator
4.3.1 Basic construction ..... 45
4.3.2 ZED components ..... 46
4.3.3 Parameter values ..... 47
5.0 ANALYSIS OF THE EXISTING WATER SYSTEM
5.1 Results
5.1.1 Water balances ..... 50
5.1.2 Mass balances ..... 50

## TOC (con't)

5.0 ANALYSIS OF THE EXISTING WATER SYSTEM (con't)
5.2 Discussion
5.2.1 Data quality ..... 51
5.2.2 System optimization ..... 54
5.3 Changing to a ZED System ..... 59
5.4 Conclusion ..... 60
6.0 RECYCLING TECHNOLOGY FOR THE FNGP
6.1 Re-evaluation of the Literature Review ..... 88
6.2 Available Options ..... 89
6.3 Conceptual ZED Designs
6.3.1 Back-end models ..... 91
6.3.2 Composite discharge designs ..... 92
7.0 DESIGN EVALUATION
7.1 Simulator Performance ..... 95
7.2 BCS Designs
7.2.1 Composite discharge models ..... 96
7.2.2 Back-end solutions: ion-exchange vs. nanofiltration ..... 96
7.2.3 Best of the back-end ion-exchangers ..... 98
7.2.4 Conclusion ..... 101
7.3 Deep Well Configurations
7.3.1 Composite discharge designs ..... 102
7.3.2 Back-end solutions: ion-exchange vs. nanofiltration ..... 102
7.3.3 Best of the back-end ion-exchangers ..... 102
7.3.4 Conclusion ..... 103
7.4 Evaluation Summary ..... 104
8.0 ZED IMPLEMENTATION AND IMPLICATIONS ..... 125
9.0 POTENTIAL LIMITATIONS
9.1 Non-Representative numbers
9.1.1 Flow data ..... 130
9.1.2 Chemical data ..... 130
9.2 Sample Variability
9.2.1 Water data ..... 131
9.2.2 Chemical data ..... 131
9.3 Computer Model ..... 132
9.4 ZED Evaluation Process ..... 133
9.5 Conclusion ..... 134

## TOC (con't)

10.0 CONCLUSIONS ..... 136
11.0 RECOMMENDATIONS ..... 138
References ..... 139
Appendix A - Water flow data ..... 146
Appendix B - Water balance equations and assumptions ..... 186
Appendix C - Water chemistry data ..... 194
Appendix D - Mass balance equations and assumptions ..... 207
Appendix E-Equations used in the computer simulator ..... 213
Appendix F - Simulator summary sheets ..... 268
Appendix G - Cost estimates ..... 278

## LIST OF TABLES

Table 3.1: A comparison of the inherent advantages and disadvantages of reverse osmosis (RO), electrodialysis (ED) and electrodialysis reversal (EDR) ..... 36
Table 5.1: A water balance of the Fort Nelson Gas Plant based on its individual operating units ..... 63
Table 5.2: A water balance of the Fort Nelson Gas Plant based on the different types of water used at the facility ..... 70
Table 5.3 A summary of treated water losses from the Fort Nelson Gas Plant. ..... 74
Table 5.4: A mass balance on the front-end softening system ..... 75
Table 5.5: A mass balance on the lime ponds ..... 76
Table 5.6: A mass balance on the boilers ..... 77
Table 5.7: A mass balance on the sulphur plant ..... 78
Table 5.8: A mass balance on the polishing pond ..... 79
Table 5.9: A comparison between water flow readings from January 1995 and overall averages ..... 80
Table 5.10: Water balances on the raw water storage tank and front-end softeners (FESs) using metered and calculated inflow and outflow volumes from the FESs ..... 82
Table 5.11: Recalculated mass balance on the FNGP's front-end softening system ..... 83
Table 5.12: Recalculated mass balance on the FNGP's lime ponds ..... 84
Table 5.13: Changes that occurred in selected areas of the Fort Nelson Gas Plant's water system before and after water observed to be escaping from the effluent treatment plant was recovered ..... 85
Table 5.14: Selected water flows in the Fort Nelson Gas Plant's water distribution network with and without 8 psi steam reuse ..... 86
Table 5.15: The average inlet gas profile for January 1994 at the Fort Nelson Gas Plant ..... 87
Table 6.1: A chemical comparison of the treated water and Fort Nelson River (FNR) discharge flows. ..... 93
Table 7.1: Comparison of 10 selected BCS ZED designs ..... 106
Table 7.2 Changes in RO and B/C feedwater flows, calcium pretreatment demands and product water quality, in the 1 -stage RO, 2-stage RO and BCS-only back-end, BCS, ion-exchange configurations, triggered by given alterations in raw water and effluent plant outflow chemistry ..... 108

## LIST OF TABLES (con't)


Table 7.4: Comparison of 8 selected deep well ZED designs113

Table 7.5: Changes in RO feedwater flow, calcium pretreatment demands and product water quality, in the 1 -stage and 2-stage RO back-end, deep well, ion-exchange configurations, triggered by given alterations in raw water and effluent plant outflow chemistry115

Table 7.6: Changes in treated water quality, in both the 1 -stage and 2-stage RO back-end,
deep well, ion-exchange configurations, triggered by given alterations in raw
water and effluent plant outflow chemistry ..... 117

Table 9.1: Changes to the configuration of the back-end, BCS, 1-stage RO, ion-exchange
design with various flow alterations ..... 135

## LIST OF FIGURES

Figure 1.1: Water balance at the Fort Nelson Gas Plant before and after ZED implementation .....  5
Figure 1.2: Salt balance at the Fort Nelson Gas Plant before and after ZED implementation. ..... 6
Figure 2.1: An illustration of Westcoast Energy Inc.'s operations in British Columbia ..... 13
Figure 2.2: General layout of the Fort Nelson Gas Plant ..... 14
Figure 2.3: An illustration of gas processing flows at the Fort Nelson Gas Plant ..... 15
Figure 2.4: A simplified illustration of the Fort Nelson Gas Plant's water system ..... 16
Figure 2.5: A simplified illustration of the Fort Nelson Gas Plant's steam system ..... 17
Figure 3.1: Typical operating total dissolved solid (TDS) concentrations for different desalination technologies ..... 37
Figure 3.2: An illustration of the electrodialysis process. ..... 38
Figure 3.3a: Flows through an EDR stack when electricity travels from right to left ..... 39
Figure 3.3b: Flows through an EDR stack when electricity travels from left to right ..... 39
Figure 3.4: Pore size distribution in three types of water permeable membranes ..... 40
Figure 3.5: An illustration of reverse osmosis ( RO ) filtration ..... 41
Figure 4.1: A simplified diagram of the water distribution system at the Fort Nelson Gas Plant ..... 49
Figure 6.1: A graphic representation of the available ZED options ..... 94
Figure 7.1: An illustration of the projected flow patterns in a composite discharge, reverse osmosis ZED configuration ..... 119
Figure 7.2: An illustration of the changes in equipment and flow patterns in a given ZED system using nanofiltration instead of ion-exchange softening. ..... 120
Figure 7.3: An illustration of the projected flow patterns in a back-end, BCS-only ZED configuration ..... 121
Figure 7.4: An illustration of the projected flow patterns in a back-end, reverse osmosis ZED configuration ..... 122
Figure 7.5: A simplified illustration of a 1 -stage RO, BCS ZED system. ..... 123
Figure 7.6: A simplified illustration of a 2-stage RO, BCS ZED system. ..... 124

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Finally, using words borrowed from Nelson Mandela, "...I have discovered the secret that after climbing a great hill, one only finds that there are many more hills to climb. I have taken a moment here to rest, to steal a view of the glorious vista that surrounds me, to look back on the distance I have come. But I can rest only for a moment ... for my long walk is not yet ended."

### 1.0 INTRODUCTION

A zero-effluent discharge (ZED) facility, as defined by some, should not release any waste products into the environment. There would be no wastewater discharges, no atmospheric emissions and no solid waste output. The plant would essentially exist as closed system. There are few such facilities in existence, due to the cost, complexity and possible impracticality of a completely closed system. ZED is, as a result, more commonly used to describe a facility which does not release any wastewater into its surroundings.

Incentives for implementing a ZED program can include a reduction in overall operating costs, improved environmental quality and/or regulatory compliance. Westcoast Energy Inc. (Westcoast) initiated the ZED project as a potential alternative to upgrading the current effluent treatment facilities at its Fort Nelson Gas Plant (FNGP), as will be required by stricter environmental guidelines. By eliminating the wastewater discharge stream now released to the Fort Nelson River (FNR), the FNGP will be able to operate independent of British Columbia's wastewater legislation.

The feasibility of implementing a ZED system at the FNGP is, however, bound by two initial conditions:

1) Westcoast has stipulated that while domestic wastewaters are part of the FNR discharge and will, as a result, have to be included in any ZED program, no recycled water will enter the domestic system; it will continue to draw fresh water from the plant's raw water reservoir.

## and

2) Most of the water used in the operational sections of the plant is softened water.

By condition (1), domestic wastewater will become a source of water for the plant (Fig. 1.1). If operational water losses do not exceed domestic wastewater input, the ZED project is doomed to
fail. The surplus of domestic wastewater would necessitate a liquid discharge from this facility - a violation of the founding principle of a ZED system. It will therefore be impossible to eliminate the FNR discharge if the plant's domestic wastewater production proves to be greater than its operational water losses, so long as recycled water is not used in the domestic system.

While boundary condition (2) does not affect the FNGP's water balance, it will still have profound effects on the ZED project. All of the chemical contaminants stripped from raw water now used in the operational sections of this facility are currently discharged along with the plant's wastewater to the FNR. When this wastewater is recycled, as will be the case in a ZED program, these elements will be returned to the head of operations. Here, they will be joined by additional salts contained in the raw and domestic waters needed to replace process water losses. To maintain a chemical balance, some salts and other contaminants will have to be drained from the ZED treatment train (Fig. 1.2). Otherwise, these materials will accumulate within the water distribution network with destructive results. Although adopting a ZED framework will release the FNGP from some regulatory guidelines, the ability to implement such a program may be limited by both the size of this facility's domestic wastewater output and the continued production of a salt brine or solid waste from any ZED treatment system.

Optimizing the FNGP's water distribution network is likely to increase the viability of a ZED program, as it will minimize the limitations imposed on this project by both boundary conditions. For example, the amount of brine or solid waste ultimately produced by a ZED treatment train is directly proportional to the volume of softened water lost from the plant. Tightening up the distribution system and reducing process water losses will shrink the demand for make-up water. Less raw water will then need to be imported from the raw water reservoir and fed into the frontend softeners (FESs). As feedwater volumes to the FESs drop, so to will the mass of contaminants discharged from these vessels. Minimizing water loss in the operational sections of the plant will therefore limit ZED waste output by reducing incoming raw water flow and waste generation rates in the FESs.

Increasing the efficiency of the domestic system is similarly important to the success of the ZED project. As previously stated, domestic wastewater production rates must be smaller than operational water losses as a prerequisite to initiating a ZED program. On one hand, this condition restricts process system optimization, in that one does not want to reduce process water losses to the point where they are less than the domestic output. At the same time, it also implies that minimizing domestic wastewater production will theoretically permit a greater amount of process optimization and increase the probability of fulfilling the prerequisite flow regime (i.e. domestic wastewater production < operational water losses).

In addition to these benefits, network optimization, on the whole, will help to minimize the cost and complexity of the required ZED treatment train. Reducing process water losses will, as previously stated, reduce the mass of contaminants contained in the FNR discharge by limiting their production in the FESs. Similarly, reusing relatively uncontaminated wastewaters, instead of releasing them into the effluent system, will reduce the volume of FNR discharge. Since the goal of the ZED project is to eliminate the release of wastewater to the FNR, the FNGP's discharge stream will ultimately need to be cleaned and recycled back into the plant. Reducing the volume and level of contamination of this wastewater, as achieved through system optimization, can only help to simplify this task. Increasing the efficiency of the existing water network is, therefore, a key step in establishing a cost effective ZED program at the FNGP; it will lead to a smaller, less contaminated final plant outflow which will be easier to recycle and produce less final solid waste than the current FNR discharge stream.

In order to identify if operational water losses at the plant are greater than its domestic output, as well as where water is now lost from the facility, flow diagrams and water balances were constructed. Plant waters were then analyzed to determine which of the escaping streams could be directly reused and which others would require pretreatment before reuse. Conceptual ZED designs, which detail the configuration of the FNGP as a ZED facility, were developed by combining information from available literature with the assembled flow diagrams and collected
water chemistry data. Computer modeling was used to evaluate the efficiency of each design. The process by which all of this work occurred is discussed in the following report, starting with a brief description of the FNGP, a review of the current literature and an outline of the methodology used in this study. Later chapters detail an analysis of the plant's existing water distribution network, the appropriate recycling technology and the best ZED configurations available to this facility. How a ZED program should be implemented, together with this project's potential limitation, conclusions and recommendations, are all examined in the final sections of this document.

## Before ZED implementation



## After ZED implementation



Figure 1.1: Water balance at the Fort Nelson Gas Plant before and after ZED implementation.

## Before ZED implementation



## After ZED implementation



Figure 1.2: Salt balance at the Fort Nelson Gas Plant before and after ZED implementation

### 2.0 BACKGROUND

### 2.1 Gas Plant Operations

### 2.1.1 General

Natural gas reserves are rarely pure methane. They generally c̣ontain a mixture of water, carbon dioxide, hydrogen sulphide and methane (Medici 1974, Ikoku 1984). Natural gas containing high levels of carbon dioxide and hydrogen sulfide is referred to as "sour" or "high acid" gas (Medici 1974, Ikoku 1984). Conversely, "sweet" gas is virtually free of pollutants (Medici 1974, Ikoku 1984). Normally raw field or production gas is dehydrated close to the well head. The remaining contaminants are removed downstream at processing plants. Production gas may also contain various heavier hydrocarbons, including ethane, propane, and butane; such gas being referred to as wet gas (as opposed to dry gas) (Medici 1974, Ikoku 1984). Processing facilities remove these liquids in addition to other undesirables, producing dry sweet sales gas.

In British Columbia, most natural gas processing is undertaken at 5 major plants, including the FNGP (Fig. 2.1), all of which are owned and operated by Westcoast Energy Inc.. As shown in Figure 2.1, field production is moved through a myriad of gas gathering pipelines to these 5 major processing facilities, and sales gas is then transported via Westcoast's major trunk pipeline system to markets in British Columbia and the U.S.

### 2.1.2 Fort Nelson Gas Plant

As field gas enters the FNGP, which itself is illustrated in Figure 2.2, it first passes through a liquid separator where liquid hydrocarbons and the remaining water are isolated from the gas (Fig. 2.3). At this point, the gas stream typically contains $87 \%$ methane, $12 \%$ carbon dioxide and $1 \%$ hydrogen sulphide by volume (E/F Manual). The carbon dioxide and hydrogen sulphide are stripped from the gas using either diethanolamine (DEA), monoethanolamine (MEA), potassium carbonate $\left(\mathrm{KCO}_{3}\right)$ or combination thereof ( $\mathrm{E} / \mathrm{F}$ and $\mathrm{C} / \mathrm{D}$ Manuals). The "scrubbed" gas, although free of carbon dioxide and hydrogen sulphide, is now saturated with water from the stripping
solutions. This water is removed in dehydrators, and the sweet dry gas is released into the sales pipeline. Water trapped by the dehydrators drains to the flare pits for disposal.

The DEA, MEA and $\mathrm{KCO}_{3}$ solutions are not used on a "once-through" basis. Contaminated liquids are regenerated by steam cleaning ( $\mathrm{E} / \mathrm{F}$ and $\mathrm{C} / \mathrm{D}$ Manuals). Carbon dioxide and hydrogen sulphide are leached out of the spent stripping solutions by counter-current steam flows (Fig. 2.3). The contaminated steam exits the gas processing trains and travels to a sulphur recovery unit.

Hydrogen sulphide is a poisonous gas which cannot, under current government regulations, be freely released into the environment. Hydrogen sulphide extracted from sour gas is changed into elemental sulphur by reacting it with sulphur dioxide. This transformation, known as the "Claus" process, typically recovers $98 \%$ of the incoming hydrogen sulphide (S/P Manual). Remaining hydrogen sulphide is burnt to sulfur dioxide and released along with all of the carbon dioxide and water contained in the vapor stream evolved when regenerating the DEA, MEA and $\mathrm{KCO}_{3}$ solutions (Fig. 2.3).

In simple terms, gas processing at the FNGP consists primarily of stripping liquid and gaseous contaminants from incoming sour gas to produce sweet sales gas, the conversion of hydrogen sulphide to elemental sulphur, and the atmospheric release of carbon dioxide, low levels of sulphur dioxide and steam.

### 2.2 FNGP Water System

Water at the FNGP is used for processing and power generation, as well as for cleaning and domestic purposes. Depending on temperature and pressure, it exists as either a liquid or a gas.

### 2.2.1 Liquid Phase

All of the water required by the plant originates from either the Fort Nelson River (FNR) or Burger creek (a minor water flow used in the spring and summer months). Water from either source initially travels to the raw water reservoir. From the reservoir, it is pumped, as required, to
a storage tank closer to the facility (Fig. 2.4). Water passes into the plant in one of three different flow systems:
i) Cleaning water is drawn directly from the tank and travels through the plant's "raw water" piping.
ii) Water destined for domestic use moves from the tanks through two activated carbon filters; it is then chlorinated and dispersed throughout the plant in the domestic system.
iii) The remaining water by-passes the domestic treatment train and flows into a water softening system; it is henceforth called treated water.

Dirty cleaning and domestic waters are released into the effluent system, which transports them to the effluent treatment plant (Fig. 2.4). All of the surface drains within the FNGP also drain to the effluent facility. Rainwater, collected in a series of ditches and culverts, does not enter the plant's water network. It is directed off-site, unless it has been contaminated by an on-site spill (Plant Modifications 1994). If rainwater does become contaminated, then it is pumped to the effluent plant for treatment. Similarly, polluted groundwater around the facility is treated at the effluent plant (Plant Modifications 1994). In general, very little "outside" water (i.e. precipitation and/or groundwater) enters the FNGP's water network, and cleaning \& domestic waters travel through separated pipelines until they mix in the effluent system.

Treated water produced by the front-end softeners (FESs) similarly remains isolated from the other water systems. Most treated water is used as boiler-feed make-up or dilutant for the natural gas stripping solutions (Fig. 2.4). Remaining treated water is used to regenerate the FESs, backwash the domestic filters or for cleaning equipment too sensitive to be washed with raw water. Spent regeneration and backwash waters drain to the lime ponds, while dirty cleaning water goes to the effluent plant.

Four of the 6 natural gas processing trains at the FNGP use treated water to replace water lost from the natural gas stripping solutions (i.e. regeneration steam and flare water - Fig. 2.3) (E/F Manual). The other 2 processing trains replace their lost water with water from the condensate return line (Fig. 2.4). Regardless of their origins, waters headed to "Process" are, for the most part, lost from the water network upon arrival. They are either eventually released to the atmosphere via the thermo-oxidizer or vapourized in the flare pits (Fig. 2.4). A small volume of flare water may remain in the pits not having evaporated during flaring. This residue will drain into the effluent treatment plant. In general, water which comes into direct contact with natural gas does not cycle into any other part of the water network, except from the small volumes of flare water draining to the effluent treatment facility. Contaminants carried into the plant in the gas stream can, however, be transported down to the effluent plant in waters used to clean process vessels.

Boiler feedwater (BFW) is a combination of treated water and returned condensate (Fig. 2.4). It is pumped into the 3 boilers and various vessels within the sulphur plant. A small volume of BFW drains from the boilers and the sulphur plant to control contaminant concentrations within the steam system, while the remaining water is vapourized and released into the steam system. Blowdown waters collect in the lime ponds.

The lime ponds receive wastewater from, as indicated, the boilers and the sulphur plant, as well as from the domestic filters, the FESs and several treated water drains within the FNGP (Fig. 2.4). The effluent plant, on the other hand, mainly receives domestic sewage, dirty cleaning water and a tiny amount of water from the flare pits. While the lime ponds are nothing more than collection basins, the effluent plant is comprised of a 20 -day stabilization pond, a 5 -day activated sludge tank and 3 clarifiers (E/P Manual). Lime pond and effluent plant outflows are combined in a mixing tank, flow through a 1-day retention pond (referred to as the "polishing" pond), and are released to the FNR.

Raw water entering the FNGP is either directly used for cleaning, is filtered, chlorinated and used for domestic purposes, or is softened and called treated water. Spent cleaning and domestic waters are collected, treated and released to the FNR. Some treated water, following its use as either cleaning, regenerating or blowdown water, is also collected and discharged to the FNR. The rest of the treated water appears to be lost through the flare pits and thermo-oxidizer.

### 2.2.2 Vapour Phase

Some of the treated water produced by the FESs mixes, as previously indicated, with condensate, forming BFW. Most of the BFW is pumped into the boilers, although some of it goes into the sulphur plant (Fig. 2.4). All of the BFW is transformed into steam, regardless of its destination.

The boilers produce 450 psi steam, which is used to power electrical turbines, high pressure pumps and gas blowers (Fig. 2.5). BFW going to the sulphur plant changes to either 150, 45 or 15 psi steam, depending on where it was vapourized in the sulphur recovery train. Petrosul, a neighboring firm, purchases a small quantity of 150 psi steam. The rest of it is used to drive the sulphur plant's air intake blowers.

Gas blowers, electrical turbines and high pressure pumps all release 45 psi steam (Fig. 2.5). Combined with the 45 psi steam originally created in the sulphur plant, most of this water vapour travels to process train reboilers. Here it heats clean natural gas stripping solutions to their boiling point. Steam produced from the boiling stripping liquids is used for self-cleaning (i.e. steam from uncontaminated solution is used to clean spent liquor) and released to the sulphur plant. Treated water and/or condensate is imported into "Process", as previously described, to replace these water losses. With its heat energy transferred to the stripping liquors, the 45 psi steam condenses and leaves the reboilers as a liquid.

Forty-five psi steam not destined for process equipment is used for heating elsewhere in the plant. It eventually cools and condenses. Condensed steam is collected and returned to the BFW
tank (Fig. 2.5). The only exception is a small amount of 45 psi steam blown through the deaerators, wherein its pressure drops to 8 psi. The emerging 8 psi steam is vented to the atmosphere.

Five and 150 psi steam created in the sulphur plant are used therein and collected as condensate (Fig. 2.5). It is similarly returned to the head to the plant with the rest of the condensate.

The steam system at the FNGP is already quite efficient. As each steam loop remains isolated within the natural gas processing and sulphur plant equipment, there is no opportunity for contamination of these high quality waters. As a result, most of the water pumped through the steam system is reused. The need for BFW make-up, however, indicates the presence of steam leaks within the facility, possibly beyond the boiler and sulphur plant blowdowns and Petrosul outlet previously identified.

Overall, water, existing as either a liquid or a gas, is used for a number of purposes at the FNGP, ranging from cleaning to power generation. A large proportion of water in the system is reused. Never the less, the demand of treated water make-up suggests that there are leaks in the operational water loops. All of the missing water may be escaping through the thermo-oxidizer, the flare pits and the steam outlet to Petrosul. This premise can, however, only be evaluated with a more thorough investigation of the plant's water network, hence the inclusion of a system diagnostic in the ZED project.


Figure 2.1: An illustration of Westcoast Energy Inc.'s operations in British Columbia

## $-$ <br> 


Effluent Plant

## Polishing Pond


$=-1-$

Figure 2.2: General layout of the Fort Nelson Gas Plant


Figure 2.3: An illustration of gas processing flows at the Fort Nelson Gas Plant.


Figure 2.4: A simplified illustration of the Fort Nelson Gas Plant's water system.


Figure 2.5: A simplified illustration of the Fort Nelson Gas Plant's steam system.

### 3.0 AVAILABLE RECYCLING TECHNOLOGY

Reusing process wastewaters and/or secondary domestic effluent is not an uncommon practice, even at power plants and other steam generating facilities (e.g. Abdullaev et. al. 1992, Pankratz \& Johanson 1992, Bowlin \& Ludlum 1992, Pierce \& Sbei 1993, Strauss 1994, Strauss 1995). The recycling systems tend to involve one or more of the following principles: distillation (Brew \& Blackwell 1991, Bowlin \& Ludlum 1992, Strauss 1994), ion exchange (Kalinske et. al. 1979, Egozy et. al. 1980, Abdullaev et. al. 1992, Bowlin \& Ludlum 1992), and/or membrane filtration (Pankratz \& Johanson 1992, Bowlin \& Ludlum 1992, Strauss 1994, Strauss 1995).

### 3.1 Distillation

Distillation uses heat energy to produce high quality steam from the contaminated wastewater (Kalinske et. al. 1979, Wood 1987, Parekh 1991). The steam can either be reused directly, or first cooled to a liquid. The volume of the remaining waste stream is significantly reduced, if not completely transformed to a solid. This relatively energy intensive process is generally most efficient and cost-effective for waters with total dissolved solid (TDS) levels above $10000 \mathrm{mg} / \mathrm{L}$ (Fig. 3.1).

### 3.2 Ion-exchange

Ion-selective resins are the backbone of the ion exchange process (Kalinske et. al. 1979, Hill \& Lorch 1987, Parekh 1991). These resins, or beads, are coated with relatively inert ions. As wastewater travels through an ion exchanger, unwanted ionic species are adsorbed, and replaced with more process friendly ions. The influent water is now clean and ready for reuse. The resins themselves eventually become saturated with contaminants. They are then regenerated, and the process starts anew. The optimal TDS range for ion exchanger is generally between 100 and 800 mg/L (Fig. 3.1).

### 3.3 Membrane Filtration

Unlike either of the previous techniques, membrane filtration involves neither heat energy nor ion replacement. As the name implies, undesirable elements in the wastewater are isolated using membrane filters (Kalinske et. al. 1979, Applegate 1984, AWWA 1989, Huang \& Koseoglu 1993). The configuration of the filter determines what elements are removed, and which others are not (Applegate 1984, AWWA 1989, Cartwright 1991). Membrane filtration has been found to work best with wastewater TDS concentrations between 100 and $10000 \mathrm{mg} / \mathrm{L}$ (Fig. 3.1).

There are, however, two general classes of membrane filters; those that are ion selective, and others which are water permeable (Kalinske et. al. 1979, Applegate 1984, AWWA 1989, Cartwright 1991, Huang \& Koseoglu 1993). Due to the specificity of their membranes, the two processes are fundamentally different from one another. Each one has its own inherent advantages and disadvantages.

### 3.3.1 Ion selective separation

Ion selective separation, more commonly referred to as electrodialysis (ED), uses filters which are designed to allow either only cation or only anion passage through the membrane (Kalinske et. al. 1979, Applegate 1984, Solt \& Foster 1987, AWWA 1989). In its most basic application, cationic and anionic filters are alternately stacked between an anode and a cathode (Fig. 3.2). Each cationic/anionic group is called a cell, and contained within each cell are two open areas bounded by the membranes. Water, flowing parallel to the filters, fills these spaces. When a direct current is applied to the system, cations and anions in solution move toward the cathode and anode, respectively (Fig. 3.2). With alternating cationic/anionic filters, water in one compartment will be drained of its ionic contaminants, while water in the neighbouring space will become increasingly contaminated. The clean and dirty waters are then collected separately as they exit the cells.

Several adaptations have been made to the basic ED configuration in attempts to either increase the robustness of the system or to improve effluent water quality. Sealed-cell electrodialysis,
which involves sealing each cationic/anionic membrane pair in a "bag", has shown some promise in bench-scale experiments (Schoeman \& van Staden 1991). It is, however, harder to clean and maintain, as compared to the "un-sealed" ED process described above (Schoeman \& van Staden 1991).

Similarly, continuous deionization, which includes ion exchange resins in the clean water compartments of an ED stack, has been shown to produce very high quality effluent (Parekh 1991, Ganzi et. al. 1992). It also appears to be rather complex and potentially difficult to operate, as well as expensive to build (Ganzi et. al. 1992). Given the drawbacks inherent in both systems, neither sealed-cell ED, nor continuous deionization, appears to be a better option than the simpler, unaltered ED stack.

One variation on the basic ED configuration which has significant benefits is electrodialysis reversal (EDR) (Applegate 1984, Schoeman 1985, Solt \& Foster 1987, AWWA 1989). EDR is almost identical to ED. The only alteration has been the installation of transformable cathodes and anodes. Rather than having the electrical current always travel in the same direction through the stacked filters (as it does in ED), EDR uses an alternating current. The relative position of the anode and cathode are then dependent on the direction of electron flow (Figs. 3.3a \& b). As the locations of anode and cathode change, so too does the direction of travel for ions in solution. Cations and anions previously moving one way are now traveling in the opposite directions (Figs. $3.3 \mathrm{a} \& \mathrm{~b}$ ). Similarly, the compartments previously producing clean water now contain more contaminated water. The great advantage of this system is that it is "self-cleaning" (Applegate 1984, Schoeman 1985, Solt \& Foster 1987, AWWA 1989).

While operating under a given current, ions concentrate in certain areas of the ED or EDR filters, or stacks as they are more commonly known (Figs. 3.2 \& 3.3a). Certain mineral salts, such as calcium carbonate $\left(\mathrm{CaCO}_{3}\right)$ or calcium sulfate $\left(\mathrm{CaSO}_{4}\right)$ have limited solubilities in water (Kotz \& Purcell 1987). If those solubilities are exceeded, then salts will start to precipitate. For example, if, as calcium and carbonate accumulate in the "contaminated" waters in the ED or EDR
stack, their combined ionic concentration is greater than the $\mathrm{CaCO}_{3}$ solubility limit, then $\mathrm{CaCO}_{3}$ will begin to precipitate out of solution and onto the surrounding membranes. Precipitation will continue until the combined abundance of calcium and carbonate ions is equivalent to the $\mathrm{CaCO}_{3}$ solubility limit of water. When the direction of electricity changes in EDR, ions accumulate in different parts of the stack (Fig. 3.3b). As the previously contaminated compartment are drained of their ionic constituents, the concentrations of calcium and carbonate drop below the $\mathrm{CaCO}_{3}$ solubility limit. Any precipitated $\mathrm{CaCO}_{3}$ then re-dissolves, leaving an essentially clean membrane. Precipitates formed in a normal ED stack have to be removed with acid or some other cleaning agent (Alsakari et. al. 1977, Schoeman 1985, Hughes et. al. 1992). Current reversal allows EDR to do this automatically and internally.

Of the 4 available ion-selective technologies, ED and EDR are the most promising. EDR does appear to have certain inherent advantages over the simpler ED alternative. A more detailed comparison of the two systems has, however, yet to be discussed.

### 3.3.2 Water permeable membranes

Unlike ion selective processes, water filtering systems are essentially sieves which capture and remove contaminants carried in the water as it passes through the membrane. Exactly which elements are removed is dependent upon filter pore size. As water slips through the pores, any particles or molecules larger than these spaces will be trapped above the membrane. Filters with smaller pores can, therefore, filter out a broader range of substances.

Tiny holed membranes do have certain restrictions. Water flow rates drop significantly as pores sizes shrink; it can reach a point where water has to forcibly pumped through the filter (Kalinske et. al. 1979, Applegate 1984, AWWA 1989, Cartwright 1991). Tighter membranes also have a higher probability of becoming clogged, as larger elements in the feedwater block pore spaces, or even cover up entire portions of the membrane wall (Applegate 1984, Parekh 1991). Although filters with tiny pores can strip a broader range of contaminants from a feed stream, using the tightest membrane is not always the best choice given the slower permeation rates and
higher clogging potentials associated with these filters. The success of a filtering system at the FNGP is, however, likely to depend on its ability to eliminate the smaller molecular contaminants found within the plant's wastewater. Using a very tight membrane would, in this case, be unavoidable.

There are several types of membranes designed for molecular sieving: ultrafilters (Applegate 1984, Jordain 1987, AWWA 1989, Ericsson \& Hallmans 1994a), nanofilters (Rohe et. al. 1990, Kopp et. al. 1993, Wiesner et. al. 1994, Turner \& Kadubandi 1994, Jacangelo et. al. 1995) and reverse osmosis filtration (Kalinske et. al. 1979, Applegate 1984, AWWA 1989, Parekh 1991, Morin 1994). Ultrafiltration is the least effective of the three options; it can only remove substances measuring more than $0.005 \mu \mathrm{~m}$ in diameter (Fig. 3.4). While bacteria, viruses and most suspended particles fall within these specifications (Applegate 1984, Jordain 1987, AWWA 1989, Ericsson \& Hallmans 1994a), calcium, sulfate and other ionic elements will pass unhindered through ultrafiltration membranes (Applegate 1984, AWWA 1989, Ericsson \& Hallmans 1994a). Ultrafilters are restricted to relatively large particle removal.

Nanofilters are slightly tighter membranes. They can screen out ions as small as $0.0009 \mu \mathrm{~m}$ (Fig. 3.4). This includes many divalent species, such as calcium, sulfate, phosphate and magnesium (Cluff 1992, Kopp et. al. 1993, Fu et. al. 1994, Ericsson \& Hallmans 1994b, Jacangelo et. al. 1995). Depending on their configuration, some monovalent ions, like sodium and chloride, can also be stripped out of solution using nanofiltration (Taylor et. al. 1989, Cluff 1992, Fu et. al. 1994). Nanofilters are not, however, very efficient at removing these singlely-charged elements (i.e. divalent removal rates can $=>95 \%$, monovalents rates are only around $35 \%$ Conlon \& McClellan 1989, Taylor et. al. 1989, Cluff 1992, Comb 1994, Fu et. al. 1994). While able to eliminate a broader range of contaminants than ultrafiltration systems, nanofilters are generally only effective at removing the larger multi-valent ions from a feedwater stream.

Reverse osmosis ( RO ) is the tightest membrane filter available. It can contain pores as small as $0.0001 \mu \mathrm{~m}$ (Fig. 3.4). As a result, it can efficiently strip monovalent ions (Hrubec et. al. 1979,

Osantowski \& Geinopolos 1979, Kosarek 1979, Morin 1994, Comb 1994, Page 1995) and even uncharged atoms (Kalinske et. al. 1979, Kosarek 1979, Applegate 1984, Eisenberg \& Middlebrooks 1986, AWWA 1989, Wethern et. al. 1991) out of water. Of the three molecular sieves, RO is the most effective one at removing a wide range of contaminants. It therefore seems to be the best of the water permeable membrane filters.

### 3.3.3 Reverse osmosis vs. electrodialysis

Within their respective groups, ED/EDR and RO are the most promising of the available filtering technologies. Each method has inherent advantages and limitation. Key factors include their respective water recovery ratios, cleaning ability and robustness, as well as the technical expertise required to operate each system.

## Water recovery ratios

RO: Unlike ordinary paper filters where water flows perpendicular to the membrane surface, RO feedwater travels parallel to the filter (Fig. 3.5). By pressurizing the feed stream and limiting the exiting water flow, some water is forced through the RO membrane. The contaminants left on top of the filter are carried out of the unit in the remaining water. Therefore, two separate stream are produced by a RO membrane: a clean water, or permeate flow and a dirty water, or concentrate stream (Fig. 3.5). Permeate production is generally limited by 2 natural phenomena: osmotic pressure and particle diffusion (Applegate 1984, , Pohland 1987, Cartwright 1991).

Unless acted upon by an outside force, water moves from areas of low solute concentration to zones of higher molecular abundance (Applegate 1984, Pohland 1987, Kotz \& Purcell 1987, AWWA 1989). In RO, the reverse flow pattern is desired. High quality water is obtained by pushing liquid from the contaminated feed stream through to the relatively pure permeate side of the filter assembly (Pohland 1987, AWWA 1989, Parekh 1991). Feedwater pressure's must be greater than the osmotic pressure drawing clean water back across the membrane for there to be any permeate production (Applegate 1984, Pohland 1987, AWWA 1989, Parekh 1991).

As water moves across a RO membrane, contaminant concentrations in the waste stream increase. The more concentrated the waste becomes, the greater the osmotic force pulling permeate back through the filter. Consequently, larger pressures are needed to ensure that the net flow of water is from concentrate to permeate. Eventually, as concentrate volumes get smaller and smaller, the pressure needed to drive the system will exceed filter design, and the membrane will burst. Water recovery rates in RO are therefore limited by the increasing osmotic pressure associated with higher permeate flows.

Just as water naturally travels up the concentration gradient - from low concentrations to high ones, molecules diffuse down it - from areas of high concentration to those of low molecular abundance (Applegate 1984, Pohland 1987, Kotz \& Purcell 1987, Cartwright 1991). Travel rates are determined by the magnitude of the concentration gradient (Applegate 1984, Pohland 1987, Cartwright 1991). The larger the difference between the two zones, the greater the flow of molecules from one area to the other. So, as the concentrate stream in a RO filter gets more concentrated with higher permeate recovery rates, the gradient gets steeper. The flow of salts and other undesirable elements across the membrane subsequently increases and permeate quality decreases (Applegate 1984, Pohland 1987, Cartwright 1991). Due to the increases in salt passage and required feedwater pressures resulting from higher permeate production rates, individual RO units are typically restricted to recovering $50 \%$ of the incoming wastewater flow (Applegate 1984, Parekh 1991)

To increase overall wastewater recovery rates, RO vessels can be linked in series. The concentrate produced in the first is directed into the second and so on until the desired output flow is obtained (Applegate 1984, Parekh 1991). Total recoveries are, however, limited by feedwater chemistry.

Certain mineral salts are relatively insoluble in water (e.g. $\mathrm{CaCO}_{3}$ ) (Kotz \& Purcell 1987). As a result, when ionic concentrations (e.g. $\mathrm{Ca}^{2+}$ and $\mathrm{CO}_{3}{ }^{2-}$ ) exceed solubility limits, these salts will begin to precipitate out of solution. This is relevant to RO when one considers that any feedwater
entering an RO unit contains a given volume of water and a given mass of solute. The produced permeate will be mainly water with few contaminants (if recovery rates $\leq 50 \%$ ). The concentrate will, on the other hand, consist of almost all of the influent contaminant mass dissolved in the remaining volume of water. Ionic concentrations in the RO waste stream are, therefore, higher than in the feed stream. If, at these "new" levels, the solubility of the sparingly soluble salts has been exceeded, then precipitation will occur. The amount of water that can be withdrawn prior to precipitation is dependent on feedwater quality. If a wastewater is initially already close to the solubility limits, then very little water can be recovered before solids start to form. Alternatively, a good deal of water could be reclaimed from a waste stream with very low scaling potential.

Once precipitation has started to occur, membrane efficiency will decrease rapidly, as portion of the membrane become coated with salt (Kosarek 1979, Pohland 1987, AWWA 1989, Huang \& Koseoglu 1993, Noshita 1994). Although some solids can be removed with adequate cleaning (Eisenberg \& Middlebrooks 1986, Pohland 1987), most filters will need to be replaced if salt precipitation has been extensive (Eisenberg \& Middlebrooks 1986, Pohland 1987).

RO water recovery rates are limited by osmotic pressure and molecular diffusion to around $50 \%$ per RO vessel. Greater recoveries can be achieved by running two or more units in series. Influent chemical characteristics will, however, restrict the success of multi-stage RO filtration. Permeate can only continue to be produced if the concentration of the sparingly soluble salts is low. Once solubility limits have been exceeded, system performance will fall, and costly repairs may be required.

ED/EDR: For ion-selective processes, such as ED and EDR, product water volumes are independent of feedwater quality. They are, instead, determined solely by the configuration of an ED or EDR stack. As indicated earlier, an ED or EDR unit consists of alternating cationic and anionic membranes grouped between a cathode and an anode, with water flowing through the spaces between the membranes (Figs. 3.2 \& 3.3a). When electricity is added to the system, ions will move out of certain compartments into the others (Figs. $8 \& 9 \mathrm{a}$ ). Clean water is produced in
areas drained of their ionic elements, while the waters in neighbouring spaces become increasingly contaminated. The ratio of clean to dirty water compartments is always $x: x-1$, where " $x$ " is equal to the number of spaces in the stack producing clean water (e.g. Fig. 3.2). Water recovery ratios for individual ED and EDR units are, therefore, between 50 and $60 \%$ of the incoming flow (i.e. recovery rate $=$ number of clean spaces $/$ total spaces $=x /(x+x-1)=x /(2 x-1))$. To increase overall wastewater recoveries, concentrate can be reprocessed either by recycling it back through the same stack or treating it in one or more downstream units (Applegate 1984, Schoeman 1985, Solt \& Foster 1987).

## Cleaning efficiency

RO: RO membranes contain extremely small pores. As a result, they are able to screen ions and other atomic particles from a liquid. RO filters have been found to have the following mean removal efficiencies: 94 \% for total dissolved solids (TDS) (Light et. al. 1984, Pohland 1987, Marquardt et. al. 1987, Shah et. al. 1993, Noshita 1994, Abdula'aly \& Chammem 1994), 96 \% divalent ions (Hrubec et. al. 1979, Kosarek 1979, Light et. al. 1984, Pohland 1987, Marquardt et. al. 1987, Shah et. al. 1993, Abdula'aly \& Chammem 1994, Noshita 1994), $94 \%$ - monovalents (Hrubec et. al. 1979, Kosarek 1979, Light et. al. 1984, Pohland 1987, Marquardt et. al. 1987, Shah et. al. 1993, Abdula'aly \& Chammem 1994, Noshita 1994), and $82 \%$ for total organic carbon (TOC) (Kosarek 1979, Marquardt et. al. 1987, Chin \& Ong 1991).

ED/EDR: ED and EDR systems, on the other hand, have much lower removal efficiencies. The product water from a given unit typically contains $53 \%$ of the TDS (Alsakari et. al. 1977, Kawanishi et. al. 1994, Kawahara 1994), as well as between 40 to $45 \%$ of the mono- \& divalent ions (Alsakari et. al. 1977, Hughes et. al. 1992, Kawanishi et. al. 1994, Kawahara 1994), found in the feedwater. Furthermore, these systems can only act on charged particles (Applegate 1984, Schoeman 1985, Solt \& Foster 1987, AWWA 1989, Huang \& Koseoglu 1993). The abundance of silica, natural organics and other inert materials in both product and concentrate streams will be identical and unchanged from feedwater concentrations.

Ionic diffusion is probably a key factor which restricts the cleaning ability of ED/EDR. As ions accumulate in the concentrate compartments, concentration gradients across the separating membranes would increase. This, in turn, should promote a greater flux of charged particles back into the product water. Given the average removal rates described in the literature (described above), the optimum balance between active ion displacement and passive diffusion appears to occur when product water concentrations are around $50 \%$ of those of the feedwater.

Feedwater quality will influence how close to this apparent $50 \%$ maximum ED/EDR processes can operate. The ability to concentrate ions within the dirty water spaces in an ED/EDR stack will be bound by the solubility limits of sparingly soluble salts (Schoeman 1985, AWWA 1989, Huang \& Koseoglu 1993). If a feedwater is rich in (say) carbonate and calcium, then few calcium and carbonate ions can be transferred into the contaminant compartments before calcium carbonate begins to precipitate onto the surrounding membranes. Given the ability of EDR units to periodically shed precipitated solids (see "Ion Selective Separation"), the performance of these systems will be less affected by salt scaling. They should, therefore, be able to operate closer to the $50 \%$ maximal removal rates. Just as overall RO water recovery ratios are restricted by feedwater scaling potential, so too is the cleaning ability of individual ED and, to a less extent, EDR stacks.

Product water quality from one ED/EDR unit can always be improved by putting it through other downstream stacks (e.g. Alsakari et. al. 1977, Hughes et. al. 1992, Kawahara 1994). There are, as one might expect, limits to this process. Finished water volumes will be reduced by 40 to 50 \% per ED/EDR unit (see "Water Recovery Ratios"). Recycling the resulting wastewater would minimize these water losses, but the larger feedwater flows (i.e. original water + recycled concentrate) would necessitate a larger ED/EDR system. More importantly, however, overall cleaning efficiencies are in themselves restricted by product water purity (Applegate 1984, Solt \& Foster 1987).

The driving force in this process is electricity. Water by itself is a relatively poor conductor of electricity. As a result, as ion concentrations in product water compartments decrease, the electrical resistance of these waters increases. More energy needs to be added to the system to maintain a current through the entire stack. Eventually, it becomes impossible to import enough electricity into the process to promote ionic displacement from product to concentrate waters. Finished water quality can no longer be improved, regardless of how many units it travels through (Applegate 1984, Solt \& Foster 1987).

ED and EDR systems can only influence charged particles. A single ED or EDR stack can remove approximately $50 \%$ of the ionic load of the feedwater stream. Although cleaning efficiencies appear to be limited by ionic diffusion, an abundance of sparingly soluble salts in the feedwater can further reduce the effectiveness of a ED or EDR system. Finished water quality can be improved by running it through additional ED/EDR units. The design of such a system would, however, have to balance cleaner water against the costs of obtaining \& maintaining the desired product water quality \& quantity.

## Optimum TDS

RO: Reverse osmosis units tend to most effective when feedwater TDS concentrations are between 2000 and $3000 \mathrm{mg} / \mathrm{L}$ (Kalinske et. al. 1979, Morin 1994).

ED/EDR: Both electrodialysis and electrodialysis reversal are generally most effective when dealing with an inlet TDS of less than $2600 \mathrm{mg} / \mathrm{L}$ (Solt \& Foster 1987, Cluff 1992, Kawahara 1994).

## Membrane fouling

RO: RO filters can be fouled by a number of substances. Some of these elements will affect any RO system, while others are membrane specific. The former category includes suspended and colloidal solids (Larson \& Argo 1976, Kalinske et. al. 1979, Kosarek 1979, Pohland 1987, AWWA 1989, Suemoto et. al. 1994), silicates (Marquardt et. al. 1987, Kawanishi et. al. 1994,

Noshita 1994, Abdula'aly \& Chammem 1994), salt scaling (Kalinske et. al. 1979, Kosarek 1979, Pohland 1987, AWWA 1989, Huang \& Koseoglu 1993, Comb 1994, Noshita 1994), and microbes (Pohland 1987, AWWA 1989, Chin \& Ong 1991, Noshita 1994, Abdula'aly \& Chammem 1994, Suemoto et. al. 1994). "Extreme" pH conditions (Applegate 1984, Eisenberg \& Middlebrooks 1986, AWWA 1989) and various oxidizing agents, such as oxygen \& chlorine (Applegate 1984, Eisenberg \& Middlebrooks 1986, Pohland 1987, AWWA 1989), are examples of the latter group.

To protect RO equipment, these contaminants are removed from the feed stream prior to its filtration. Pretreatment trains can include multi-media filters (Hrubec et. al. 1979, Osantowski \& Geinopolos 1979, Kaakinen \& Moody 1984, Wethern et. al. 1991, Pankratz \& Johanson 1992, Shah et. al. 1993), ultrafiltration (Kaakinen \& Moody 1984, Wethern et. al. 1991), activated carbon columns (Osantowski \& Geinopolos 1979, Kaakinen \& Moody 1984, Pohland 1987, Marquardt et. al. 1987, Wethern et. al. 1991), and chlorination (Larson \& Argo 1976, Kaakinen \& Moody 1984, Suemoto et. al. 1994) and/or dechlorination (Pohland 1987, AWWA 1989) stations all depending on the extent of feedwater contamination, as well as which elements are of concern. They also tend to be more extensive than systems used for ED and EDR protection, since RO filters, on the whole, are generally more sensitive to fouling than either ED or EDR (AWWA 1989).

ED/EDR: Similar contaminating agents affect ED and EDR membranes, including suspended and colloidal solids (Applegate 1984, Schoeman 1985, Solt \& Foster 1987, Huang \& Koseoglu 1993), oxidizers (Applegate 1984, Schoeman 1985), and microbes (Applegate 1984, Huang \& Koseoglu 1993). Silicates, on the other hand, pose no threat to ED or EDR (Kawahara 1994), but organics can (Kalinske et. al. 1979, Applegate 1984, Schoeman 1985, Solt \& Foster 1987, Huang \& Koseoglu 1993). ED systems can also be detrimentally affected by salt scaling (Schoeman 1985, AWWA 1989, Huang \& Koseoglu 1993). This is of less concern for EDR stacks, due to the self-cleaning mechanisms inherent in the process (see "Ion Selective Separation").

Pretreatment trains for ED and EDR contain similar elements as RO units; they are, however, generally simpler than RO systems, due to the greater robustness of ED and especially EDR membranes (AWWA 1989).

## Technical operation \& maintenance

RO: Reverse osmosis is a relatively straight forward process. Pretreated feedwater flows into a RO vessel, and exits either as concentrate or permeate (Fig. 3.5). The whole process is controlled by flow meters, pumps and valves (e.g. Applegate 1984, Pohland 1987, Parekh 1991).

Though regular chemical cleaning will help to maintain optimal filter performance (Applegate 1984, Pohland 1987), RO membranes will eventually need to be replaced. They have an average life span of about 5 years (Pohland 1987). Inadequate pretreatment, highly contaminated feedwaters or harsh operating conditions will significantly reduce filter life (Applegate 1984, Light et. al. 1984, Pohland 1987). In any case, since RO membranes exist as individual units (e.g. Applegate 1984, Eisenberg \& Middlebrooks 1986, Pohland 1987, Parekh 1991), it is relatively easy to isolate and replace a faulty or spent filter.

ED/EDR: By their very configuration, ED and EDR units are complex and potentially hard to maintain. An ED/EDR stack contains between 100 and 600 membrane pairs sandwiched between a cathode and an anode (Applegate 1984, Solt \& Foster 1987). Bounded by each pair is a clean and dirty water compartment (Fig. 3.2). Water is carried to and from each opening in separate tubes. Fresh water is also continually traveling over the cathode and anode to keep them free of contaminants (Applegate 1984, Solt \& Foster 1987). This results in an extensive array of pipes and valves, in addition to the equipment controlling the flow of electricity to and from the stack.

EDR has the additional complications associated with periodic current changes. When the flow of electricity changes direction, product water compartments become filled with concentrate, and vice versa (Figs. 3.3a \& b). The stack has to be purged before product water can be collected from its new locations to prevent contamination with any remaining concentrate (Applegate 1984). As a
result, EDR units have even more complex valving systems than ED stacks (Applegate 1984, Schoeman 1985). Whicheyer process is used, ED or EDR, the operators will have to be familiar with both water flow and electrical instrumentation, and they must be able to identify potentially problematic situations before they disrupt the ED or EDR cleaning process.

As previously mentioned, there are between 100 and 600 cell pairs in a ED/EDR unit, and water travels to and from each membrane set in a separate tube. There are, as a result, hundreds of tiny pipes running into and out of a single stack! If one of these tubes were to become blocked, locating the problem could take some time. Similarly, if a membrane should fail or wear out, a good portion of the entire stack would have to be disassembled to find and replace the spent filter (Hughes et. al. 1992). Any time a stack is taken apart, be it for manual cleaning or membrane change, there is generally a high probability of membrane damage (Applegate 1984, Hughes et. al. 1992). Clearly, maintaining a ED or EDR system can be both time and labour intensive.

Where as RO technology offers simplicity and ease of operation, ED and EDR are complex, potentially time consuming reclamation processes. Technical personnel would require greater training to operate and maintain such systems; maintenance cost are also likely to be higher than those associated with RO.

## The better system

RO is slightly more sensitive to fouling than either ED or EDR, and, as a result, generally requires more extensive pretreatment systems (Table 3.1). Even though average water recovery rates per operating unit are almost equivalent for ED, EDR and RO (Table 3.1), overall product water volumes also tend to be greater for ED and EDR. Product water quality is, on the other hand, likely to be better with RO, due to more efficient removal of a broader range of contaminants (Table 3.1). The best feature of reverse osmosis is its simplicity and operational ease, especially when compared to the more complex ED and EDR technologies (Table 3.1). In face of all of the above criteria, reverse osmosis appears to be the most effective membrane filtration system.

### 3.3.4 RO membranes selection

RO membranes are derived from either cellulose acetate (Applegate 1984, Eisenberg \& Middlebrooks 1986, Pohland 1987, AWWA 1989) or poly-organic compounds, such as polyamide (Applegate 1984, Eisenberg \& Middlebrooks 1986, Pohland 1987, AWWA 1989). Cellulose-based membranes (CAs) are generally cheaper than polyamides (PAs) (AWWA 1989). They are also more resistant to oxidizing agents, such as chlorine and dissolved oxygen, than PAs (Applegate 1984, Eisenberg \& Middlebrooks 1986, Pohland 1987, AWWA 1989). On the other hand, PAs can operate under broader temperature and pH conditions, (Eisenberg \& Middlebrooks 1986, AWWA 1989 for the former; Applegate 1984, Eisenberg \& Middlebrooks 1986, AWWA 1989, Abdula'aly \& Chammem 1994 for the latter). They are also resistant to biological degradation (Shields 1979, Applegate 1984, Pohland 1987), unlike CAs (Applegate 1984, Eisenberg \& Middlebrooks 1986, AWWA 1989), and PAs require lower water pressures than CAs to produce a given permeate flow (AWWA 1989). As both membrane types generally have the same filtering abilities, either one could be used in any treatment system; the RO pretreatment train would simply have to be designed to produce the required feedwater characteristics (e.g. proper pH , temperature, dissolved oxygen and chlorine levels...).

RO filters, regardless of their individual make-up, come in one of 4 different configuration: tubular (Eisenberg \& Middlebrooks 1986, Pohland 1987, AWWA 1989, Parekh 1991), plate-andframe (Pohland 1987, Eisenberg \& Middlebrooks 1986, AWWA 1989, Parekh 1991), hollowfibre (Eisenberg \& Middlebrooks 1986, Pohland 1987, AWWA 1989, Parekh 1991), and spiral wound (Eisenberg \& Middlebrooks 1986, Pohland 1987, AWWA 1989, Parekh 1991, Morin 1994). Plate-and-frame, as well as tubular designs, while effective, have limited applications. Their relatively small membrane surface area to unit volume ratios $\left(165 \& 335 \mathrm{~m}^{2} / \mathrm{m}^{3}\right.$, respectively - Pohland 1987) makes treating anything by very small feedwater flows extremely expensive (Pohland 1987, Parekh 1991).

Both spiral wound and hollow-fibre membranes have much larger area to volume ratios (1000 \& $16500 \mathrm{~m}^{2} / \mathrm{m}^{3}$, respectively - Pohland 1987). The use of hollow fibre systems has been restricted in the past by their relative sensitivity to fouling (Applegate 1984, Pohland 1987), and their need for higher feedwater pressures (Morin 1994). As a result, the operating costs associated with hollow fibres are greater than those for spiral would designs (Morin 1994). The dominance of spiral would membranes is also due in part to the ease with which recent advances in membrane technology have been incorporated into these designs as compared to hollow fibre configurations (AWWA 1989). Thus, although membrane type is a somewhat arbitrary choice based on feedwater characteristics, treatment system generally always use spiral-wound RO filters, because of the relative robustness and low cost of this configuration.

### 3.4 Brine Disposal

Regardless of whether distillation, ion-exchange or membrane filtration is used, each system will produce a final wastewater requiring further treatment; these waters could be either regeneration fluids from the ion-exchanges, very concentrated blowdown flows from the distillation vessels or RO membrane concentrates. As the goal of the ZED project is to completely eliminate the need for a wastewater discharge from the FNGP, none of the previously discussed technologies is, on its own, sufficient to fulfill this objective.

Power plants and other zero-effluent facilities described in the literature (e.g. Brew \& Blackwell 1991, Pankratz \& Johanson 1992, Bowlin \& Ludlum 1992, Pierce \& Sbei 1993, Strauss 1994) which have similar water demands as the FNGP (i.e. make-up water to replace boiler and other process losses) incorporate one of two systems into their wastewater treatment trains. Concentrated liquid wastes are either disposed of in evaporation ponds (Bowlin \& Ludlum 1992, Pierce \& Sbei 1993), or they are solidified prior to discharge and landfilled (Brew \& Blackwell 1991, Pankratz \& Johanson 1992, Bowlin \& Ludlum 1992). Evaporation ponds are rather self-explanatory. Wastewater drains into the holding ponds and evaporates. The remaining solids collect in the basins, which are eventually capped with earth.

Wastewater solidification, on the other hand, tends to be best carried out in two steps. Concentrate volumes are first reduced through forced evaporation using a brine concentrator (B/C) (Brew \& Blackwell 1991, Bowlin \& Ludlum 1992, Pankratz \& Johanson 1992). This robust piece equipment is specifically designed to handle very contaminated liquids. It is not affected by the precipitation of salts and other solids, typically formed during wastewater evaporation. Rather than coating the inner workings of the $\mathrm{B} / \mathrm{C}$, these substances precipitate onto seed elements added to the feed stream as it enters the unit (Pankratz \& Johanson 1992, Bowlin \& Ludlum 1992). B/Cs can recover up to $90 \%$ of the incoming wastewater (Brew \& Blackwell 1991, Strauss 1994), and product water quality is very high ( $<10 \mathrm{mg} / \mathrm{L}$ TDS - Bowlin \& Ludlum 1992, Pankratz \& Johanson 1992). The remaining $10 \%$ exits the B/C and travels to the second part of the solidification process.

Brine concentrator wastes can go to either a spray dryer (Brew \& Blackwell 1991, Bowlin \& Ludlum 1992), or a crystallizer (Bowlin \& Ludlum 1992, Pankratz \& Johanson 1992). Wastewater entering a spray drying is atomized into tiny droplets, which then fall through a hot air chamber. Water evaporates, and the remaining solids fall into a collection basin at the bottom of the unit. The produced material tends to be very fine and powdery, because it forms from tiny water drops. It is completely dry, and can be directly landfilled (Brew \& Blackwell 1991, Bowlin \& Ludlum 1992).

Crystallizers work somewhat differently. Hot, pressurized wastewater is injected into a hollow-bodied vessel (Bowlin \& Ludlum 1992, Pankratz \& Johanson 1992). As the pressure in the tank is less than that of the feedwater, flash evaporation occurs. Not all of the feed stream vapourizes at once; some of it remains as a liquid. This water falls to the bottom of the vessel, along with any newly-formed solids. Some of this solution is recycled back through the process, while the remaining portion is drained from the system. This bleed stream is generally further dewatered with a filter press (Bowlin \& Ludlum 1992, Pankratz \& Johanson 1992). The resulting solid filter cake can then be landfilled.

About $75 \%$ of the feedwater flowing to a crystallizer will be turned to steam, which can be recovered and reused (Pankratz \& Johanson 1992, Bowlin \& Ludlum 1992). Spray dryers are not designed for steam recovery; any and all vapourized water is lost to the atmosphere (Brew \& Blackwell 1991, Bowlin \& Ludlum 1992). These units are, on the other hand, a final solution. Incoming wastewater is completely transformed to a dry solid. There is no need for additional dewatering equipment, as is the case with crystallizers. Capital and operating costs are, therefore, likely to be lower for ZED operations using spray dryers instead of crystallizers. The fine, powdery nature of the solids produced by a spray dryer can, however, make them difficult to handle (Brew \& Blackwell 1991).

In summary, the small volumes of wastewater draining from distillation, ion-exchange or membrane filtration processes can be disposed of either in evaporation ponds or by more energy intensive forced evaporation systems. Such a system is likely to include a brine concentrator and a crystallizer or spray dryer. Regardless of its configuration, any brine disposal technology will produce two outflows: relatively pure water, which either escapes to the atmosphere or is collected and reused, and a solid waste, either in the form of precipitates in an evaporation pond, fine powder from a spray dryer or a filter cake from a crystallizer/filter press assembly.

### 3.5 Conclusion

Current literature clearly indicates that the FNGP can become a ZED facility. The required technology exists, and other facilities with similar water demands as the FNGP already operate under a ZED framework. A wastewater recycling program at the FNGP is likely to use either distillation, ion-exchange or RO as the main treatment process. Remaining concentrates will be directed into either an evaporation pond or a brine concentrator - spray dryer or crystallizer assembly. Wastewater currently leaving the plant will be reduced to a solid waste.

Table 3.1: A comparison of the inherent advantages and disadvantages of reverse osmosis (RO), electrodialysis (ED) and electrodialysis reversal (EDR).

| Parameter | RO | ED | EDR |
| :--- | :---: | :---: | :---: |
| Water recovery rate / vessel | $50 \%$ max. | $50-60 \%$ | $50-60 \%$ |
| Removal rates: |  |  |  |
| total dissolved solids (TDS) | $94 \%$ | $53 \%$ | $53 \%$ |
| monovalent ions | $94 \%$ | $55 \%$ | $55 \%$ |
| divalent ions | $96 \%$ | $60 \%$ | $60 \%$ |
| total organic carbon | $82 \%$ | - | - |
| Can it remove non-ionic particles? | Yes | No | No |
| Optimal TDS - mg/L | $2000-3000$ | $<2600$ | $<2600$ |
| Fouling elements: | Yes | Yes | No |
| Salts | Yes | Yes | Yes |
| Microbes | Yes | No | No |
| Silica | No | Yes | Yes |
| Organics | Yes | Yes | Yes |
| Solids | High | Moderate | Low |
| Relative pretreatment needs | Simple | Complex | Highly complex |
| Operation \& maintenance |  |  |  |



Figure 3.1: Typical operating total dissolved solid (TDS) concentrations for different desalination technologies.

- adapted from AWWA 1989

$\begin{array}{ll}A=\text { Anion selective membrane } & A-=\text { Anion } \\ C=\text { Cation selective membrane } & C+=\text { Cation }\end{array}$

Figure 3.2: An illustration of the electrodialysis process.
adapted from Applegate 1984


Figure 3.3a: Flows through an EDR stack when electricity travels from right to left.


Figure 3.3b: Flows through an EDR stack when electricity travels from left to right.

- adapted from Applegate 1984

Figure 3.5: An illustration of reverse osmosis (RO) filtration.
- adapted from Applegate 1984


### 4.0 METHODOLOGY

### 4.1 Analysis of the Existing Water Distribution Network

### 4.1.1 Flow data

Data collection: Water flow data used herein are a collection of metered readings, derived values, and numbers provided by plant personnel. Metered measurements originated from a number of sources. Some of the information was available in operational reports and other hard copy documentation. Other readings were extracted from the facility's continual monitoring computer system. Only several months worth of data was retrievable from the computer network. January 1995 was one of the few months where a complete set of measured values was available. All of the metered numbers used in this study were average flow rates recorded over those 31 days.

Data analysis: Water balances were used in conjunction with flow diagrams to identify leak points in the FNGP's water distribution network. One set of balances was developed around the individual operating units within the plant (e.g. Powerhouse, E \& F Process Trains and the Sulphur Plant). A second group followed each water loop through the entire facility (e.g. 450 and 45 psi steam systems). If the difference between total input and total output in a given balance was less the $10 \%$, the system was considered balanced.

### 4.1.2 Chemical data

Data collection: Water samples were collected by plant personnel from a number of different locations in the water distribution network (Fig. 4.1). A 1 L and a 200 or 250 mL acid-washed polyethylene bottle were filled at each sampling station. Nitric acid was added to the smaller container to lower its pH to $<2 \mathrm{pH}$ units; the acid was used as a preservative to stabilize metal concentrations within these samples (Greenberg et. al. 1992). No acid was added to the larger bottle, because this water was used to determine pH , alkalinity and a number of other "acid-
sensitive" parameters. Sampling generally occurred twice a week for three weeks in May $199{ }^{1}$. An additional set of samples was later collected in mid-July to evaluate developing trends in the data. At the completion of each round of sampling, full bottles were placed into a cooler with frozen ice packs, sent to the University of British Columbia's Environmental Engineering Lab, and, upon arrival, were stored at $4^{\circ} \mathrm{C}$ until they could be analyzed.

Upon removal from the refrigerator, the 1 L bottles were shaken and 80 mL of water was withdrawn from each for a total solids determination. The remaining waters were then filtered through $0.45 \mu \mathrm{~m}$ pore membranes and analyzed for acidity, conductivity, alkalinity, total carbon content (both organic and inorganic), and various inorganics (i.e. sulphate - $\mathrm{SO}_{4}$, phosphate $\mathrm{PO}_{4}$, chloride - Cl and silica - Si ). Acidity and conductivity were measured with a Beckman Ø44 pH probe and a Fisher Scientific Accumet ${ }^{\circledR}$ conductivity meter, respectively. Alkalinity was tested by titration (Greenberg et. al. 1992). Inorganic concentrations were determined by a Lachat Quickchem Flow Injection Analyzer, while a Shimadzu TOC-500 (total organic carbon) analyzer was used to identify both inorganic and organic carbon content.

Acidified waters from the smaller 200 / 250 mL containers were also filtered prior to examination. These waters were analyzed for calcium $(\mathrm{Ca})$, magnesium $(\mathrm{Mg})$, sodium $(\mathrm{Na})$ and iron ( Fe ) using AAS (atomic absorption spectrophotometry) following methods described in Thermo Jarrel Ash (1986).

All of the tested parameters were chosen based on their importance to boiler maintenance and operation (Jackson 1980, Robertson 1981, Schroeder 1991). A key part of the FNGP water network are the three boilers used to generate steam for the facility. They have the most stringent water quality requirements of all of the equipment in place at the plant (Shang Su , personal com.).

[^0]By identifying the abundance of the selected contaminants in different parts of the plant, ZED equipment can be designed to produce water clean enough to maintain current BFW quality, thereby ensuring a successful wastewater recycling program.

Data analysis: Some waters contained contaminant concentrations below detection limits. These samples were subsequently assigned the detection limit value. This value replacement procedure pertained only to the inorganics and the metals; the detection limits were $0.1 \mathrm{mg} / \mathrm{L}$ for $\mathrm{Cl}, 1.0 \mathrm{mg} / \mathrm{L}-\mathrm{SO}_{4}, 0.05 \mathrm{mg} / \mathrm{L}-\mathrm{PO}_{4}, 0.1 \mathrm{mg} / \mathrm{L}-\mathrm{Si}, 0.05 \mathrm{ppm}-\mathrm{Ca}, 0.10 \mathrm{mg} / \mathrm{L}-\mathrm{Mg}, 0.04$ $\mathrm{mg} / \mathrm{L}-\mathrm{Fe}$ and $0.02 \mathrm{mg} / \mathrm{L}$ for sodium.

Several individual observations were deemed non-representative and dropped from the data set, because they were either illogical (e.g. total carbon < organic carbon) or equal to $\pm 2 \mathrm{x}$ the average value of the remaining readings. Two entire samples were also tossed out of the study, since more than $50 \%$ of their defining parameters were found to be non-representative, as defined above. For example, discarded measurements included sodium and alkalinity readings from the May 24th treated water sample (Appendix C). The two completely omitted samples were the May 3rd reservoir and July 12th raw water samples (Appendix C).

The remaining data set was used in conjunction with water flow information to form mass balances of the water network. These balances were used to determine if all pertinent chemical and water flow paths had been accounted for, as well as to determine which of the plant's discharged waters could be directly reused, and which others would require treatment before being pumped back into the water system.

### 4.2 Select the Appropriate Recycling Technology for the FNGP

Components for the FNGP's ZED system were chosen based on their:

- ability to perform effectively at the plant
- relative capital and operating costs
- ease of operation
- associated safety risks


### 4.3 The Computer Simulator

### 4.3.1 Basic construction

A computer simulator was constructed to determine the "best method" of transforming the FNGP into a ZED facility. It was built in a 2-step process. All of the data collected at the plant was first imported into Microsoft Excel ${ }^{\circledR}$ to form a static image of the water distribution network. Sources and sinks identified in previously assembled mass balances were included to this model ${ }^{2}$. Most of the inputted data were then replaced with numerical formulas. The values displayed in one part of the simulator were now either directly or indirectly linked to the rest of the system. With the interlinking equations in place, changing the value of a given cell resulted in the recalculation of the whole worksheet.

The formulas themselves were derived from either assumption, information available in various FNGP training manuals (e.g. calcium concentrations in lime treater blowdown were calculated from data in the Water Treatment Manual), patterns in the collected data (e.g. chloride concentration in lime treater blowdown were the same as the raw inlet water) or a combination thereof. Every equation and its derivation is listed in Appendix E. Key assumptions used in building the simulator were:

- All identified sources and sinks were assumed to stay constant over time, and they were incorporated into the simulator as a percent increase or decrease, respectively.
- Regardless of influent characteristics, softened water leaving the ion-exchangers always had calcium and magnesium concentrations equal to those now found in the treated water used at the
${ }^{2}$ A source or sink was defined as a difference of $>15 \%$ between the total incoming and outgoing mass flux through a given system. The only source or sink that was not included in the simulator was a $>15 \%$ discrepancy in iron levels across the polishing pond. This observation was dismissed as an error; when the metal samples were acidified just after collection, bound iron was probably released from algae present in these waters.

FNGP (i.e. 1.4 mg of $\mathrm{Ca} / \mathrm{L}$ and 0.5 mg of $\mathrm{Mg} / \mathrm{L}$ - Appendix C ). Changes in feedwater hardness would only influence the rate at which the ion-exchangers needed to be regenerated, and not product water quality.

- All treated water losses from the system, except for the boiler, sulphur plant and lime treater blowdowns, as well as the 8 psi steam and ion-exchanger regeneration flows, were constant over time.
- Raw water chemistry was also constant over time, as were the pH readings of the boiler, sulphur plant and lime treater blowdowns and the flow of domestic filter backwash and pump sealant to the lime pond.

These assumptions, combined with observable trends in the original data, were used to assemble formulas which transformed the static model into a predictive simulator. All of the worksheet cells were interlinked. A change in any part of the water network now either directly or indirectly affected the rest of the system.

### 4.3.2 ZED components

The different recycling technologies (e.g. RO, B/C and spray dryers) were themselves represented by equations. Contaminant loads in concentrate and product waters were derived from waste removal rates found in the literature. Maximal water recovery rates for RO, B/C and other ZED vessels were also taken from published work. Together, contaminant load and exiting flow calculations served to define the products leaving any of the ZED components. Building different ZED treatment trains was simply a matter of linking the pertinent recycling units together. The assembled systems were then individually connected to the water network. As each cell in each computer worksheet was influenced by all of the other cells, the model automatically re-adjusted itself to account for any changes induced by the different ZED designs.

### 4.3.2 Parameter values

Incorporated into the simulator are a number of variables which can be altered by a user. They include, for example, the volume of 8 psi steam lost through venting and the final waste disposal system. All told, there are between 16 and 44 variables open to manipulation, depending on the chosen ZED configuration.

During testing, the values of certain parameters, such as the number of RO stages incorporated into a treatment train and the presence or absence of a brine concentrator, were automatically determined by the nature of the simulation. If the selected ZED design was a 1 -stage RO with a $B / C$, then clearly only 1 RO stage was used and a $B / C$ was included in the simulation. Each ZED configuration therefore imposed certain restrictions on a user's freedom to further manipulate the simulator.

Of the independent variables not affected by a given ZED design, a few were kept constant throughout the evaluation process to maintain comparability between different scenarios. They included:

- RO and $\mathrm{B} / \mathrm{C}$ contaminant removal efficiencies
- 8 psi steam losses, which were kept at zero
- a continual flow of $171 \mathrm{~m}^{3} /$ day of condensed 8 psi steam to the hot lime treater
- water temperatures of $35^{\circ} \mathrm{C}$ within the RO filters
- water pH values of 5.8 at the entrance of every RO filter stage
- water recovery rates of $50 \%$ on the first RO unit
- $\mathrm{B} / \mathrm{C}$ water recovery rates of $90 \%$

The specific values were selected based on either current conditions at the plant (i.e. 171 $\mathrm{m}^{3} /$ day of 8 psi steam to the lime treater), opportunities for system optimization (i.e. closing the 8 psi steam vents), or information in the available literature (e.g. previous research indicates that membrane filters tend to work most effectively when feedwater temperature and pH are between 30
to $50^{\circ} \mathrm{C}$ (Kalinske et. al. 1979, Applegate 1984, AWWA 1989) and 4 to 7 (Applegate 1984, AWWA 1989, Suemoto et. al. 1994), respectively).

The remaining parameters were changed during each simulation to provide a realistic, yet optimal view of the water distribution network under a ZED framework. For example, literature indicates that maximal wastewater recovery rates for RO filtration are $50 \%$ per stage (Applegate 1984, Parekh 1991). Ideally, the FNR discharge could be filtered through a 2 -stage RO, with each unit operating at $50 \%$ recovery. The waste stream would then be reduced to $25 \%$ of its original volume. Calcium, carbonate and sulfate concentrations in the composite wastewater are, however, high enough that at this level of treatment the RO membranes would quickly become contaminated with precipitated calcium salts. The process, while appearing to be very effective in terms of waste reduction and water recovery, would be unrealistic, due to the high salt content of the wastewater. A balance was therefore maintained between calcium pretreatment and water recovery rates to maximize the rate of wastewater reclamation while ensuring filter integrity. The final configuration of every tested ZED design was recorded, and the inputted parameter values are listed in Appendix F.

Figure 4.1: A simplified diagram of the water distribution system at the Fort Nelson Gas Plant

### 5.0 ANALYSIS OF THE EXISTING WATER DISTRIBUTION NETWORK

### 5.1 Results

### 5.1.1 Water balances

All of the individual operating units were in balance (Table 5.1), as was each water loop (Table 5.2). The only exception was the effluent plant. Almost $1 / 3$ of the wastewater collected at the effluent plant seemingly disappeared during treatment (Table 5.1). Plant personnel have indicated that this discrepancy may have resulted from faulty or inaccurate flow meters, as opposed to actual water losses (Bruce Kosugi, personal com.). The general balance of the water system seemed to indicate that all major flow paths had been identified and accounted for.

Numerous steam leaks and/or water discharges were discovered in the water system. Obviously, the effluent plant was one area of potentially significant water loss, should the flow meters prove to be accurate. Other pathways by which water escapes from the plant are listed in Table 5.3. While some of these flows were quite large (i.e. $>150 \mathrm{~m} 3 /$ day ), others were relatively insignificant (i.e. $<5.0 \mathrm{~m}^{3} /$ day $)$ (Table 5.3). All in all, the water distribution network seems to have been properly detailed, and the presence of leaks and other drainage points within the plant's flow network may indicate that the existing system can be optimized.

### 5.1.2 Mass balances

Mass balances were performed on 5 areas of the plant: the front-end softeners, the lime and polishing ponds, the sulphur plant and the boilers. There were large inequalities in each balance. The front-end softener (FES) and lime pond mass balances were perhaps the worst of the bunch. Significant amount of mass (i.e. $>15 \%$ of the total incoming flux) were missing from the FES balance for all but 2 parameters (Table 5.4). Similarly, the lime pond appeared to be a source of calcium, chloride, magnesium and a number of other chemicals, while also acting as a carbon, iron and phosphate sink (Table 5.5). Generally fewer than half of the 16 monitored parameters were significantly skewed in the remaining three mass balances (Tables 5.6, $5.7 \& 5.8$ ). In any case,
the presence of so many large discrepancies calls into questions the quality of the collected chemical data.

### 5.2 Discussion

### 5.2.1 Data quality

Water flow numbers: The representative value of the water flow data set may at first seem somewhat suspect considering that it contains a number of readings averaged over only 1 month of the year. The January 1995 averages were, however, generally within $10 \%$ of mean flow volumes derived over longer periods of time (Table 5.9). Even when the difference between the 2 average values was greater that $10 \%$, it was rarely a significant difference $\left(\mathrm{t}_{\text {statistic }}>1.96, \mathrm{p}<\right.$ 0.05 ) (Table 5.9). The January 1995 data should, therefore, be representative of general flow patterns at the plant.

Although the January 1995 metered data may be representative of longer term trends, several of the flow meters from which these numbers originate may not in themselves be accurate. As previously mentioned, plant personnel have indicated that the observed water leak at the effluent plant may have been the result of faulty flow meters. Furthermore, according to the inlet and outlet flow meters on the front-end softening system, about 538 and $725 \mathrm{~m}^{3} /$ day entered and exited the FESs, respectively (Table 5.9). At these flow rates, there would appear to be a major hole in the incoming raw water pipeline, as well as insufficient treated water make-up to replace all of the treated water now lost from the plant (Table 5.10). If one calculates the raw and treated water flow rates by a heat balance on the 8 psi steam loop (Appendix B ), then the volumes of incoming raw water and outgoing treated water increase and the pipeline leak and treated water deficit disappear (Table 5.10). This discrepancy between the metered values and the calculated flows indicates that there may be, or at least have been, inaccurate flow meters at the plant, and that part of the collected data set may contain inaccurate numbers.

If incorrect information was used to analyze the water distribution network, the resulting water and mass balances could be wrong. Individual flow paths could have been missed, and/or leak
points may have been overlooked or underestimated. Yet, considering that plant personnel have repeatedly examined the resulting water balances (hence the discovery of the apparently inaccurate flow meters on the FESs), it is unlikely that the flow data continue to contain any serious flaws. It is, therefore, reasonable to conclude that the plant's water distribution network has been adequately defined by the collected data to proceed with the ZED project.

Chemical data: The large number of inequalities found in the 5 mass balances performed on different areas of this facility seemed to indicate that plant waters were either mis-sampled and/or improperly analyzed, or that additional flow paths remained unaccounted for. With respect to the FES balance, it was concluded that given the amount of chloride and sodium missing from the balance (Table 5.4) the regeneration waters from the ion-exchangers were never correctly sampled. Information in the Water Treatment Manual also indicates that the description of the hot lime treater blowdown waters was probably incorrect

The hot lime treater removes calcium, magnesium, silica and inorganic carbon from incoming raw water by causing these substances to precipitate. The resulting solids are then carried to the lime ponds within this vessel's blowdown stream. As previously discussed, filtered non-acidified water was used for silica and alkalinity measurements. Any solidified silica and/or carbonate within the collected blowdown waters would not have been detected. Similarly, the low recorded concentrations of calcium and magnesium in these waters (Appendix C) suggests that not all of the solidified calcium and magnesium was available for observation in the acidified metal samples. As a result, the large discrepancies in the FES mass balance were likely due to incorrect sampling of the ion-exchangers' regeneration brine and underestimated contaminant concentrations in the hot lime blowdown sample, rather than unidentified outflows from the softening system.

Using information from the Water Treatment Manual, it was possible to estimate what the ionexchange and hot lime treater wastewaters should have looked like (Appendix D). Recalculating the mass balance with these altered outflows showed the front-end softening system to be in near
perfect balance; there were significant differences in only 4 ,-instead of 10 , parameters (Table 5.11).

Problems with the original characterization of the front-end softening system was also responsible a large number of the inequalities found across the lime pond. For example, the lime pond appeared to be a source of calcium, magnesium, sodium and chloride (Table 5.5), because the influx of these elements from the hot lime treater and ion-exchangers was underestimated. When the lime pond mass balance was reworked with the calculated FES outflows, these chemical surpluses disappeared (Table 5.12). Yet, rather than resulting in an even balance for all 4 elements, the lime pond turned from a source of calcium and magnesium to a sink for these 2 chemicals (Table 5.12). It also continued to be a sink for silica and inorganic carbon (Table 5.12). Considering that these contaminants are flushed from the hot lime treater as precipitates, these findings are, in retrospect, hardly surprising. The calcium, magnesium, silica and inorganic carbon flowing into the lime pond from the hot lime treater were part of the suspended solids which settled out of solution within the lime pond, hence its appearance as a sink for these chemicals. By the same logic, it is not unreasonable to assume that the iron, organic carbon and phosphates missing from the lime pond outflow also precipitated out of solution, either on their own or as part of larger settlable solids.

The 3 remaining mass balances were unaffected by the problems with the ion-exchanger and hot lime treater samples. Discrepancies in these systems can, however, be attributed to other causes. Conditioning chemicals added to the boiler feedwater used in the both the boilers and the sulphur plant, while having very little effect on flow volumes, alter the chemical characteristics of this water (Nalco Chemical Program). They are, as a result, responsible for several of the inconsistencies observed across these 2 systems (Tables $5.6 \& 5.7$ ). Similarly, water samples taken from the polishing pond always contained algae. These organisms were likely the cause of the significant differences between total incoming and outgoing suspended solids, phosphate, dissolved organic carbon and iron (Table 5.8). For example, their mere presence would have
increased the level of suspended solids within these waters, and iron bound within the algae was also probably released into solution when the metal samples were acidified just after collection.

The results from the initial mass balances suggested that there were major problems within the assembled chemical data set. Further investigation showed that several areas of the FNGP had indeed been mis-sampled. There was, however, information available at the plant which was used to more accurate characterize of these flows. Reworking the mass balances with these new estimates eliminated a large number of the originally observed inconsistencies. Some of the remaining problems were attributed to other previously overlooked factors, including the addition of conditioning chemicals to the boiler feedwater tanks and the presence of algae in the polishing pond. Since many of the discrepancies observed in the original mass balances can be explained or completely eliminated, the chemical data collected at the plant, including that contained in several operation manuals, provided sufficient information to indicate that all pertinent water flows had been identified and characterized, and that one could now proceed with the design of a ZED system for the FNGP.

### 5.2.2 System optimization

As previously discussed in the introduction of this report, optimizing the plant's water distribution network will not only reduce the demand for treated water make-up, which can lead to subsequent reductions in a ZED system's final waste output, it will also limit the size, complexity and cost of the required ZED treatment train by limiting the volume of wastewater requiring purification prior to reuse. Increasing the efficiency of the existing water network can be achieved by closing off leaks in the system and by reusing relatively uncontaminated waste streams currently released from this facility.

Water leaks: The FNGP already appears to have quite a tight distribution network, as only 2 leak points were uncovered while compiling the water balances. They consisted of steam escaping from the plant's steam tracing pipes and wastewater lost from the effluent plant during treatment (Tables $5.3 \& 5.1$, respectively). While sealing up holes in the steam tracing network will
theoretically reduce the demand for treated water make-up, the actual benefits of such action may be somewhat inconsequential. The volume of escaping steam was estimated at $5 \mathrm{~m}^{3} /$ day (Table 5.3). This is less than $1 \%$ of the total volume of treated water now produced by the FESs (Table 5.1). Clearly, altering the state of the steam tracing system will have very little, if any, affect on the rest of the water network and/or the success of a ZED program at the plant.

The leak at the effluent plant was somewhat larger than the one in the steam tracing pipes. It was estimated at $50 \mathrm{~m}^{3} /$ day, almost $1 / 3$ of the effluent plant's total incoming water flow (Table 5.1). Despite its apparent size, the disappearance of wastewater from the effluent plant does not affect the treated water system. The effluent plant mainly receives domestic sewage and dirty raw water. If more or less water were to escape from the effluent plant, treated water make-up demands would remain unaffected. Recovering the lost effluent plant wastewater is not, therefore, going to affect a ZED system's waste output in the classic sense of increasing or decreasing waste production at the FESs. It will, however, alter the size of the required ZED treatment train and the volume of waste drained from this system, as indicated by tests with the computer simulator.

When the leak was initially sealed, the volume of the FNR discharge jumped from 282 to 332 $\mathrm{m}^{3} /$ day (Table 5.13). Similarly, the mass of solid waste leaving the ZED system increased from 415 to $483 \mathrm{~kg} /$ day (Table 5.13 ). Although both the volume of wastewater going into the ZED treatment train and the mass of solid waste leaving said system dropped when reclaimed wastewater was recycled back through the plant, they still remained at higher levels than when the leak at the effluent plant was left untouched (i.e. 285 vs. $240 \mathrm{~m} 3 /$ day and $294 \mathrm{vs} .233 \mathrm{~kg} /$ day, respectively - Table 5.13). The reason for the sustained increase in both measurements is the abundance of contaminants in the effluent plant outflow relative to the FNGP's raw water inflow (e.g. TDS readings of $2850 \mathrm{vs} .400 \mathrm{mg} / \mathrm{L}$, respectively - Appendix C). When the leak at the effluent plant was sealed, all of the contaminants previously escaping through this "hole" were added to the water system, and, although the flow of raw water into the FNGP did decrease below "open leak" levels once wastewater recycling was initiated, this flow reduction was insufficient to
prevent a-net increase in the flux of contaminants through the plant's water network. As a result, the plant produced a larger waste flow containing more contaminant mass than when the leak was left untouched.

Eliminating possible wastewater losses from the effluent plant will not, however, outright abolish the opportunity to initiate a ZED program at the FNGP. Operational water losses were found to be in excess of $600 \mathrm{~m}^{3} /$ day (Table 5.3). Effluent plant outflows were only $109 \mathrm{~m}^{3} /$ day (Table 5.1). Even if all of the wastewater lost from the effluent plant were recovered, effluent plant outflow would still be far less than current operational losses (i.e. $160 \mathrm{~m}^{3} /$ day vs. $>600$ $\mathrm{m}^{3} /$ day - Tables $5.1 \& 5.3$, respectively). The limits imposed by this project's first boundary condition have not been violated; the FNGP's domestic wastewater production rates are smaller than their operational water losses, and they would continue to remain so even if all of the wastewater observed to be escaping from the effluent plant were recovered. A ZED program can therefore be implemented at the FNGP without recycling any reclaimed wastewater through the domestic water system.

It is was initially suggested that sealing off all of the leaks present in the FNGP's water distribution network would increase the success of a ZED program by reducing both the size of the required ZED treatment train and the waste flow generated by said system. Given the nature of the 2 leak points identified in the water network, this assumption may not completely be valid. The loss of water from the steam tracing lines was so minute that eliminating these leaks would have virtually no affect on a ZED program. On the other hand, recovering wastewater apparently lost from the effluent plant, while not violating this project's boundary conditions, is likely to increase both the size of the required ZED treatment system and its final waste output. From the prospective of implementing a cost effective wastewater recycling program at the FNGP, there appears to be few benefits to closing off the 2 leak point identified in the water distribution network. Yet for the FNGP to be recognized as a ZED facility, the leak at the effluent plant will have to be sealed, regardless of the economical consequences of this action.

As indicated in the introduction, a.true ZED operation would not release any waste products into its surroundings. Given the cost, complexity and possible impracticality of such a process, the term ZED has been somewhat watered down to indicate that no wastewater is expelled from a ZED facility. By this definition, the FNGP will have to seal off the leak at the effluent plant as part of its transformation into a ZED operation. As for the steam escaping from the steam tracing lines, this water was not considered wastewater, because of its purity. Considering that evaporation ponds, spray dryers and other systems used to achieve ZED also release water vapour, this assumption appears to be the industry norm. Therefore, while attempts to recapture steam venting off of the steam tracing lines is relatively inconsequential to the ZED project, eliminating the leak at the effluent plant is a fundamental step, albeit an uneconomical one, in establishing a ZED program at the FNGP.

Water reuse: One of the FNGP's process engineers, Shang Su, has indicated that only high quality wastewater can be directly reused, because most of the process vessels have relatively stringent water quality requirements. As a result, identifying opportunities for wastewater reuse was limited to an examination of the plant's waste steam flows. Unfortunately, most of the uncovered steam vents were small leaks ( $<5.0 \mathrm{~m}^{3} /$ day ) (Table 5.3). Cost associated with collecting and reusing this steam overshadow the possible benefits (Shang Su, personal com.). One obvious exception was the loss of steam from the deaerators. Approximately $376 \mathrm{~m}^{3} /$ day of 8 psi steam was released from these vessels (Table 5.3). Given the quantity and quality of water involved, the 8 psi steam should be condensed and reused. To evaluate the potential benefits of reusing this water, a simulation was run wherein the 8 psi steam currently venting off the deaerators was condensed and returned to the boiler feedwater hotwells. No other changes were made, and no ZED equipment was used.

When the $376 \mathrm{~m}^{3} /$ day of steam was initially returned to the water system, the ratio of raw water to 8 psi steam entering the hot lime treater fell from 855:171 to $479: 171$. The increased dilution of raw water with high quality condensed steam led to improved treated water quality.

Cleaner treated water meant cleaner boiler feedwater, which, in turn, translated into less wastewater from the boilers and the sulphur plant (Table 5.14). At the same time, with less raw water traveling through the FESs, softener blowdown and regeneration rates also dropped. Reduction in all of these waste flows culminated in an additional saving of $72.1 \mathrm{~m}^{3} /$ day of raw water (Table 5.14). Overall, reusing the $376 \mathrm{~m}^{3} /$ day of 8 psi steam triggered changes throughout the water system, which ultimately resulted in a total raw water savings of the $448 \mathrm{~m}^{3} / \mathrm{day}$.

Aside from affecting raw water flows, reusing the 8 psi steam also influenced the FNR discharge. Wastewater volumes leaving the polishing pond fell in response to smaller incoming blowdowns, and, with less raw water being softened, the mass of contaminants stripped out by the softeners, traveling through the distribution network, and eventually exiting the plant similarly decreased (Table 5.14). Therefore, reusing the $376 \mathrm{~m}^{3} /$ day of 8 psi steam no only resulted in a total water savings of $448 \mathrm{~m}^{3} / \mathrm{day}$, it also led to a smaller, less contaminated plant outflow. Clearly reusing the 8 psi steam currently discharge from the deaerators is a key step to optimizing the existing water distribution system.

Vapour traveling to the thermo-oxidizer and later released to the atmosphere may be another potential water source ready for direct reuse (Table 5.3). Originally leached from natural gas stripping solutions, the vapour stream contains significant levels of hydrogen sulphide $\left(\mathrm{H}_{2} \mathrm{~S}\right)$ and carbon dioxide $\left(\mathrm{CO}_{2}\right)\left(\mathrm{S} / \mathrm{P}\right.$ Manual). Most of the $\mathrm{H}_{2} \mathrm{~S}$ is removed as elemental sulphur in the sulphur plant, while the steam, $\mathrm{CO}_{2}$ and remaining $\mathrm{H}_{2} \mathrm{~S}$ move onto the thermo-oxidizer and are eventually released to the atmosphere (Fig. 2.3). If this gaseous mixture was cooled prior to its arrival at the thermo-oxidizer, condensed steam could be collected and returned to the stripping liquors.

The role of the thermo-oxidizer is to transform remaining $\mathrm{H}_{2} \mathrm{~S}$ into sulfur dioxide $\left(\mathrm{SO}_{2}\right)$. This process is temperature dependent; good rates of conversion only occur under extremely hot conditions (i.e. $>500^{\circ} \mathrm{C}$ ) (S/P Manual). Although the vapour stream could theoretically be cooled
and the condensing water collected, it is unclear how a drop in feed temperature might affect-the $\mathrm{H}_{2} \mathrm{~S}-\mathrm{SO}_{2}$ conversion within the thermo-oxidizer. Furthermore, significant levels of $\mathrm{CO}_{2}$ and $\mathrm{H}_{2} \mathrm{~S}$ may return with the condensed water. Contaminated reflux may reduce the effectiveness of the stripping liquids. Further research would be required to assess if water can indeed be recovered from the thermo-oxidizer feed stream without affecting the performance of either the thermo-oxidizer or the stripping solutions. Reusing this water may also pose a problem to the entire ZED project. If all of the water released through the thermo-oxidizer was reused, together with the 8 psi steam venting off of the deaerators, operational water losses would fall dangerously close to domestic wastewater production rates (i.e. $55.3 \mathrm{~m}^{3} /$ day vs. $68.5 \mathrm{~m}^{3} /$ day - assuming that domestic outflow is equal to half of the dirty water reaching the effluent plant - Tables $5.1 \& 5.3$ ).

Most of the FNGP's steam discharges are too small to be effectively recaptured and reused. The 8 psi steam vent off of the deaerators is the obvious exception. Reusing this water-should greatly reduce raw water flows and lead to a smaller, less contaminated final plant outflow: Although the vapour stream traveling from process, through the sulphur plant and eventually to the thermo-oxidizer may be another potential source of reusable water, further research is needed to be sure that reusing this water will not negatively affect other parts of the FNGP, or that reusing this steam in conjunction with the aforementioned 8 psi steam will not jeopardize the whole ZED project by lowering operational water losses below domestic system output.

### 5.3 Changing to a ZED System

To establish a ZED program at the FNGP, wastewater generated at this facility which cannot be directly reused will have to be treated and recycled either individually or as a combined flow. There are inherent problems to dealing with each waste stream individually, as illustrated by the mechanisms needed to recycle process waters currently drained to the flare pits. Flare water, originally part of the natural gas stripping solution (Fig. 2.3), contains various hydrocarbons leached from the gas stream during processing. As a result, plant personnel have expressed strong apprehension about reusing any flare water for fear of a hydrocarbon build up within the
processing system, which could eventually have explosive results (Shang Su, personal com.). Installing a hydrocarbon separator would provide a means of removing the worrisome elements from the flare water. This wastewater could then be cleaned and safely poured back into the stripping solutions.

Flare water flow is, however, inconsistent. It does not, as suggested in Table 5.3, continually drain to the flare pits (Shang Su, personal com.). The hydrocarbon characteristics of this water still remain to be defined. Judging by the inlet gas profile (Table 5.15), it is bound to contain a number of different contaminants, all of which are too dilute to visibly separate out of solution (personal observation). Inconsistent flow and the potential abundance of low level hydrocarbons complicate the design of a hydrocarbon separator, and cast doubts on the benefits of recycling flare water.

According to plant schematics, water not vapourized during flaring travels out of the flare pits and into the effluent treatment plant. (Fig. 4.1). The effluent plant produces a relatively constant outflow (Table 5.9). The chemistry of this discharge stream has also been defined (Appendix C). The objective of optimizing the FNGP water network is to reduce the complexity of the required ZED system by limiting the size of the FNR discharge and the demand for raw water. Although recycling the flare water back into process would help to accomplish this goal, the inconsistent production of, and (to a lesser extent) the small concentration of hydrocarbons in the flare water make this option difficult. It would be more effective to simply deal with it as part of the effluent plant outflow. Since the other wastewater discharges listed in Table 5.3 similarly suffer from inconsistent flow rates (Shang Su, personal com.), the ZED treatment system will be designed around the combined FNR discharge.

### 5.4. Conclusion

The water flow and water chemistry data collected at the FNGP initially appeared to be of rather limited value in terms of their representative strength. Metered flow readings were average values derived from only one month's worth of data, and the mass balances were riddled with
inconsistencies. The January 1995 flow data were, however, generally representative of longer term trends, even if the accuracy of some flow meters remains in question (i.e. those surrounding the FESs and effluent plant). Similarly, although the regeneration wastes from the ion-exchangers were never properly sampled, calculations based on information from several of the plant's operating manuals provided a rough characterization of the blowdown waters leaving the FESs. When the mass balances were reconstructed, most of the original discrepancies either disappeared or could be attributed to contaminants in the sampled waters (e.g.. algae in the polishing pond samples) or conditioning chemicals added to the water system. In the end, the assembled data base provided sufficient detail to continue on with the ZED project.

Opportunities for optimizing the existing water network were limited. Only two water leaks were detected; one was in the steam tracing lines, while the other appeared to be a significant loss of wastewater from the effluent plant. It is unlikely that tightening the steam tracing system will have any impact on the plant's water flows, because only about $5 \mathrm{~m}^{3} /$ day of steam escapes through these vents. This represents less that $1 \%$ of current treated water production. Recovering the wastewater lost from the effluent plant, while not violating the first boundary condition limiting this project, will ultimately necessitate a larger ZED treatment system. Waste production from this system will also be larger than it would be with the current FNR discharge. Clearly, the original premise that minimizing the amount of water leaking out of the FNGP's water network would simplify the required ZED program does not apply in this case; closing off the 2 observed leaks is unlikely to benefit the ZED project, but the FNGP can only become a ZED facility if the leak at the effluent plant is eliminated.

The same cannot be said for reusing some of the relatively uncontaminated steam now discharged from this facility. Although most of the steam vents are too small to warrant their recapture, the 8 psi deaerator vent is an obvious exception. Reusing the $376 \mathrm{~m}^{3} /$ day of lost 8 psi steam would result in a net raw water savings of $448 \mathrm{~m}^{3} / \mathrm{day}$. Computer simulations also indicate that the FNR discharge will fall by $72 \mathrm{~m}^{3} /$ day, and that its incumbent contaminant mass will be
reduced by $291 \mathrm{~kg} /$ day when this water is reused. The only other steam vent which could offer similar savings is the vapour stream released through the thermo-oxidizer. Further research is first needed to investigate if this water can be reused without compromising either the plant's gas processing efficiency or the ability to install a ZED system at the FNGP.

Given the stringent water quality requirements of the process equipment, none of the wastewater discharged into the effluent system can be directly reused. Individually, these waters suffer from inconsistent flow rates. As a result, it is more efficient to focus on treating and recycling these water after they combine to form the FNR discharge. A ZED program for the FNGP was therefore designed around recycling the FNR discharge wherein no other changes were made to the water distribution network, except that the 8 psi steam now venting from the deaerators was condensed and returned, along with other condensate, to the boiler feedwater hotwells. The apparent effluent plant leak was not altered as the only further research will show if the observed water loss actually occurred, or if it. was a simple manifestation caused by faulty flow meters.
Table 5.1: A water balance of the Fort Nelson Gas Plant based on its individual operating units.

| Input ( $\mathrm{m}^{3} /$ day ) |  | Operation | Output (m³/day) |  |
| :---: | :---: | :---: | :---: | :---: |
| Powerhouse |  |  |  |  |
| - Treater system |  |  |  |  |
| Raw water | 859a | Production of soft water | Treated water (T.W.) | $1.01 \times 10^{3 \mathrm{a}}$ |
| 8 psi steam | $547{ }^{\text {a }}$ |  | \#3 Hot lime treater blowdown | $16.3^{\text {a }}$ |
|  |  |  | Venting off \#3 treater | $376^{\text {a }}$ |
| Brine solution | $39.1{ }^{\text {a }}$ | Softener regeneration | Used brine to lime pond | $39.1{ }^{\text {a }}$ |
| T.W. from \#3 treater | $17.7{ }^{\text {a }}$ |  | Dirty T.W. back to \#3 treater | $11.9{ }^{\text {a }}$ |
| Raw water | $6.8{ }^{\text {b }}$ | \#3 treater recirc. pump | Seal water to lime pond | $6.8{ }^{\text {b }}$ |
| Raw water | 866 | Net flows | Treated water | $1.01 \times 10^{3}$ |
| 8 psi steam | 547 |  | Discharge to lime pond | 62.2 |
| Brine solution | 39.1 |  | Venting | 376 |
|  | $1.45 \times 10^{3}$ |  |  | $1.45 \times 10^{3}$ |
| Input - output $=0.0 \quad$ System in balance |  |  |  |  |
| - Treated water tanks |  |  |  |  |
| Treated water | $1.01 \times 10^{3}$ | T.W. collection | T.W. to Process | $151{ }^{\text {a }}$ |
|  |  |  | Domestic filter backwash | $16.3^{\text {a }}$ |
|  |  |  | Softener regeneration brine | $39.1^{\mathrm{a}}-$ |
|  |  |  | T.W. to deaerators | 808 |
| - Deaerators |  |  |  |  |
| T.W. from T.W. tanks | 808 | Boiler feedwater formation | Boiler feedwater (BFW) to E/F/G/H | $296{ }^{\text {c }}$ |
| Condensate return | $1.15 \times 10^{4}$ |  | BFW to sulphur plant | $1.74 \times 10^{3 \mathrm{~d}}$ |
| 45 psi steam | $546{ }^{\text {e }}$ | Purge non-condensible gases | 8 psi venting | $546{ }^{\text {d }}$ |
|  |  |  | . BFW to boilers | $1.02 \times 10^{4}$ |

Table 5.1 (con't)

| Input ( $\mathrm{m}^{3} /$ day ) |  | Operation | Output ( $\mathrm{m}^{3} /$ day $)$ |  |
| :---: | :---: | :---: | :---: | :---: |
| Boilers |  |  |  |  |
| BFW from deaerators | $1.02 \times 10^{4}$ | 450 psi steam production | 450 psi steam | $9.76 \times 10^{3 \mathrm{c}}$ |
|  |  |  | Blowdown to lime pond | $81.6^{\mathrm{c}}$. |
| Raw water | $6.8{ }^{\text {b }}$ | \#5 boiler uniloc cooler | Spent cooling water | $6.8{ }^{\text {b }}$ |
|  | $\overline{1.02 \times 10^{4}}$ |  |  | $\overline{9.85 \times 10^{3}}$ |
| Input - output $=402$ Difference $<\mathbf{1 0 \%}$ of the input - System balances |  |  |  |  |
| - Power generation |  |  |  |  |
| 450 psi steam | $2.97 \times 10^{3 \mathrm{~d}}$ | Turbines | 45 psi steam | $2.97 \times 10^{3 \mathrm{a}}$ |
|  |  |  | Venting | $3.0{ }^{\text {a }}$ |
| 45 psi steam | $\underline{610^{\mathrm{e}}}$ | Preheaters | Condensate | $610^{\text {d }}$ |
|  | $3.58 \times 10^{3}$ |  |  | $\overline{3.58 \times 10^{3}}$ |
| Input - output $=1.0 \quad$ Difference $<\mathbf{1 0 \%}$ of the input - System balances |  |  |  |  |
| - Other |  |  |  |  |
| 45 psi steam | $501{ }^{\text {d }}$ | Steam tracing | Condensate | $496^{\text {a }}$ |
|  |  |  | Leaks | $5.0^{\text {a }}$ |
| 45 psi steam | $1.09 \times 10^{3 \mathrm{c}}$ | Excess 45 psi condenser | Condensate | $1.09 \times 10^{3 \mathrm{c}}$ |
| Raw \& Dom. water Water into $\mathrm{P} / \mathrm{H}$ lab | q | Cleaning, Showers.... <br> Water quality monitoring | Dirty R \& D water | q |
|  | $8.3{ }^{\text {b }}$ |  | Discharge to lime pond | $8.3{ }^{\text {b }}$ |
|  | $1.60 \times 10^{3}$ |  |  | $1.60 \times 10^{3}$ |
| Input -output $=0.0 \quad$ System in balance |  |  |  |  |

Table 5.1 (con't)

| Input ( $\mathrm{m}^{3} /$ day ) |  | Operation | Output (m³/day) |  |
| :---: | :---: | :---: | :---: | :---: |
| Process <br> - C/D Trains |  |  |  |  |
| 450 psi steam | $1.76 \times 10^{3 \mathrm{~d}}$ | Turbines | 45 psi steam for reboilers | $1.76 \times 10^{3 \mathrm{a}}$ |
|  |  |  | Venting | $1.8{ }^{\text {a }}$ |
| 45 psi steam | $3.06 \times 10^{3 \mathrm{c}}$ | Reboilers | Condensate | $3.06 \mathrm{e} 3^{\text {d }}$ |
| T.W. | $10.9{ }^{\text {a }}$ | T.W. cleaning | Dirty T.W. to effluent plant | $10.9{ }^{\text {a }}$ |
| Raw \& Dom. water | y | Cleaning, Showers.... | Dirty R \& D water | y |
| Condensate | $70.6^{\text {a }}$ | MEA/Carbonate make-up | Water in sweet gas | $0.4{ }^{\text {a }}$ |
| Sour water | v |  | Lost to flare pits | $7.3{ }^{\text {a }}$ |
|  |  |  | Vapour in wet acid gas | $62.9{ }^{\text {a }}$ |
|  |  |  | Liquid in wet acid gas | $v$ |
| 450 psi steam | $1.76 \times 10^{3}$ | Net flows | Venting | 1.8 |
| 45 psi steam | $1.30 \times 10^{3}$ |  | Condensate | $2.99 \times 10^{3}$ |
| T.W. | 10.9 |  | Dirty water to effluent plant | $10.9+y$ |
| Raw \& Dom. water | y |  | Water in sweet gas | 0.4 |
| Sour water | v |  | Lost to flare pits | 7.3 |
|  |  |  | Vapour in wet acid gas | 62.9 |
|  |  |  | Liquid in wet acid gas | v |
|  | $3.07 \times 10^{3}$ |  |  | $3.07 \times 10^{3}$ |
| Input - output $=0.0 \quad$ System in balance |  |  |  |  |
| - E/F/G/H Trains |  |  |  |  |
| 450 psi steam | $2.17 \times 10^{3 \mathrm{~d}}$ |  | Turbines 45 psi steam for reboilers | $2.17 \times 10^{3 \mathrm{a}}$ |
|  |  |  | Venting | $2.2{ }^{\text {a }}$ |

Table 5.1 (con't)

| Input ( $\mathrm{m}^{3} /$ day ) |  | Operation | Output (m³/day) |  |
| :---: | :---: | :---: | :---: | :---: |
| - E/F/G/H Trains (con't) |  |  |  |  |
| 45 psi steam | $5.47 \times 10^{3 \mathrm{c}}$ | Reboilers | Condensate | $5.47 \times 10^{3 \mathrm{c}}$ |
| T.W. | $10.9{ }^{\text {a }}$ | T.W. cleaning | Dirty T.W. to effluent plant | $10.9{ }^{\text {a }}$ |
| Raw \& Dom. water | f | Cleaning, Showers.... | Dirty R \& D water | f |
| 450 psi steam | $208{ }^{\text {c }}$ | 450 psi steam breakdown | 45 psi steam | $208{ }^{\text {c }}$ |
| BFW | $269{ }^{\text {c }}$ | Cooling superheated steam | 45 psi steam | $269{ }^{\text {c }}$ |
| T.W. | 129b | DEA make-up | Water in sweet gas | $0.7{ }^{\text {a }}$ |
| Sour water | w |  | Lost to flare pits | $13.3{ }^{\text {a }}$ |
|  |  |  | Vapour in wet acid gas | $115^{\text {a }}$ |
|  |  |  | Liquid in wet acid gas | w |
| 450 psi steam | $2.38 \times 10^{3}$ | Net flows | Venting | 2.2 |
| 45 psi steam | $2.80 \times 10^{3}$ |  | Condensate | $5.47 \times 10^{3}$ |
| BFW | 296 |  | Dirty water to effluent plant | $10.9+\mathrm{f}$ |
| T.W. | 140 |  | Water in sweet gas | 0.7 |
| Raw \& Dom. water | f |  | Lost to flare pits | 13.3 |
| Sour water | w |  | Vapour in wet acid gas | 115 |
|  |  |  | Liquid in wet acid gas | w |
|  | $\overline{5.61 \times 10^{3}}$ |  |  | $\overline{5.61 \times 10^{3}}$ |
| Input - output $=0.0 \quad$ System in balance |  |  |  |  |
| Booster Station 12 |  |  |  |  |
| 450 psi steam | $1.89 \times 10^{3 \mathrm{~d}}$ | Gas compression | 45 psi steam to header | $1.89 \times 10^{3 \mathrm{a}}$ |
|  |  |  | Venting | $1.9{ }^{\text {a }}$ |
|  | $1.89 \times 10^{3}$ | System in balance |  | $1.89 \times 10^{3}$ |

Table 5.1 (con't)

| Input ( $\mathrm{m}^{3} /$ day ) |  | Operation | Output ( $\mathrm{m}^{3} /$ day ) |  |
| :---: | :---: | :---: | :---: | :---: |
| Sulphur Plant |  |  |  |  |
| 450 psi steam | $686{ }^{\text {d }}$ | Gas blowers | 45 psi steam | 685 ${ }^{\text {a }}$ |
|  |  |  | Venting | $0.7{ }^{\text {a }}$ |
| BFW | $447{ }^{\text {a }}$ | Reaction furnace | 150 psi steam | 425a |
|  |  |  | Blowdown to lime pond | $21.8{ }^{\text {a }}$ |
| 150 psi steam | $405{ }^{\text {d }}$ | Air blowers | 5 psi steam | 405 ${ }^{\text {d }}$ |
|  |  |  | Venting | $0.4{ }^{\text {a }}$ |
| - | - | - | 150 psi steam to Petrosul | $19.6{ }^{\text {c }}$ |
| BFW | $438{ }^{\text {a }}$ | \#1 condensers | 45 psi steam | $422^{\text {a }}$ |
|  |  |  | Blowdown to lime pond | $16.3^{\text {a }}$ |
| BFW | $857{ }^{\text {a }}$ | \#2 condensers | 45 psi steam | $446{ }^{\text {d }}$ |
|  |  |  | 15 psi steam | $378{ }^{\text {d }}$ |
|  |  |  | Blowdown to lime pond | $32.7{ }^{\text {a }}$ |
| 15 psi steam | $378{ }^{\text {d }}$ | 15 psi condenser | Condensate | $378{ }^{\text {d }}$ |
| 5 psi steam | 405 ${ }^{\text {d }}$ | 5 psi condenser | Condensate | 405 ${ }^{\text {d }}$ |
| Sour water | v+w | Wet acid gas purification | Sour water | v+w |
| T.W./Condensate | $178{ }^{\text {a }}$ |  | T.W./Condensate lost to T/O | $178{ }^{\text {a }}$ |
|  |  |  | Water formed in S formation | $260^{\text {a }}$ |
| 450 psi steam | 686 | Net flows | 150 psi steam | 19.6 |
| BFW | $1.74 \times 10^{3}$ |  | 45 psi steam | $1.55 \times 10^{3}$ |
| Sour water | $\mathrm{v}+\mathrm{w}$ |  | Condensate | 783 |
| T.W./Condensate | 178 |  | Blowdown to lime pond | 70.8 |
|  |  |  | Venting | 1.1 |
|  |  |  | Sour water | v+w |
|  |  |  | Water lost in acid gas to T/O | 438 |

Table 5.1 (con't)

| Input ( $\mathrm{m}^{3} /$ day ) |  | Operation | Output (m³/day) |  |
| :---: | :---: | :---: | :---: | :---: |
| Sulphur Plant (con't) |  |  |  |  |
|  | $2.61 \times 10^{3}$ |  |  | $\overline{2.87 \times 10^{3}}$ |
| Input - output $=$ | -260 D | Difference $<10 \%$ of the inpu | m balances |  |
| Thermo-oxidizer |  |  |  |  |
| 450 psi steam | $280{ }^{\text {d }}$ | Blowers | 45 psi steam | $280{ }^{\text {a }}$ |
|  |  |  | Venting | $0.3{ }^{\text {a }}$ |
| 45 psi steam | $43.5{ }^{\text {d }}$ | Preheaters | Condensate | $43.5{ }^{\text {d }}$ |
| Water in acid gas | $438{ }^{\text {a }}$ | Burn \& release of acid gas | Water released out the stack | $438{ }^{\text {a }}$ |
| 450 psi steam | 280 | Net flows | 45 psi steam | 236 |
| Water in acid gas | 438 |  | Venting | 0.3 |
|  |  |  | Condensate | 43.5 |
|  |  |  | Water released out the stack | 438 |
|  | 718 |  |  | 718 |
| Input - output= | 0.0 S | System in balance |  |  |
| Lime Pond |  |  |  |  |
| S/P blowdown | $70.8{ }^{\text {a }}$ | Blowdown collection | Discharge to polishing pond | $238{ }^{\text {c }}$ |
| \#3 Hot lime treater | $16.3^{\text {a }}$ |  |  |  |
| Used softerner brine | $39.1{ }^{\text {a }}$ |  |  |  |
| Boiler blowdown | $81.6{ }^{\text {c }}$ |  |  |  |
| Dom. filter backwash | $16.3^{\text {a }}$ |  |  |  |
| $\mathrm{P} / \mathrm{H}$ lab sample lines | $8.3{ }^{\text {b }}$ |  |  |  |
| \#3 pump water seal | $6.8{ }^{\text {b }}$ |  |  |  |

Table 5.1 (con't)
Lime Pond (con't)


| 8 | 0 |
| :--- | :--- |
| 0 |  |
| 0 |  | | $\stackrel{\sim}{n}$ |  |
| :---: | :---: |
| $\underset{n}{n}$ |  |

a Calculated or assumed value - see Appendix B for description of its derivation b Number supplied by John Martin - FNGP employee
c Metered reading = average from January 1995 data
${ }^{\mathrm{d}}$ Number supplied by Shang Su - Process Engineer at FNGP
Table 5.2: A water balance of the Fort Nelson Gas Plant based on the different types of water used at the facility

| Origin (m³/day) | Used in/as ( $\mathrm{m}^{3} /$ day $)$ |  |  | Becomes |
| :---: | :---: | :---: | :---: | :---: |
| Raw water |  |  |  |  |
| Reservoir inflow | $1.00 \times 10^{3 \mathrm{a}}$ | \#5 uniloc cooling water | $6.8{ }^{\text {b }}$ | Wastewater to lime pond |
|  |  | \#3 recir. pump seal | $6.8{ }^{\text {b }}$ | Wastewater to lime pond |
|  |  | Cleaning \& Domestic water | $137{ }^{\text {a }}$ | Wastewater to E/P |
|  |  | Treater system | $859{ }^{\text {a }}$ | Treated water |
|  |  | Export to Petrosul | $32.7{ }^{\text {c }}$ | - |
|  | $\overline{1.00 \times 10^{3}}$ |  | $\overline{1.04 \times 10^{3}}$ |  |
| Input - output $=-38.7$ Difference $>\mathbf{1 0 \%}$ input |  |  |  |  |
| Treated water |  |  |  |  |
| Raw water | $859{ }^{\text {a }}$ | \#3 treater blowdn. | $16.3{ }^{\text {a }}$ | Wastewater to lime pond |
| Condensed 8 psi steam | 547 a | \#3 treater venting | $376{ }^{\text {a }}$ | Lost to atmosphere |
|  |  | Softener regeneration | $39.1{ }^{\text {a }}$ | Wastewater to lime pond |
|  |  | Domestic filter backwash | $16.3^{\text {a }}$ | Wastewater to lime pond |
|  |  | Cleaning water | $21.8{ }^{\text {a }}$ | Wastewater to E/P |
|  |  | DEA make-up water in E/F/G/H process trains | 129a | Lost through: <br> - Stack, Flare \& Sweet gas |
|  | $1.41 \times 10^{3}$ |  | 598 |  |
| Going to hotwells $=808$ |  |  |  |  |
| Boiler feedwater |  |  |  |  |
| Treated water | 808 | Cooling water for 45 psi | $269{ }^{\text {c }}$ | 45 psi steam |
| Condensate return | $1.15 \times 10^{3}$ | steam in $\mathrm{E} / \mathrm{F} / \mathrm{G} / \mathrm{H}$ Steam production in S/P | $1.74 \times 10^{3 \mathrm{~d}}$ | 150 psi \& 45 psi steam |

Table 5.2 (con't)

Table 5.2 (con't)

| Origin ( $\mathrm{m}^{3} /$ day $)$ |  | Used in/as ( $\mathrm{m}^{3} /$ day ) |  | Becomes |
| :---: | :---: | :---: | :---: | :---: |
| 150 psi steam (con't) |  |  |  |  |
|  | 425 | Shipped to Petrosul | $\frac{19.6^{\mathrm{c}}}{425}-$ | - |
| Input -output $=0.0 \quad$ System in balance |  |  |  |  |
| 45 psi steam |  |  |  |  |
| $\mathrm{P} / \mathrm{H}$ turbines | $2.96 \times 10^{3 \mathrm{a}}$ | C/D reboilers | $3.06 \times 10^{3 \mathrm{c}}$ | Condensate |
| C/D turbines | $1.76 \times 10^{3 \mathrm{a}}$ | E/F/G/H/ reboilers | $5.47 \times 10^{3 \mathrm{c}}$ | Condensate |
| E/F/G/H turbines | $2.17 \times 10^{3 \mathrm{a}}$ | $\mathrm{P} / \mathrm{H}$ preheaters | $610^{\text {d }}$ | Condensate |
| S/P gas blowers | $685{ }^{\text {a }}$ | T/O preheaters | $43.5{ }^{\text {d }}$ | Condensate |
| T/O blowers | $279{ }^{\text {a }}$ | Steam tracing | $501{ }^{\text {d }}$ | Condensate |
| Booster St. 12 turbines | $1.89 \times 10^{3 \mathrm{a}}$ | Venting on steam tracing | $5.0{ }^{\text {a }}$ | Lost to atmosphere |
| Breakdn. stations | $208{ }^{\text {c }}$ | X-cess 45 psi condenser | $1.09 \times 10^{3 \mathrm{c}}$ | Condensate |
| S/P - Condensers | $868{ }^{\text {d }}$ | Purge steam to deaerators | $547{ }^{\text {d }}$ | 8 psi steam |
| B.F.W. to E/F/G/H | ${ }^{2988}{ }^{\text {c }}$ |  |  |  |
|  | $1.11 \times 10^{4}$ |  | $1.13 \times 10^{4}$ |  |
| Input - output | -198 | ence $<10 \%$ of the input | ystem bala |  |
| 15 psi steam |  |  |  |  |
| S/P - \#2 condensers | $380^{\text {a }}$ | 15 psi condensers | $380^{\text {a }}$ | Condensate |
| Input - output | 0.0 S | $m$ in balance |  |  |
| 8 psi steam |  |  |  |  |
| Vented from deaerators | 547a | \#3 hot lime treater | $171{ }^{\text {a }}$ | Treated water |

Table 5.2 (con't)

a Calculated or assumed value - see Appendix B for description of its derivation b Number supplied by John Martin - FNGP employee
c Metered reading $=$ average from January 1995 data
${ }^{\text {d Number supplied by Shang Su - Process Engineer at FNGP }}$

Table 5.3: A summary of treated water losses from the Fort Nelson Gas Plant.

| Source | Rate of loss ( $\mathrm{m}^{3} /$ day $)$ | Destination |
| :---: | :---: | :---: |
| Venting |  |  |
| 8 psi steam off the deaerators | 376 | Atmosphere |
| Turbine vents: |  |  |
| Powerhouse | 3.0 | Atmosphere |
| Process | 3.9 | Atmosphere |
| Sulphur plant | 1.1 | Atmosphere |
| Thermo-oxidizer | 0.3 | Atmosphere |
| Booster Station 12 | 1.9 | Atmosphere |
| Steam tracing | $\begin{array}{r} 5.0 \\ \mathbf{3 9 1} \end{array}$ | Atmosphere |
| Export |  |  |
| 150 psi steam | $\frac{19.6}{19.6}$ | Petrosul |
| Discharge |  |  |
| Blowdown |  |  |
| - Boilers | 81.6 | Lime pond |
| Sulphur plant | 70.8 | Lime pond |
| Hot lime treater | 16.3 | Lime pond |
| Domestic backwash | 16.3 | Lime pond |
| Softener regeneration | 39.1 | Lime pond |
| Filter cleaning | 21.8 | Effluent plant |
| Process water | 20.7 | Flare pits |
|  | 178 | Thermo-oxidizer |
|  | 1.0 | Sweet gas |
|  | 445 |  |
| Tota <br> Treated wate | $\begin{array}{lc} \text { oss }= & \mathbf{8 5 6} \\ \text { put }= & 808 \\ \hline-47.9 \end{array}$ |  |

Table 5.4: A mass balance on the front-end softening system.

| Parameter | Total in ${ }^{\text {a }}$ | Total out ${ }^{\text {a }}$ | Difference | \% diff. | Explanation |
| :---: | :---: | :---: | :---: | :---: | :---: |
| Water flow ( $\mathrm{m}^{3} /$ day) | $1.45 \times 10^{3}$ | $1.45 \times 10^{3}$ | 0 | 0 |  |
| Total Alkalinity (kg/day) | 162 | 43.1 | 118 | 73 | mis-sampled |
| Solids (kg/day) |  |  |  |  |  |
| Total | - | - | - | - |  |
| Suspended | - | - | - | - |  |
| Dissolved | - | - | - | - |  |
| Carbon content (kg/day) |  |  |  |  |  |
| Total | 54.8 | 26.1 | 28.7 | 52 | mis-sampled |
| Inorganic | 44.4 | 17.5 | 26.9 | 61 | mis-sampled |
| Organic | 10.4 | 8.7 | 1.7 | 16 | mis-sampled |
| Metals (kg/day) |  |  |  |  |  |
| Calcium | 70.0 | 2.8 | 67.1 | 96 | mis-sampled |
| Magnesium | 15.8 | 0.7 | 15.1 | 95 | mis-sampled |
| Sodium | 154 | 59.6 | 94.1 | 61 | mis-sampled |
| Iron | 0.5 | 0.7 | -0.2 | -50 | mis-sampled |
| Inorganics (kg/day) |  |  |  |  |  |
| Phosphates | 0.1 | 0.1 | 0.0 | -13 | mis-sampled |
| Chlorides | 217 | 3.0 | 214 | 99 | mis-sampled |
| Sulfates | 78.6 | 88.3 | -9.6 | -12 | mis-sampled |
| Silica | 4.6 | 1.0 | 3.6 | 78 | mis-sampled |

[^1]Table 5.5: A mass balance on the lime ponds.

| Parameter | Total in ${ }^{\text {a }}$ | Total out ${ }^{\text {a }}$ | Difference | \% diff. | Explanation ${ }^{\text {c }}$ |
| :---: | :---: | :---: | :---: | :---: | :---: |
| Water flow ( $\mathrm{m}^{3} /$ day $)$ | 246 | 238 | 8.0 | 3 |  |
| Total Alkalinity (kg/day) | 51.3 | 17.3 | 34.0 | 66 | precipitated |
| Solids (kg/day) |  |  |  |  |  |
| Total | 218 | 610 | -393 | -180 | FESs error |
| Suspended | 2.2 | 1.6 | 0.5 | 25 | precipitated |
| Dissolved | 134 | 609 | -475 | -354 | FESs error |
| Carbon content (kg/day) |  |  |  |  |  |
| Total | 12.7 | 6.4 | 6.3 | 50 | precipitated |
| Inorganic | 5.2 | 2.5 | 2.7 | 53 | precipitated |
| Organic | 7.6 | 3.9 | 3.7 | 49 | precipitated |
| Metals (kg/day) |  |  |  |  |  |
| Calcium | 2.5 | 52.7 | -50.1 | $-1.97 \times 10^{3}$ | FESs error |
| Magnesium | 0.5 | 4.7 | -4.2 | -808 | FESs error |
| Sodium | 64.6 | 149 | -84.6 | -131 | FESs error |
| Iron | 0.2 | 0.0 | 0.2 | 82 | precipitated |
| Inorganics (kg/day) |  |  |  |  |  |
| Phosphates | 1.1 | 0.0 | 1.1 | 98 | precipitated |
| Chlorides | 2.7 | 217 | -214 | $-7.90 \times 10^{3}$ | FESs error |
| Sulfates | 97.0 | 148 | -50.5 | -52 | FESs error |
| Silica | 1.0 | 0.9 | 0.1 | 6 | insig. difference |
| a Refer to Appendix D for calculations |  |  |  |  |  |
| ${ }^{\mathrm{b}}$ precipitated $=$ missing mass settled out of solution within the lime ponds; FESs error $=$ front end softeners mis-sampled; insig. difference $=$ insignificant difference |  |  |  |  |  |

Table 5.6: A mass balance on the boilers.

| Parameter | Total in ${ }^{\text {a }}$ | Total out ${ }^{\text {a }}$ | Difference | \% diff. | Explanation ${ }^{\text {b }}$ |
| :---: | :---: | :---: | :---: | :---: | :---: |
| Water flow ( $\mathrm{m}^{3} /$ day ) | $1.03 \times 10^{4}$ | $9.85 \times 10^{3}$ | 402 | 4 |  |
| Total Alkalinity (kg/day) | 104 | 75.5 | 28.2 | 28 | ? |
| Solids (kg/day) |  |  |  |  |  |
| Total | 840 | 828 | 12.5 | 1 | insig. difference |
| Suspended | 7.6 | 3.1 | 4.5 | 59 | ? |
| Dissolved | 832 | 824 | 8.0 | 1 | insig. difference |
| Carbon content (kg/day) |  |  |  |  |  |
| Total | 67.7 | 63.0 | 4.6 | 7 | insig. difference |
| Inorganic | 26.9 | 20.1 | 6.8 | 25 | ? |
| Organic | 40.7 | 42.9 | -2.2 | -5 | insig. difference |
| Metals (kg/day) |  |  |  |  |  |
| Calcium | 3.0 | 2.3 | 0.7 | 22 | ? |
| Magnesium | 1.8 | 1.6 | 0.2 | 9 | insig. difference |
| Sodium | 61.7 | 53.1 | 8.7 | 14 | chemical add'n |
| Iron | 2.2 | 2.1 | 0.2 | 8 | insig. difference |
| Inorganics (kg/day) |  |  |  |  |  |
| Phosphates | 0.5 | 1.1 | -0.6 | -126 | chemical add'n |
| Chlorides | 2.5 | 2.5 | 0.0 | 0 |  |
| Sulfates | 67.7 | 79.4 | -11.7 | -17 | chemical add'n |
| Silica | 1.5 | 1.6 | -0.1 | -3 | insig. difference |

a Refer to Appendix D for calculations
b chemical add' $\mathrm{n}=$ changes induced by addition BFW polishing chemicals; ? = unknown source or sink; insig. difference $=$ insignificant difference

Table 5.7: A mass balance on the sulphur plant.

| Parameter | Total in ${ }^{\text {a }}$ | Total out ${ }^{\text {a }}$ | Difference | \% diff. | Explanation ${ }^{\text {b }}$ |
| :---: | :---: | :---: | :---: | :---: | :---: |
| Water flow ( $\mathrm{m}^{3} /$ day) | $2.61 \times 10^{3}$ | $2.87 \times 10^{3}$ | 260 | 10 |  |
| Total Alkalinity (kg/day) | 22.0 | 22.9 | -0.9 | -4 | insig. difference |
| Solids (kg/day) |  |  |  |  |  |
| Total | 197 | 224 | -26.2 | -13 | insig. difference |
| Suspended | 1.5 | 0.5 | 1.0 | 65 | ? |
| Dissolved | 196 | 223 | -27.2 | -14 | insig. difference |
| Carbon content (kg/day) |  |  |  |  |  |
| Total | 16.0 | 17.1 | -1.1 | -7 | insig. difference |
| Inorganic | 6.1 | 5.3 | 0.8 | 13 | chemical add'n |
| Organic | 10.0 | 11.9 | -1.9 | -19 | ? |
| Metals (kg/day) |  |  |  |  |  |
| Calcium | 0.7 | 0.6 | 0.1 | 17 | ? |
| Magnesium | 0.4 | 0.4 | 0.0 | 4 | insig. difference |
| Sodium | 11.8 | 18.7 | -6.8 | -58 | chemical add'n |
| Iron | 0.5 | 0.5 | 0.0 | 8 | insig. difference |
| Inorganics (kg/day) |  |  |  |  |  |
| Phosphates | 0.1 | 0.5 | -0.4 | -333 | chemical add'n |
| Chlorides | 0.5 | 0.9 | -0.4 | -70 | ? |
| Sulfates | 13.1 | 29.8 | -16.8 | -128 | chemical add'n |
| Silica | 0.3 | 0.5 | -0.1 | -45 | ? |
| a Refer to Appendix D for calculations |  |  |  |  |  |
| b chemical add' $\mathrm{n}=$ changes induced by addition BFW polishing chemicals; ? = unknown source or sink; insig. difference $=$ insignificant difference |  |  |  |  |  |

Table 5.8: A mass balance on the polishing pond.

| Parameter | Total in ${ }^{\text {a }}$ | Total out ${ }^{\text {a }}$ | Difference | \% diff. | Explanation ${ }^{\text {b }}$ |
| :---: | :---: | :---: | :---: | :---: | :---: |
| Water flow (m³/day) | 347 | 352 | 5 | 1 |  |
| Total Alkalinity (kg/day) | 36.7 | 35.3 | 1.4 | 4 | insig. difference |
| Solids (kg/day) |  |  |  |  |  |
| Total | 922 | 973 | -51.0 | -6 | insig. difference |
| Suspended | 2.9 | 9.5 | -6.6 | -232 | algae |
| Dissolved | 919 | 964 | -44.3 | -5 | insig. difference |
| Carbon content (kg/day) |  |  |  |  |  |
| Total | 28.4 | 34.3 | -5.9 | -21 | algae |
| Inorganic | 8.9 | 10.1 | -1.2 | -13 | insig. difference |
| Organic | 19.5 | 24.2 | -4.8 | -25 | algae |
| Metals (kg/day) |  |  |  |  |  |
| Calcium | 58.7 | 51.3 | 7.4 | 13 | insig. difference |
| Magnesium | 6.6 | 7.2 | -0.6 | -9 | insig. difference |
| Sodium | 210 | 205 | 4.9 | 2 | insig. difference |
| Iron | 0.4 | 0.6 | -0.2 | -41 | algae |
| Inorganics (kg/day) |  |  |  |  |  |
| Phosphates | 0.0 | 0.1 | -0.1 | -130 | algae |
| Chlorides | 239 | 220 | 18.9 | 8 | insig. difference |
| Sulfates | 183 | 191 | -7.6 | -4 | insig. difference |
| Silica | 1.7 | 1.8 | -0.2 | -10 | insig. difference |

a Refer to Appendix D for calculations
b algae $=$ changes induced by algae in polishing pond; insig. difference $=$ insignificant difference

Table 5.9: A comparison between water flow readings from January 1995 and overall averages.

| Flow Path | January |  | Overall |  |  | \% <br> Difference |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  | Avg. | C.V. ${ }^{\text {a }}$ | Avg. | C.V. ${ }^{\text {a }}$ | Months ${ }^{\text {b }}$ |  |
| Reservoir - m³/day | $1.0 \times 10^{3}$ | 0.44 | $1.1 \times 10^{3}$ | 0.39 | 14 | 9 |
| Raw water inlet - $\mathrm{m}^{3} /$ day | 538 | 0.62 | 623 | 0.52 | 14 | 14 |
| Treated water $-\mathrm{m}^{3} /$ day | 724 | 0.61 | 854 | 0.50 | 14 | 15 |
| Boiler \#5 - KLBH |  |  |  |  |  |  |
| Steam production | 297 | 0.04 | 285 | 0.13 | 14 | -4 |
| Blowdown | 1.8 | 0.40 | 2.1 | 0.38 | 2 | 14 |
| Boiler \#6-KLBH |  |  |  |  |  |  |
| Steam production | 301 | 0.02 | 276 | 0.23 | 14 | -9 |
| Blowdown | 3.1 | 0.19 | 3.1 | 0.21 | 2 | 0 |
| Boiler \#7 - KLBH |  |  |  |  |  |  |
| Steam production | 299 | 0.02 | 281 | 0.12 | 14 | -6 |
| Blowdown | 2.6 | 0.25 | 2.7 | 0.25 | 2 | 4 |
| 450 psi steam breakdn. - KLBH |  |  |  |  |  |  |
| $\mathrm{E} / \mathrm{F}$ trains | 0.2 | 1.37 | 0.5 | 1.39 | 2 | 60* |
| G/H trains | 18.9 | 0.56 | 17.0 | 0.53 | 2 | -11 |
| Reboiler steam - KLBH |  |  |  |  |  |  |
| C train - carbonate | 103 | 0.05 | 106 | 0.18 | 5 | 3 |
| C train - MEA | 33.7 | 0.04 | 32.0 | 0.18 | 5 | 5 |
| D train - carbonate | 109 | 0.05 | 107 | 0.14 | 5 | 2 |
| D train - MEA | 35.0 | 0.03 | 33.1 | 0.12 | 5 | 6 |
| E train - DEA | 125 | 0.06 | 128 | 0.08 | 5 | 2 |
| F train - DEA | 120 | 0.04 | 126 | 0.08 | 5 | 5 |
| G train - DEA | 131 | 0.03 | 129 | 0.08 | 5 | 2 |
| H train - DEA | 127 | 0.05 | 128 | 0.07 | 5 | 1 |
| Processed gas - mmscfd | 569 | 0.03 | 569 | 0.04 | 5 |  |
| Stack emissions - mmscfd | 6.2 | 0.16 | 7.3 | 0.18 | 14 | $15^{*}$ |

Table 5.9 (con't)

|  | January <br> Flow Path |  | C.V.a | Avg. | Overall <br> C.V.a | Months ${ }^{\text {b }}$ |
| :--- | :---: | :---: | :---: | :---: | :---: | :---: | | \% |
| :---: |
| Difference |

${ }^{\text {a C.V. }}$ = Coefficient of variation $=$ standard deviation $/$ average
b Number of months over which flow readings were recorded and averaged

* Significant difference between the January 1995 and overall averages ( $\mathrm{t}_{\text {statistic }}>1.96, \mathrm{p}<$ 0.05)

Table 5.10: Water balances on the raw water storage tank and front-end softeners (FESs) using metered and calculated inflow and ouflow volumes from the FESs.

| Flow path | Metered values ( $\mathrm{m}^{3} /$ day) | Calculated values ( $\mathrm{m}^{3} /$ day) |
| :---: | :---: | :---: |
| Raw water storage tank |  |  |
| Inflow |  |  |
| Reservoir pipline | $1.00 \times 10^{3}$ | $1.00 \times 10^{3}$ |
| Outflows |  |  |
| \#5 unlioc cooling water | 6.8 | 6.8 |
| Pump sealant | 6.8 | 6.8 |
| Cleaning \& domestic systems | 137 | 137 |
| Treated water system | 538 | 859 |
| Export to Petrosul | 32.7 | 32.7 |
| Inflow - outflow = | $=282$ | -38.7 |
| Front-end softeners |  |  |
| Inflow |  |  |
| Raw water storage tank | 538 | 859 |
| 8 psi steam | 547 | 547 |
|  | $1.08 \times 10^{3}$ | $1.41 \times 10^{3}$ |
| Outflows |  |  |
| Hot lime treater blowdown | 16.3 | 16.3 |
| Ion-exchanger regeneration | 39.1 | 39.1 |
| Cleaning water | 21.8 | 21.8 |
| Stripping solution make-up | 129 | 129 |
| Steam venting off lime treater | 360 | 376 |
| Domestic filter backwash | 16.3 | 16.3 |
|  | 583 | 599 |
| Product water volume = | $=503$ | 808 |
| Total treated water losses ${ }^{\text {a }}=$ | $=\quad 856$ | 856 |
|  | -353 | -47.9 |

[^2]Table 5.11: Recalculated mass balance on the FNGP's front-end softening system.

| Parameter | Total in ${ }^{\text {a }}$ | Total out ${ }^{\text {a }}$ | Difference | \% diff. | Explanation ${ }^{\text {b }}$ |
| :---: | :---: | :---: | :---: | :---: | :---: |
| Water flow ( $\mathrm{m}^{3} /$ day) | $1.0 \times 10^{3}$ | $1.0 \times 10^{3}$ | 0.0 | 0 |  |
| Carbon content |  |  |  |  |  |
| Inorganic (mol/day) | $3.2 \times 10^{3}$ | $3.3 \times 10^{3}$ | -95 | -3 | insig. difference |
| Organic (kg/day) | 8.7 | 6.6 | 2.0 | 23 | ? |
| Metals (kg/day) |  |  |  |  |  |
| Calcium | 69.9 | 69.8 | 0.1 | 0 |  |
| Magnesium | 15.7 | 15.7 | 0.1 | 0 |  |
| Sodium | 11.0 | 141 | -130 | -1181 | Ion-x regeneration |
| Iron | 0.4 | 0.6 | -0.3 | -69 | ? |
| Inorganics (kg/day) |  |  |  |  |  |
| Phosphates | 0.1 | 0.1 | 0.0 | 0 |  |
| Chlorides | 2.5 | 217 | -215 | -8554 | Ion-x regeneration |
| Sulfates | 74.8 | 84.2 | -9.4 | -13 | insig. difference |
| Silica | 4.5 | 4.5 | 0.0 | 0 |  |

[^3]Table 5.12: Recalculated mass balance on the FNGP's lime ponds.

| Parameter | Total in ${ }^{\text {a }}$ | Total out ${ }^{\text {a }}$ | Difference | \% diff. | Explanation ${ }^{\text {b }}$ |
| :---: | :---: | :---: | :---: | :---: | :---: |
| Water flow ( $\mathrm{m}^{3} /$ day $)$ | 246 | 238 | 7.6 | 3 | insig. difference |
| Carbon content |  |  |  |  |  |
| Inorganic (mol/day) | $3.2 \times 10^{3}$ | $2.4 \times 10^{2}$ | $2.9 \times 10^{2}$ | 93 | precipitated |
| Organic (kg/day) | 7.3 | 3.9 | 3.4 | 47 | precipitated |
| Metals (kg/day) |  |  |  |  |  |
| Calcium | 69.6 | 52.6 | 17.0 | 24 | precipitated |
| Magnesium | 15.6 | 4.7 | 10.9 | 70 | precipitated |
| Sodium | 148 | 149 | -1.0 | 0 |  |
| Iron | 0.2 | 0.0 | 0.2 | 100 | precipitated |
| Inorganics (kg/day) |  |  |  |  |  |
| Phosphates | 1.1 | 0.0 | 1.1 | 100 | precipitated |
| Chlorides | 217 | 217 | 0.0 | 0 |  |
| Sulfates | 97.6 | 148 | -49.9 | -51 | ? |
| Silica | 4.5 | 0.9 | 3.6 | 80 | precipitated |

[^4]Table 5.13: Changes that occurred in selected areas of the Fort Nelson Gas Plant's water system before and after water observed to escaping from the effluent treatment plant ( $\mathrm{E} / \mathrm{P}$ ) was recovered.

| Parameter | E/P leak untouched | E/P leak closed |
| :---: | :---: | :---: |
| Raw water flow ( $\mathrm{m}^{3} /$ day $)$ |  |  |
| - initial | 407 | 407 |
| - with wastewater recycling | 138 | 90.1 |
| FNR discharge flow ( $\mathrm{m}^{3} /$ day ) |  |  |
| - initial | 282 | 332 |
| - with wastewater recycling | 240 | 285 |
| ZED solid waste output (kg/day) |  |  |
| - initial | 415 | 483 |
| - with wastewater recycling | 233 | 294 |

Table 5.14: Selected water flows in the Fort Nelson Gas Plant's water distribution network with and without 8 psi steam reuse.

|  | Current system <br> $\left(\mathrm{m}^{3} / \mathrm{day}\right)$ | $\mathbf{8}$ psi steam reused <br> $\left(\mathrm{m}^{3} / \mathrm{day}\right)$ | Difference <br> $\left(\mathrm{m}^{3} / \mathrm{day}\right)$ |
| :--- | :---: | :---: | :---: |
| Flow path | 855 | 407 | $\mathbf{4 4 8}$ |
| Raw water inflow | 171 |  |  |
| 8 psi steam flow to lime treater | 171 | 0.0 | $\mathbf{3 7 6}$ |
| Lost 8 psi steam | 376 | 13.9 | 3.1 |
| Lime treater blowdown | 17.0 | 19.0 | 19.9 |
| Ion-exchange blowdown | 38.9 | 54.6 | 26.0 |
| Boiler blowdown | 80.6 | 46.8 | 23.1 |
| Sulphur plant blowdown | 69.9 | 173 | 72.0 |
| Lime pond outflow | 245 | 109 |  |
| Effluent plant outflow | 109 | 282 | $\mathbf{7 2 . 0}$ |
| Polishing pond | 354 | 422 | $\mathbf{2 9 1}$ |
| Discharged contaminant mass <br> (kg/day) | 713 |  |  |

Table 5.15: The average inlet gas profile for January 1994 at the Fort Nelson Gas Plant.

| Substance | Relative abundance |
| :--- | :---: |
| Nitrogen $\left(\mathrm{N}_{2}\right)$ | 0.0103 |
| Hydrogen sulfide $\left(\mathrm{H}_{2} \mathrm{~S}\right)$ | 0.0168 |
| Carbon dioxide $\left(\mathrm{CO}_{2}\right)$ | 0.1241 |
| Organic carbon compounds: | 0.8456 |
| 1 - carbon elements (C1) | 0.0023 |
| 2 - carbon elements (C2) | 0.0004 |
| 3 - carbon elements (C3) | 0.0002 |
| 4 - carbon elements (C4) | 0.0001 |
| 5 - carbon elements (C5) | 0.0002 |
| $6+$ - carbon elements (C6+) |  |
|  | Average gas volume $=18707000 \mathrm{~m}^{3} / \mathrm{day}$ |
|  |  |

### 6.0 RECYCLING TECHNOLOGY FOR THE FNGP

### 6.1 Re-evaluation of the Literature Review

Three principle treatment technologies are available to the FNGP: distillation, ion-exchange and membrane filtration. Each systems functions optimally within a given total dissolved solids (TDS) range (Fig. 3.1). The FNR discharge had a TDS concentration of $2700 \mathrm{mg} / \mathrm{L}$ (Table 6.1). According to Figure 3.1, this level of contamination would be best treated with membrane filtration.

As previously discussed, there are two types of membrane filters: ion-selective and water permeable. ED and EDR are the best ion-selective units. RO is the most suitable water permeable membrane for the FNGP, considering the large amounts mono- and divalent ions which need to be removed from the FNR discharge before it can be reused (Table 6.1). A preliminary comparison between the 3 alternatives indicated that RO, while more sensitive to fouling than both ED and EDR, generally produces the cleanest permeate of the 3 alternatives. Furthermore, despite the higher product water output with ED and EDR, RO units are simpler and easier to operate than ED and EDR. As a result, RO was chosen to be the best membrane filtration system.

A further argument to support the use of RO at the FNGP instead of ED or EDR concerns personal safety. As most of the equipment used in RO is already in use at this facility, plant personnel are already familiar with the potential dangers associated with RO (e.g. moving parts on pumps and valves, weakened pipes bursting...). On the other hand, the driving force in electrodialysis is electricity. The abundance of electrical current and highly conductive waters in, and potentially around, ED/EDR stacks is a definite health hazard, one that does not currently exist at the plant. So, while installing a RO filter is unlikely to introduce any new hazards into the FNGP, incorporating an ED or EDR unit into the ZED program would open plant personnel up to increased heath risks. Given the inherent advantages of RO over ED and EDR, it was selected as the most suitable treatment technology for the FNGP.

The RO membranes should themselves be of a spiral wound design, as this is the best of the available configurations (see Section 3.3.4). At this stage, it makes little difference if cellulose acetate of polyamide filters are used; they both provided the same level of treatment. When the final ZED treatment train is developed, filter type will become important, because it may influence the layout of the RO pretreatment system (see Section 3.3.4).

According to the literature, ZED plants generally dispose of RO concentrates in either evaporation ponds of by solidification (see Section 3.4). Evaporation ponds are only effective in climates where ambient air conditions encourage water evaporation. This is clearly not the case in Fort Nelson ( $58^{\circ} \mathrm{N} \times 124^{\circ} \mathrm{W}$ ) for a better part of the year. Winter temperature are generally well below $0^{\circ} \mathrm{C}$, and winter can last from October to May. RO concentrates will have to be disposed of by a more active approach.

A brine concentrator followed up with a spray dryer or crystallizer seems to be the proven method for solidifying RO concentrates (see Section 3.4). Spray dryers produce a solid effluent, while waste products from a crystallizer still require some processing before they can be landfilled. The fine, powdery solids exiting a spray dryer can, however, be difficult to handle. Given the reassurance of manufacturers that this is not likely to be a problem in a properly designed system (RCC - personal com.), and their lower costs, spray dryers appear to be the more appropriate technology to follow-up a brine concentrator as part of the ZED program at the FNGP.

Instead of using a BCS (brine concentrator - spray dryer) assembly, RO concentrates can, as suggested by some Westcoast personnel, be disposed of in a deep well. In any case, re-evaluating the available literature reveals that the most appropriate technology from the FNGP would be a reverse osmosis filtration unit followed up with a BCS assembly or a deep well.

### 6.2 Available Options

The most suitable ZED technology for the FNGP has been previously identified as an RO filtration unit paired up with a waste disposal system consisting of either a BCS assembly or a deep
well. There are, however, a number of different ways of applying this technology. The available options include:

- using a 1,2 or multi-stage RO filtration system
andlor
- disposing of RO concentrates in a deep well
andlor
- disposing of RO concentrates with a BCS assembly
andlor
- replacing the ion-exchangers in the front-end softening system with nanofilters (NF) to reduce solid waste production (i.e. no longer adding the large quantities of sodium chloride used to regenerate the ion-exchangers into the water network)

Preliminary work with the computer simulator has also indicated that calcium, carbonate and sulphate concentrations in the FNR discharge flow are so high that, without adequate pretreatment, solids will precipitate onto the RO membranes. Furthermore, regardless of the extent of pretreatment, solids will continue to form in a RO system containing more than 2 successive filtering units. These findings led to the inclusion of a calcium-removal pretreatment step into the computer model, as well as the development of two additional ZED alternatives:

1) placing the RO filters and associated disposal equipment between the hot lime treater and the ion-exchangers, rather than downstream of the polishing pond.
and/or
2) RO technology is completely excluded from the ZED system, and only a BCS unit is used

With configuration (1), effluent from the polishing pond would flow directly into the hot lime treater, which is currently in operation only 1 out of every 2 or 3 days (Shang Su, personal com.).

Here it would mix with incoming raw water, and the combined mixture would pass into the RO assembly. The inherent advantage of this design is the use of the lime treater to partially soften the RO feedwaters, rather than building a new calcium removal process for this task.

As for option (2), the performance of a brine concentrator (B/C) is unaffected by the precipitation of calcium and magnesium salts. Instead of attaching to the sides of the vessel, solids precipitate onto seed particles introduced into the feed stream as it enters the B/C (Pankratz \& Johanson 1992, Bowlin \& Ludlum 1992). As a result, a B/C can operate effectively even with highly concentrated wastewaters, hence the original intention of using a $B / C$ to treat RO concentrates.

Given the abundance of potential options, there are numerous ZED combinations available to the plant. Initial work with the simulator indicated that any RO filtration system would be limited to two stages and that calcium pretreatment would be necessary to prevent membrane fouling even in a 1-stage set-up. With these restrictions in mind, a variety of conceptual ZED-designs were developed.

### 6.3 Conceptual ZED Designs

Each ZED design was developed around the FNGP's water network as it is expected to perform when the 8 psi steam released from the deaerators is recaptured and reused. Eighteen different configurations were built, and they can be classified into one of two groups based on the relative position of the ZED equipment.

### 6.3.1 Back-end models

In these scenarios, the required ZED equipment was attached to the polishing pond. Recovered wastewater re-entered the water network through the front-end softening system. Ten back-end models were tested with the simulator. Four of them used 1-stage RO units, while another 4 simulations incorporated 2 -stage RO filtration. The remaining 2 designs did not contain any RO
membranes; they relied solely on a BCS assembly. All 10 scenarios can be loosely represented by the formula:
$\mathrm{I}-\mathrm{X}$ (ion-exchange) or NF softening +x -stage $\mathrm{RO}+\mathrm{BCS}$ or deep well disposal

### 6.3.2 Composite discharge designs

As opposed to the back-end models, composite discharge from the polishing pond now flowed directly into the hot lime treater, and then into downstream RO filters. Eight composite discharge solutions were developed. Four of them used 1 -stage RO filtration, while the other 4 contained a 2-stage RO process. The following equation symbolizes the different composite discharge options:

I-X or NF softening + direct effluent reuse +x -stage $\mathrm{RO}+\mathrm{BCS}$ or deep well disposal

No composite discharge/BCS-only configurations were built, since the basis for directly reusing polishing pond effluent was to circumvent the need for an external RO pretreatment train. As no RO technology was used in the BCS-only models, this premise became irrelevant. Figure 6.1 summarizes the basic structure of all 18 ZED designs.

Table 6.1: A chemical comparison of the treated water and Fort Nelson River (FNR) discharge flows.

| Parameter | FNR discharge | Treated water |
| :---: | :---: | :---: |
| Total dissolved solids - $\mathrm{mg} / \mathrm{L}$ | 2744 | 279 |
| Total organic carbon - $\mathrm{mg} / \mathrm{L}$ | 97.7 | 20.7 |
| Inorganics: |  |  |
| Chlorides - $\mathrm{mg} / \mathrm{L}$ | 627 | 2.4 |
| Sulfates - $\mathrm{mg} \mathrm{SO} 4 / \mathrm{L}$ | 543 | 81.9 |
| Silica - mg SiO $2 / \mathrm{L}$ | 5.2 | 0.9 |
| Metals: |  |  |
| Calcium $-\mathrm{mg} / \mathrm{L}$ | 146 | 1.4 |
| Magnesium $-\mathrm{mg} / \mathrm{L}$ | 20.6 | 0.5 |
| Sodium $-\mathrm{mg} / \mathrm{L}$ | 584 | 50.5 |

Softening technology


$$
=\text { not evaluated }
$$

$=$ brine concentrator + spray dryer
Figure 6.1: A graphic representation of the available ZED options

### 7.0 DESIGN EVALUATION

The 18 proposed ZED designs were previously described as either back-end or composite discharge models depending on the relative position of the installed ZED equipment. The presence or absence of a brine concentrator - spray dryer (BCS) assembly is, however, a more significant difference between the various conceptual configurations. Those solutions with a deep well in place of a BCS assembly can only transform the FNGP into a ZED facility in the strict sense that there will no longer a surface discharge from the plant to the FNR; the water network will, however, continue to generate a liquid waste, which will be disposed of as a liquid. The water loop will never be completely closed, unlike the BCS models which will generate a solid waste. The only water that will escape from a BCS design will be waste steam vented from the spray dryer. Given this fundamental difference between the 18 ZED solutions, they were evaluated separately as either BCS or deep well designs. Factors used to determine which are the most effective ZED programs included cost, complexity, waste generation rates and calcium pretreatment requirements.

### 7.1 Simulator Performance

The computer simulator was, as previously detailed, originally constructed from the water flow and water chemistry data collected at the plant. The inputted data were then replaced with formulas, transforming the static model into a predictive tool. Preliminary work with the simulator identified various bugs in the system; it occasionally crashed or did not respond as expected. All of the equations and the layout of the model were reviewed and, if necessary corrected. The simulator was then repeatedly run through a series of simulations to examine if the problems had been properly addressed. This exercise was also used to ensure that the computer's projected image of the FNGP as a ZED facility was consistent over time and independent of previous simulations. Once the simulator passed through these tests without error, it was used to evaluate the 18 proposed ZED configurations summarized in Figure 6.1. At no time during the design evaluations were any major problems encountered. The program appeared to be bug free.

## 7,2 BCS Designs

### 7.2.1 Composite discharge models

The hot lime treater successfully removed enough calcium in all four composite discharge scenarios to circumvent the need to build a calcium removal process upstream of the RO filters (Table 7.1). Product water quality from each ZED assembly was such that, when all four systems were in full recycle, downstream softeners were no longer required. The ion-exchangers were subsequently dropped from their respective simulations (Table 7.1). The nanofilters remained in place, since they removed more than just calcium and magnesium.

Using in-situ equipment to pretreat RO feedwaters required significant alterations to the existing water network (e.g. Fig. 7.1). Much larger RO and BCS units were used in all 4 composite discharge solutions compared to the simpler back-end models (Table 7.1). They were, as a result, among the most expensive of the available scenarios (Table 7.1). The composite discharge simulations also produced more waste steam than many of the other designs and roughly equivalent amounts of solid waste (Table 7.1). As there are no significant cost or waste savings inherent in any of the 4 complex composite discharge configurations, compared to the other available options, none of them appear to be particularly well suited for the FNGP.

### 7.2.2 Back-end solutions: Ion-exchange vs. nanofiltration

In all 3 back-end categories (i.e. 1-stage RO, 2-stage RO and BCS-only), the scenarios containing nanofilter (NF) were at times almost identical to their respective ion-exchange (I-X) counterparts. Both models generally possessed the same size RO and/or BCS systems and produced equal volumes of waste steam, regardless of whether the ZED equipment had just been attached or reclaimed wastewater was being recycled (Table 7.1). Calcium pretreatment demands of the 1 -stage $\mathrm{RO} / \mathrm{NF}$ configuration were also equivalent to those of the 1 -stage $\mathrm{RO} / \mathrm{I}-\mathrm{X}$ option (Table 7.1). This was not the case in the 2 -stage RO designs; the "NF" scenario continually required more calcium to be removed from RO feedwaters than its "I-X" alternative (Table 7.1).

There were other trends in the "I-X" and "NF" solutions which were common to all 3 groups. The "NF" options were always more expensive than the opposing "I-X" designs, due to the capital costs associated with replacing the existing ion-exchangers with nanofilters (Table 7.1). Because the front-end treatment network had to be altered to install the NF vessels (Fig. 7.2), the "NF" configurations would in reality be harder to assemble and slightly more complex to operate than their respective "I-X" alternatives. The "NF" simulations did, on the other hand, generate less solid waste than the "I-X" systems (Table 7.1).

Before reclaimed wastewater was recirculated back through the plant, the difference in solid waste generation rates were significant. The "NF" models initially produced less than $2 / 3$ of the total waste output of their respective "I-X" counterparts (Table 7.1). These savings dropped to around $13 \%$ in both of the RO simulations after the water loops were closed (Table 7.1). The difference between the BCS-only scenarios remained above $28 \%$ (Table 7.1).

With respect to the 4 RO designs, neither "NF" configuration illustrated.significant, sustained gains which would have suggested that they were more suitable to the FNGP than their respective "I-X" alternatives. Both "NF" systems were more expensive, more complex and, at least in the case of the 2-stage RO/NF simulations, more delicate (as illustrated by higher pretreatment demands) than either RO/I-X option (Table 7.1). Their respective BCS units produced near identical amounts of waste steam (Table 7.1). Initial solid waste savings quickly fell to only a 13 \% margin of difference once wastewater recycling had started (Table 7.1). Neither RO/NF model is the best ZED design available to the FNGP.

All of the finding summarized in the last paragraph also apply to the BCS/NF scenario, except for solid waste output. This configuration, as stated above, continually generated less solid waste than the BCS/I-X option. Until disposal costs are properly assessed, the value of reduced wasting rates remains unclear. As such, all 3 "I-X" solutions were still deemed to be more efficient that their "NF" counterparts.

### 7.2.3 Best of the back-end ion-exchangers

Cost and complexity: The BCS-only model was the simplest of the available configurations (Table 7.1). It contained the shortest ZED treatment system; wastewater reclamation occurred in a single vessel (Fig. 7.3), rather than in a series of separate units (Fig. 7.4), and pretreatment will likely be limited to deaeration and pH adjustments (Fig. 7.3), instead of multiple pre-filters, pH adjustments, anti-scalant addition and possibly chlorination and/or dechlorination (Fig. 7.4). The brine concentrator ( $\mathrm{B} / \mathrm{C}$ ) was, however, the most expensive component of any BCS design (Appendix G). The BCS-only scenario contained the largest $\mathrm{B} / \mathrm{C}$ of the three "I-X" options (Table 7.1). As a result, the simplest ZED configuration was also the most expensive (Table 7.1).

Using a 2-stage RO unit in place of a "1-stage" process increased the complexity of the overall water network. Additional pumps, pipes and pH adjustment stations were required to support the second set of membrane filters (Fig. 7.5 vs . Fig. 7.6). The resulting wastewater flow path was longer and more complicated than in either the 1 -stage RO or BCS-only configurations. On the other hand, adding the second stage cut the flow of RO concentrates to the $\mathrm{B} / \mathrm{C}$ by $10 \%$ (Table 7.1). The monetary savings of using a smaller $B / C$ were such that the 2 -stage $R O$ design was slightly cheaper than its "1-stage" counterpart (Table 7.1).

It appears that the more complex the option, the cheaper its overall capital cost. Both RO designs were cheaper than the simplest ZED scenario, the BCS-only/I-X model. The more complex 2 -stage RO simulation was similarly cheaper than the simpler RO alternative.

Waste generation rates: Steam losses from the ZED equipment were proportional to the size of the B/C. As such, the BCS-only model generated the most waste steam, while the 2 -stage RO simulations produced the least (Table 7.1).

Solid waste production followed a similar trend. The BCS-only scenario had the highest generation rates of the back-end/I-X solutions (Table 7.1). The 2 -stage RO design produced less solid waste than its "1-stage" counterpart, although the difference between the two configurations
was less than $5 \%$ (Table 7.1). While the BCS-only scenario continually generated the most solid and steam waste, and the two RO models had near identical solid waste production rates, the 2stage RO design always vented less steam than the 1 -stage RO alternative.

Robustness and treated water quality: Although the robustness of a $\mathrm{B} / \mathrm{C}$ has already been discussed as it pertains to pretreatment requirements, it is the cleaning ability of each ZED system that is now being evaluated. Robustness, within this context, was assessed by observing fluctuation in product and treated water quality that occurred when given contaminant concentrations in either the raw water inflow or effluent plant outflow were changed; product water, or reclaimed wastewater, refers to water produced by the ZED equipment, while treated water is softened water leaving the FESs. Prior to any alterations, the BCS model contained the cleanest product and treated waters, followed, in order, by the 1 -stage and 2 -stage RO scenarios (Tables $7.2 \& 7.3$ ).

When different contaminant concentrations were doubled, reclaimed wastewater and treated water quality in the BCS-only configuration generally changed the least (Tables $7.2 \& 7.3$ ). The cleaning ability of this process was perhaps best demonstrated when chloride and dissolved organic carbon (DOC) levels were increased. Although these changes resulted in significant increases in the abundance of both chloride and DOC in both RO designs, there was no observed deterioration in treated water quality in the BCS-only scenario (Table 7.3). Not only did the BCSonly system initially produce the cleanest water, it also proved to be the most robust of the three back-end/I-X options.

The fragility of the 2 -stage RO model was illustrated by its calcium pretreatment requirements. Whenever calcium, sulfate or alkalinity levels were raised, the risk of membrane fouling increased, so more calcium had to be removed in the pretreatment train (Table 7.2). The 1-stage RO design did not show the same susceptibility to salt precipitation. When the water loop was first closed, calcium pretreatment was no longer necessary (Table 7.1). An observation not mirrored in the 2stage alternative (Table 7.1). Subsequent changes to raw water or effluent plant chemistry rarely
resulted in the reinstatement of a calcium removal step (Table 7.2). The probability of salt precipitating in the 2 -stage RO membranes was much higher than in the 1 -stage RO vessels, illustrating the relative fragility of this system compared to its more resilient "1-stage" counterpart.

Aside from being more delicate than the 1 -stage RO system, the 2 -stage RO configuration tended to produce the lowest quality recycled and treated waters of the three ZED options (Tables $7.2 \& 7.3$ ). Chloride and DOC levels in the treated water were, even before any chemical alterations were made, well above current concentrations (Table 7.3). Furthermore, although some chemical manipulations caused larger relative shifts in the "1-stage RO" waters, the 1 -stage RO solution always contained product and treated waters of equal or better quality than its 2-stage RO alternative (Tables $7.2 \& 7.3$ ). While the BCS-only design continually produced the highest quality waters regardless of changing water chemistry, the 1 -stage RO configuration was more resilient, and generated cleaner waters, than its " 2 -stage" counterpart.

Best design: The 1 -stage RO design was less complex and more robust than the 2 -stage RO. scenario (Tables 7.1, $7.2 \& 7.3$ ). Treated and recycled water quality were similarly better in the $1-$ stage RO simulation (Tables $7.2 \& 7.3$ ). The two RO systems generated roughly equivalent amounts of solid waste, although steam losses were higher in the 1 -stage RO solution (Table 7.1). The 2-stage RO option was also found to be $\$ 40000$ cheaper that the 1 -stage RO alternative, since it used a smaller BCS unit (Table 7.1). This was an odd finding as one would have expected a two stage RO unit to cost substantially more than a single stage RO filter. In light of the possibly questionable monetary savings associated with the more delicate and complex 2-stage RO simulation, the 1 -stage RO model seems to be the better RO design.

Choosing the "best" ZED design for the FNGP thus comes down to the relative importance of cost, as both the 1 -stage RO and BCS-only configurations performed successfully. The BCS-only scenario had the higher capital costs (Table 7.1). It was, on the other hand, simpler and more robust than the 1 -stage RO simulation (Tables $7.1,7.2 \& 7.3$ ). The BCS-only solution also produced better quality recycled water than the 1 -stage RO alternative (Table 7.2), which lead to
the observed larger solid waste generation rates (i.e. all contaminants stripped from reclaimed wastewater were released as a solid, so the cleaner the product water, the larger the mass of removed solids) (Table 7.1). It was impossible to predict how solid waste output and recycled water quality might affect the operating costs of these two ZED options. Therefore, within the limits of this study, it was impossible to determine which of the two configurations is the best ZED system for the FNGP. Only further research, encompassing both bench-scale and pilot plant experiments, will reveal which is the better design.

### 7.2.4 Conclusion

Ten BCS ZED scenarios were originally developed and evaluated. The 4 composite discharge configurations proved to be extremely bulky and complex. They were more expensive and less effective that the other 6 back-end models. None of these options were deemed to be appropriate for the FNGP.

In 3 of the back-end designs, the existing ion-exchangers were replaced with nanofilters. Due to the installation costs of these new softeners, all of the "NF" scenarios were more.expensive than their respective "I-X" counterparts. There were few, if any, benefits incurred from changing softeners that justified their relatively high costs. They were subsequently eliminated from the selection process.

Comparisons between the 3 remaining ion-exchange solutions revealed that the 2 -stage RO configurations was complex, fragile and produced relatively poor quality recycled water. The 1stage RO scenario, while a better choice than its " 2 -stage" counterpart, was less robust and more complicated than the BCS-only simulation. On the other hand, it was cheaper than the BCS-only system, and, with adequate pretreatment, would be just as successful at closing off the plant as the BCS-only design. Further research is needed to determine which of the two configurations, the 1 stage RO or the BCS-only option, is the best ZED system for the FNGP.

### 7.3 Deep Well Configurations

### 7.3.1 Composite discharge designs

Similar to the trends observed in the BCS designs, the deep well composite discharge configurations used much larger ZED equipment than any of the back-end scenarios (Table 7.4). They were more complex than the back-end systems, and they produced far more wastewater than their back-end counterparts (Table 7.4). The sole advantage of directly reusing the polishing pond outflow waters was the elimination of an external calcium removal process (Table 7.4). Using insitu equipment for RO pretreatment did not, however, translate into substantial cost reductions. Due to the size of the RO units required by these designs, the composite discharge solutions were far more expensive than either back-end/I-X option (Table 7.4). Even though the two composite discharge/I-X scenarios were cheaper than the back-end/NF alternatives (Table 7.4), all 4 composite discharge models were dismissed as possible ZED designs for the FNGP.

### 7.3.2 Back-end solutions: Ion-exchange vs. nanofiltration

RO feedwater and concentrate flows differed by at most $7 \%$ between the back-end/NF configurations and their respective "I-X" counterparts (Table 7.4). The "NF" models were more complex and expensive than either "I-X" solution (Table 7.4). As there were no immediate benefits to replacing the existing ion-exchangers with nanofilters, neither back-end/NF design appeared to be the best ZED solution available to the FNGP.

### 7.3.3 Best of the back-end ion-exchangers

Cost and complexity: The 1 -stage RO design was cheaper than its " 2 -stage" counterpart (Table 7.4). As previously indicated, expanding a RO unit to include a second set of filters increased the complexity of the resulting water network. The 2 -stage RO solution was, therefore, the more complicated and expensive of the two options (Table 7.4).

It is important to note that this price comparison was, as were all deep well cost analyses, limited to the capital costs of the RO membranes and associated supportive equipment (Appendix G). The cost of a deep well was assumed to be independent of the ZED configuration.

Waste generation rates: The 2 -stage RO simulation operated at a wastewater recovery rate of $60 \%$, while the 1 -stage RO solution was limited to $50 \%$ (Appendix F). Not surprisingly, the 1stage RO system discharged more liquid waste to the deep well than the 2 -stage RO model (Table 7.4).

Robustness: A ZED system's robustness was tested, as explained earlier, by doubling the concentrations of different constituents in the raw and effluent plant outflow waters, and observing the resulting changes in recycled and treated water quality. Prior to any alterations, the 1 -stage RO simulation contained the better quality product and treated waters (Tables $7.5 \& 7.6$ ). It continued to contain the cleaner waters when contaminant levels were changed (Tables $7.5 \& 7.6$ ). The 1 stage RO option was the more robust of the two deep well solutions; it produced better quality recycled and treated waters than its " 2 -stage" counterpart regardless of the changes in water chemistry elsewhere in the system.

Best design: The 1 -stage RO design was the better of the two ion-exchange scenarios. It was not only cheaper and simpler than the 2-stage RO option (Table 7.4), it was also produced higher quality treated and recycled water than the " 2 -stage" alternative (Tables $7.5 \& 7.6$ ). The only disadvantage of the 1 -stage RO system was its higher wastewater production rates (Table 7.4), yet, given its many inherent advantages over the other deep well ZED simulations, the 1 -stage RO configuration is felt to be the best deep well solution for the FNGP.

### 7.3.4 Conclusion

Eight deep well models were developed and evaluated with the computer simulator. The 4 composite discharge configurations were too bulky, complex and expensive compared to the other available options. The back-end simulations with nanofilters in place of existing ion-exchangers
were similarly eliminated as possible ZED solutions, as they were more complex, costly and had higher pretreatment demands than the " $\mathrm{I}-\mathrm{X}$ " alternatives.

Of the two remaining scenarios, the 1 -stage back-end RO system was cheaper, more robust and simpler than its "2-stage" counterpart. The 1 -stage RO simulation also produced the higher quality recycled and treated waters of the two models. The only advantage of the 2 -stage RO configurations was its smaller wastewater stream. In any case, the 1 -stage back-end RO design appears to be the best of the deep well options available to FNGP.

### 7.4 Evaluation Summary

Information from the available literature, combined with suggestions from Westcoast personnel and the data collected from the plant, indicated that the FNGP could achieve ZED by a number of different treatment configurations. A computer simulator was built to evaluate the effectiveness of each option. The 18 original scenarios were sub-divided into 2 categories based on their final disposal technology. Each ZED program was then tested, and the resulting output compared to that from the other designs within the group.

The best BCS systems were the back-end, 1-stage RO and BCS-only models, which maintained an ion-exchanger in the plant's front-end softening system. They were both more robust, cost effective and simpler than the alternatives. The 1-stage RO scenario would cost around $\$ 1.61$ million US dollars, while the BCS-only configuration has a price tag of around $\$ 1.79$ million US dollars.

As for the deep well designs, the best option was the back-end, 1-stage RO unit with an ionexchanger in the raw water softening system. This model was more efficient and less complex than the other deep well configurations. The 1 -stage RO system would cost approximately \$101 700 US dollars, excluding the cost of the deep well.

It is important to note that within every test run, it was assumed that the 8 psi steam now venting off of the deaerators was recaptured and reused elsewhere in the water system. The cost
analyses reported herein were also very rudimentary in nature. They were developed from RO, NF, BCS, multi-media filter and ion-exchange unit capital costs (Appendix G); operating costs were not assessed. Given the preliminary nature of this study, the reported cost estimates should be sufficient for Westcoast personnel to decide if a ZED program for the FNGP is economically viable and which of the evaluated design to pursue.
Table 7.1: Comparison of the 10 selected BCS ZED designs. Values are shown for both initial conditions and full wastewater recycling, with the latter in parentheses.

| Evaluation criteria | Back-end models |  |  |  |  |  | Composite discharge |  |  |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  | $\underset{\mathrm{I}-\mathrm{X}}{\text { 1-stage RO }} \underset{\mathrm{NF}}{\text { R }}$ |  | $\underset{\text { I-X }}{\text { 2-stage RO }}$ |  | $\underset{\mathrm{I}-\mathrm{X}}{\mathrm{BCS}-o n l y} \underset{\mathrm{NF}}{ }$ |  | $\underset{I-X^{\mathrm{a}}}{\text { 1-stage RO }} \underset{\mathrm{NF}}{\mathrm{~N}}$ |  | 2-stage RO |  |
| Feedwater flows ( $\mathrm{m}^{3} /$ day ) |  |  |  |  |  |  |  |  |  |  |
| RO | $\begin{gathered} 282 \\ (240) \end{gathered}$ | $\begin{gathered} 279 \\ (225) \end{gathered}$ | $\begin{gathered} 282 \\ (244) \end{gathered}$ | $\begin{gathered} 279 \\ (225) \end{gathered}$ | - | - | $\begin{gathered} 539 \\ (538) \end{gathered}$ | $\begin{gathered} 517 \\ (514) \end{gathered}$ | $\begin{gathered} 546 \\ (529) \end{gathered}$ | $\begin{gathered} 520 \\ (509) \end{gathered}$ |
| B/C | $\begin{gathered} 141 \\ (120) \end{gathered}$ | $\begin{gathered} 140 \\ (112) \end{gathered}$ | $\begin{gathered} 113 \\ (97.6) \end{gathered}$ | $\begin{gathered} 112 \\ (90.0) \end{gathered}$ | $\begin{gathered} 282 \\ (235) \end{gathered}$ | $\begin{gathered} 279 \\ (238) \end{gathered}$ | $\begin{gathered} 269 \\ (269) \end{gathered}$ | $\begin{gathered} 259 \\ (257) \end{gathered}$ | $\begin{gathered} 219 \\ (132) \end{gathered}$ | $\begin{gathered} 208 \\ (127) \end{gathered}$ |
| Waste generation rates |  |  |  |  |  |  |  |  |  |  |
| Steam ( $\mathrm{m}^{3} /$ day) | $\begin{gathered} 14.1 \\ (12.0) \end{gathered}$ | $\begin{gathered} 14.0 \\ (11.2) \end{gathered}$ | $\begin{aligned} & 11.3 \\ & (9.8) \end{aligned}$ | $\begin{aligned} & 11.2 \\ & (9.0) \end{aligned}$ | $\begin{gathered} 28.2 \\ (23.5) \end{gathered}$ | $\begin{gathered} 27.9 \\ (23.8) \end{gathered}$ | $\begin{gathered} 26.9 \\ (26.9) \end{gathered}$ | $\begin{gathered} 25.7 \\ (25.7) \end{gathered}$ | $\begin{gathered} 21.9 \\ (13.2) \end{gathered}$ | $\begin{gathered} 20.8 \\ (12.7) \end{gathered}$ |
| Soilds (kg/day) | $\begin{gathered} 415 \\ (233) \end{gathered}$ | $\begin{gathered} 256 \\ (203) \end{gathered}$ | $\begin{gathered} 398 \\ (230) \end{gathered}$ | $\begin{gathered} 246 \\ (200) \end{gathered}$ | $\begin{gathered} 453 \\ (313) \end{gathered}$ | $\begin{gathered} 287 \\ (225) \end{gathered}$ | $\begin{gathered} 411 \\ (200) \end{gathered}$ | $\begin{gathered} 407 \\ (200) \end{gathered}$ | $\begin{gathered} 397 \\ (195) \end{gathered}$ | $\begin{gathered} 393 \\ (196) \end{gathered}$ |
| Calcium pretreatment ( $\mathrm{mg} / \mathrm{L}$ ) |  |  |  |  |  |  |  |  |  |  |
|  | $\begin{aligned} & 53 \\ & (0) \end{aligned}$ | $\begin{aligned} & 50 \\ & (0) \end{aligned}$ | $\begin{aligned} & 68 \\ & (6) \end{aligned}$ | $\begin{gathered} 75 \\ (17) \end{gathered}$ | - | - | $\begin{gathered} 0 \\ (0) \end{gathered}$ | $\begin{gathered} 0 \\ (0) \end{gathered}$ | $\begin{gathered} 0 \\ (0) \end{gathered}$ | $\begin{gathered} 0 \\ (0) \end{gathered}$ |
| Cost (million US dollars) ${ }^{\text {b }}$ |  |  |  |  |  |  |  |  |  |  |
|  | 1.61 | 1.73 | 1.57 | 1.69 | 1.79 | 1.91 | 1.85 | 1.95 | 1.77 | 1.87 |

Table 7.1 (con't)

a When reclaimed wastewater was recycing, ion-exchangers dropped from simulation
b see Appendix G for calculations
Table 7.2: Changes in RO and $\mathrm{B} / \mathrm{C}$ feedwater flows, calcium pretreatment demands and product water quality, in the 1 -stage RO, 2 stage RO and BCS-only back-end, BCS, ion-exchange configurations, triggered by given alterations in raw water and effluent plant outflow chemistry.

| Chemical change | Model | Calcium pretreat (mg/L) | $\underset{\substack{\text { ROed flow } \\\left(\mathrm{m}^{3} / \mathrm{day}\right)}}{\substack{\text { fay }}}$ | $\begin{gathered} \text { B/C } \\ \text { feed flow } \\ \left(\mathrm{m}^{3} / \mathrm{day}\right) \end{gathered}$ | Calcium ( $\mathrm{mg} / \mathrm{L}$ ) | Recove Magnesium (mg/L) | ed water ch Sulphate ( $\mathrm{mg} / \mathrm{L}$ ) | mistry Chloride ( $\mathrm{mg} / \mathrm{L}$ ) | $\begin{gathered} \text { DOC } \\ (\mathrm{mg} / \mathrm{L}) \end{gathered}$ |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| No changes |  |  |  |  |  |  |  |  |  |
|  | 1 -stage RO | 0.0 | 240 | 120 | 1.3 | 0.2 | 5.8 | 7.9 | 9.7 |
|  | 2-stage RO | 5.6 | 244 | 97.6 | 2.3 | 0.5 | 11.4 | 15.7 | 19.0 |
|  | BCS-only |  |  | 232 | $7.2 \times 10^{-3}$ | $1.3 \times 10^{-3}$ | $3.1 \times 10^{-2}$ | $3.7 \times 10^{-2}$ | $9.9 \times 10^{-3}$ |
| Raw water |  |  |  |  |  |  |  |  |  |
| - doubled Ca \& $\mathrm{Mg}^{\text {a }}$ |  |  |  |  |  |  |  |  |  |
|  | 1 -stage RO | 0.0 | $\begin{aligned} & 266 \\ & 11 \% \end{aligned}$ | $\begin{gathered} 132 \\ 11 \% \end{gathered}$ | $\begin{gathered} 1.9 \\ 47 \% \end{gathered}$ | $\begin{gathered} 41 \% \end{gathered}$ | $\begin{aligned} & 5.3 \\ & -9 \% \end{aligned}$ | $\begin{aligned} & 20.3 \\ & 157 \% \end{aligned}$ | $\begin{aligned} & 8.8 \\ & -9 \% \end{aligned}$ |
|  | 2-stage RO | $\begin{aligned} & 27.6 \\ & 392 \% \end{aligned}$ | $\begin{aligned} & 271 \\ & 11 \% \end{aligned}$ | $\begin{aligned} & 108 \\ & 11 \% \end{aligned}$ | $\begin{aligned} & 2.5 \\ & .8 \% \end{aligned}$ | $\begin{aligned} & 0.5 \\ & 8 \% \end{aligned}$ | ${ }_{-9 \%}^{10.3}$ | $\begin{aligned} & 38.3 \\ & 144 \% \end{aligned}$ | $\begin{gathered} 17.2 \\ -10 \% \end{gathered}$ |
|  | BCS only | - | - | $\begin{aligned} & 255 \\ & 10 \% \end{aligned}$ | $\begin{gathered} 1.1 \times 10^{-2} \\ 51 \% \end{gathered}$ | $\begin{gathered} 1.6 \times 10^{-3} \\ 23 \% \end{gathered}$ | $\begin{gathered} 2.8 \times 10^{-2} \\ -8 \% \end{gathered}$ | $\begin{gathered} 7.1 \times 10^{-2} \\ 92 \% \end{gathered}$ | $9.1 \times 10^{-3}$ |
| - doubled $\mathrm{SO}_{4}$ \& T. Alk. ${ }^{\text {a }}$ |  |  |  |  |  |  |  |  |  |
|  | 1 -stage RO | 0.0 | $\begin{gathered} 24 \% \\ 3 \% \end{gathered}$ | $\begin{aligned} & 123 \\ & 3 \% \end{aligned}$ | $\begin{aligned} & 1.3 \\ & -2 \% \end{aligned}$ | $\begin{aligned} & 0.2 \\ & 16 \% \end{aligned}$ | $\begin{aligned} & 7.2 \\ & 24 \% \end{aligned}$ | $\begin{aligned} & 7.8 \\ & -1 \% \end{aligned}$ | $\begin{aligned} & 9.5 \\ & -2 \% \end{aligned}$ |
|  | 2-stage RO | $\begin{aligned} & 13.8 \\ & 147 \% \end{aligned}$ | $\begin{aligned} & 250 \\ & 3 \% \end{aligned}$ | $\begin{aligned} & 100 \\ & 3 \% \end{aligned}$ | $\begin{gathered} 1.9 \\ -19 \% \end{gathered}$ | $\begin{gathered} 0.4 \\ -11 \% \end{gathered}$ | $\begin{aligned} & 14.1 \\ & 24 \% \end{aligned}$ | $\begin{aligned} & 15.4 \\ & -2 \% \end{aligned}$ | $\begin{aligned} & 18.6 \\ & -2 \% \end{aligned}$ |
|  | BCS only | - | - | $\begin{gathered} 232 \\ 0 \% \end{gathered}$ | $\underset{2 \%}{7.3 \times 10^{-3}}$ | $\underset{1 \%}{1.3 \times 10^{-3}}$ | $\begin{gathered} 4.0 \times 10^{-2} \\ 31 \% \end{gathered}$ | $\begin{gathered} 2.1 \times 10^{-2} \\ 44 \% \end{gathered}$ | $\begin{gathered} 1.0 \times 10^{-2} \\ 1 \% \end{gathered}$ |

Table 7.2 (con't)

| Chemical <br> change | Model | Calcium <br> pretreat <br> $(\mathrm{mg} / \mathrm{L})$ | RO <br> feed flow <br> $\left(\mathrm{m}^{3} / \mathrm{day}\right)$ | B/C <br> feed flow <br> $(\mathrm{m} 3 / \mathrm{day})$ | Calcium <br> $(\mathrm{mg} / \mathrm{L})$ | Recovered water chemistry <br> Magnesium <br> $(\mathrm{mg} / \mathrm{L})$ | Sulphate <br> $(\mathrm{mg} / \mathrm{L})$ | Chloride <br> $(\mathrm{mg} / \mathrm{L})$ |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| E/P outflow |  |  |  |  |  |  |  |  |
| - doubled Ca \& Mga |  |  |  |  |  |  |  |  |
| (mg/L) |  |  |  |  |  |  |  |  |

Table 7.2 (con't)

| Chemical change | Model | Calcium pretreat ( $\mathrm{mg} / \mathrm{L}$ ) | $\begin{gathered} \text { RO } \\ \text { feed flow } \\ \left(\mathrm{m}^{3} / \text { day }\right) \end{gathered}$ | $\begin{gathered} B / C \\ \text { feed flow } \\ \left(\mathrm{m}^{3} / \text { day }\right) \end{gathered}$ | Calcium (mg/L) | Recove Magnesium (mg/L) | d water ch Sulphate ( $\mathrm{mg} / \mathrm{L}$ ) | mistry Chloride (mg/L) | $\begin{gathered} \mathrm{DOC} \\ (\mathrm{mg} / \mathrm{L}) \end{gathered}$ |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| E/P outflow |  |  |  |  |  |  |  |  |  |
| - doubled Cla |  |  |  |  |  |  |  |  |  |
|  | 1-stage RO | 0.0 | $\begin{gathered} 240 \\ 0 \% \end{gathered}$ | $\begin{gathered} 120 \\ 0 \% \end{gathered}$ | $1.3$ | $\begin{aligned} & 0.2 \\ & 19 \% \end{aligned}$ | $\begin{aligned} & 5.8 \\ & 0 \% \end{aligned}$ | $12.0$ | $\begin{gathered} 9.8 \\ 1 \% \end{gathered}$ |
|  | 2-stage RO | $\begin{aligned} & 5.4 \\ & -4 \% \end{aligned}$ | $\begin{gathered} 245 \\ 0 \% \end{gathered}$ | $\begin{gathered} 98.0 \\ 0 \% \end{gathered}$ | $\begin{gathered} 2.3 \\ -1 \% \end{gathered}$ | $\begin{aligned} & 0.5 \\ & -9 \% \end{aligned}$ | $\begin{gathered} 11.4 \\ 0 \% \end{gathered}$ | $\begin{gathered} 23.5 \\ 50 \% \end{gathered}$ | $\begin{gathered} 19.0 \\ 0 \% \end{gathered}$ |
|  | BCS only | - | - | $\begin{gathered} 232 \\ 0 \% \end{gathered}$ | $\begin{gathered} 7.3 \times 10^{-3} \\ 2 \% \end{gathered}$ | $\begin{gathered} 1.3 \times 10^{-3} \\ 2 \% \end{gathered}$ | $\begin{gathered} 3.1 \times 10^{-2} \\ 2 \% \end{gathered}$ | $\begin{gathered} 4.8 \times 10^{-2} \\ 30 \% \end{gathered}$ | $\begin{gathered} 1.0 \times 10^{-2} \\ 1 \% \end{gathered}$ |

${ }^{\mathrm{a}} \mathrm{Ca}=$ calcium, $\mathrm{Mg}=$ magnesium, $\mathrm{SO}_{4}=$ sulphate, $\mathrm{T} . \mathrm{Alk} .=$ total alkalinity, $\mathrm{Cl}=$ cloride, $\mathrm{DOC}=$ dissolved organic carbon

Table 7.3: Changes in treated water quality, in the 1 -stage RO, 2 -stage RO and BCS-only backend, BCS, ion-exchange configurations, triggered by given alterations in raw water and effluent plant outflow chemistry.

| Chemical <br> change | Model | Sulphate <br> $(\mathrm{mg} / \mathrm{L})$ | Treated water chemistrya <br> Sodium <br> $(\mathrm{mg} / \mathrm{L})$ | Chloride <br> $(\mathrm{mg} / \mathrm{L})$ | DOC <br> $(\mathrm{mg} / \mathrm{L})$ |
| :---: | :---: | :---: | :---: | :---: | :---: | | Total CO3 |
| :---: |
| $(\mathrm{mol} / \mathrm{L})$ |

## Raw water

- doubled $\mathrm{Ca} \& \mathrm{Mg}^{\text {b }}$

| 1-stage RO | 24.3 | 52.6 | 9.9 | 5.8 | $4.7 \times 10^{-4}$ |
| :--- | :---: | :---: | :---: | :---: | :---: |
|  | $-4 \%$ | $217 \%$ | $135 \%$ | $-4 \%$ | $30 \%$ |
| 2-stage RO | 26.1 | 59.6 | 18.3 | 8.7 | $4.7 \times 10^{-4}$ |
|  | $-4 \%$ | $162 \%$ | $140 \%$ | $-4 \%$ | $13 \%$ |
| BCS-only | 24.3 | 60.6 | 0.9 | 2.6 | $4.7 \times 10^{-4}$ |
|  | $-3 \%$ | $134 \%$ | $-5 \%$ | $-4 \%$ | $0 \%$ |

- doubled $\mathrm{SO}_{4} \& ~ T . ~ A l k .{ }^{\text {b }}$

| 1-stage RO | 47.8 | 16.5 | 4.1 | 6.0 | $1.3 \times 10^{-3}$ |
| :---: | :---: | :---: | :---: | :---: | :---: |
|  | $89 \%$ | $-1 \%$ | $-2 \%$ | $-1 \%$ | $256 \%$ |
| 2-stage RO | 49.8 | 22.7 | 7.5 | 9.0 | $1.3 \times 10^{-3}$ |
|  | $83 \%$ | $0 \%$ | $-1 \%$ | $-1 \%$ | $220 \%$ |
| BCS-only | 49.5 | 10.7 | 0.9 | 2.7 | $7.5 \times 10^{-4}$ |
|  | $98 \%$ | $-59 \%$ | $-5 \%$ | $0 \%$ | $57 \%$ |

E/P outflow

- doubled $\mathrm{Ca} \& \mathrm{Mg}^{\text {b }}$

| 1-stage RO | 25.3 | 16.6 | 4.1 | 6.0 | $3.45 \times 10^{-4}$ |
| :--- | :---: | :---: | :---: | :---: | :---: |
|  | $0 \%$ | $0 \%$ | $-1 \%$ | $0 \%$ | $-4 \%$ |
| 2-stage RO | 27.2 | 22.8 | 7.6 | 9.1 | $4.1 \times 10^{-4}$ |
|  | $0 \%$ | $0 \%$ | $-1 \%$ | $0 \%$ | $-2 \%$ |
| BCS-only | 25.0 | 25.9 | 0.9 | 2.7 | $4.7 \times 10^{-4}$ |
|  | $0 \%$ | $0 \%$ | $-4 \%$ | $0 \%$ | $0 \%$ |

Table 7.3 (con't)
$\left.\begin{array}{cccccc}\hline \begin{array}{c}\text { Chemical } \\ \text { change }\end{array} & \text { Model } & \begin{array}{c}\text { Sulphate } \\ (\mathrm{mg} / \mathrm{L})\end{array} & \begin{array}{c}\text { Sodium } \\ (\mathrm{mg} / \mathrm{L})\end{array} & \begin{array}{c}\text { Treated water chemistrya } \\ \text { Chloride } \\ (\mathrm{mg} / \mathrm{L})\end{array} & \begin{array}{c}\text { DOC } \\ (\mathrm{mg} / \mathrm{L})\end{array}\end{array} \begin{array}{c}\text { Total CO} \\ (\mathrm{mol} / \mathrm{L})\end{array}\right]$
${ }^{\text {a }}$ Calcium and magnesium treated water concentrations not shown since they were held at $<1.4$ and $0.5 \mathrm{mg} / \mathrm{L}$, repectively.
${ }^{\mathrm{b}} \mathrm{Ca}=$ calcium, $\mathrm{Mg}=$ magnesium, $\mathrm{SO}_{4}=$ sulphate, $\mathrm{T} . \mathrm{Alk} .=$ total alkalinity, $\mathrm{Cl}=$ cloride, Total $\mathrm{CO}_{3}=$ total carbonate content, $\mathrm{DOC}=$ dissolved organic carbon
Table 7.4: Comparison of the 8 selected deep well ZED designs. Values are shown for both initial conditions and full wastewater
recycling, with the latter in parentheses.

| Evaluation criteria | Back-end models |  |  |  | Composite discharge |  |  |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  | 1-stage RO |  | 2-stage RO |  | 1-stage R O |  | 2-stage RO | $\mathbf{O}_{\mathrm{NF}}$ |
| RO feedwater flows ( $\mathrm{m}^{3} /$ day ) |  |  |  |  |  |  |  |  |
|  | $\begin{aligned} & 282 \\ & (263) \end{aligned}$ | $\begin{gathered} 279 \\ (246) \end{gathered}$ | $\begin{gathered} 282 \\ (263) \end{gathered}$ | $\begin{gathered} 279 \\ (246) \end{gathered}$ | $\begin{gathered} 1.04 \times 10^{3} \\ \left(1.03 \times 10^{3}\right) \end{gathered}$ | $\begin{gathered} 983 \\ (979) \end{gathered}$ | $\begin{gathered} 780 \\ (691) \end{gathered}$ | $\begin{gathered} 741 \\ (662) \end{gathered}$ |
| Wastewater generation rates ( $\mathrm{m}^{3} / \mathrm{day}$ ) |  |  |  |  |  |  |  |  |
|  | $\begin{gathered} 141 \\ (131) \end{gathered}$ | $\begin{gathered} 140 \\ (123) \end{gathered}$ | $\begin{gathered} 113 \\ (105) \end{gathered}$ | $\begin{gathered} 112 \\ (98.4) \end{gathered}$ | $\begin{gathered} 518 \\ (514) \end{gathered}$ | $\begin{gathered} 491 \\ (490) \end{gathered}$ | $\begin{gathered} 253 \\ (173) \end{gathered}$ | $\begin{gathered} 241 \\ (166) \end{gathered}$ |
| Calcium pretreatment ( $\mathrm{mg} / \mathrm{L}$ ) |  |  |  |  |  |  |  |  |
|  | $\begin{gathered} 53 \\ (11) \end{gathered}$ | $\begin{gathered} 50 \\ (21) \end{gathered}$ | $\begin{gathered} 68 \\ (29) \end{gathered}$ | $\begin{gathered} 75 \\ (40) \end{gathered}$ | $\begin{gathered} 0 \\ (0) \end{gathered}$ | $\begin{gathered} 0 \\ (0) \end{gathered}$ | $\begin{gathered} 0 \\ (0) \end{gathered}$ | $\begin{gathered} 0 \\ (0) \end{gathered}$ |
|  |  |  |  |  |  |  |  |  |
|  | 1.01 | 2.27 | 1.12 | 2.37 | 1.61 | 2.81 | 1.64 | 2.86 |

Table 7.4 (con't)

| Evaluation criteria | Back-end models |  |  |  | Composite discharge |  |  |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  | $\mathrm{I}-\mathrm{X}$ | $\mathbf{R O}_{\mathbf{N F}}$ | $\mathrm{I}-\mathrm{X}$ | $\mathbf{R O}_{\mathbf{N F}}$ | $\mathrm{I}-\mathrm{X}^{\text {-sta }}$ | $\mathrm{RO}_{\mathrm{N}}^{\mathrm{F}}$ | $I-X^{2-s t a g}$ | $\mathrm{RO}_{\mathrm{NF}}$ |
| Complexity |  |  |  |  |  |  |  |  |
|  | Simple | Relatively simple | Relatively simple | Moderate | Moderately complex | Complex | Moderately complex | Complex |
|  | add | add | add | add | major | major | major | major |
|  | 1-stage RO | NF + | 2-stage RO | NF + | changes to | changes to | changes to | changes to |
|  | + | 1-stage RO | + | 2-stage RO | flow paths | flow paths | flow paths | flow paths |
|  | deep well | \& | deep well | \& | + | + | + | + |
|  |  | deep well |  | deep well | 1 -stage RO <br> \& deep well | 1-RO, NF \& deep well | 2-stage RO <br> \& deep well | $\begin{gathered} \text { 2-RO, NF \& } \\ \text { deep well } \end{gathered}$ |

[^5]Table 7.5: Changes in RO feedwater flows, calcium pretreatment demands and product water quality, in both the 1 -stage and 2 -stage


| Chemical change | Model | Calcium pretreat (mg/L) | $\begin{gathered} \text { RO } \\ \text { feed flow } \\ \left(\mathrm{m}^{3} / \text { day }\right) \end{gathered}$ | Calcium (mg/L) | $\begin{aligned} & \text { Recove } \\ & \text { Magnesium } \\ & (\mathrm{mg} / \mathrm{L}) \end{aligned}$ | ed water Sulphate (mg/L) | mistry Chloride (mg/L) | $\begin{gathered} \text { DOC } \\ (\mathrm{mg} / \mathrm{L}) \end{gathered}$ |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| No changes |  |  |  |  |  |  |  |  |
|  | 1-stage RO | 10.8 | 131 | 3.0 | 0.5 | 12.4 | 26.3 | 17.5 |
|  | 2-stage RO | 29.5 | 105 | 3.3 | 0.8 | 20.1 | 40.0 | 28.9 |
| Raw water |  |  |  |  |  |  |  |  |
| - doubled Ca \& Mga |  |  |  |  |  |  |  |  |
|  | 1-stage RO | 49.4 | 158 | 3.4 | 0.7 | 10.8 | 64.6 | 14.6 |
|  |  | 357\% | 21\% | 13\% | 34\% | -13\% | 146\% | -17\% |
|  | 2-stage RO | 66.9 | 127 | 3.6 | 1.0 | 17.5 | 97.4 | 24.2 |
|  |  | 127\% | 20\% | 10\% | 29\% | -13\% | 144\% | -16\% |
| - doubled $\mathrm{SO}_{4}$ \& T. Alk. ${ }^{\text {a }}$ |  |  |  |  |  |  |  |  |
|  | 1-stage RO | 29.7 | 133 | 2.2 | 0.5 | 17.3 | 13.9 | 17.2 |
|  |  | 175\% | 1\% | -27\% | 3\% | 40\% | -47\% | -2\% |
|  | 2-stage RO | 43.3 | 107 | 2.3 | 0.8 | 27.6 | 23.4 | 28.5 |
|  |  | 47\% | 1\% | -30\% | 2\% | 37\% | -41\% | -1\% |
| E/P outflow |  |  |  |  |  |  |  |  |
| - doubled Ca \& Mga |  |  |  |  |  |  |  |  |
| 1-stage RO |  | 34.0 | 131 | 3.0 | 0.8 | 12.4 | 26.4 | 17.5 |
|  |  | 214\% | 0\% | 0\% | 62\% | 0\% | 0\% | 0\% |

Table 7.5 (con't)

| Chemical change | Model | Calcium pretreat (mg/L) | $\begin{gathered} \text { RO } \\ \substack{\text { feed flow } \\ \left(\mathrm{m}^{3} / \text { day }\right)} \end{gathered}$ | Calcium (mg/L) | Recove Magnesium ( $\mathrm{mg} / \mathrm{L}$ ) | ed water c Sulphate (mg/L) | mistry Chloride (mg/L) | $\begin{gathered} \text { DOC } \\ (\mathrm{mg} / \mathrm{L}) \end{gathered}$ |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| E/P outflow |  |  |  |  |  |  |  |  |
| - doubled $\mathrm{Ca} \& \mathrm{Mg}\left(\right.$ con't) ${ }^{\text {a }}$ |  |  |  |  |  |  |  |  |
|  | 2-stage RO | $\begin{aligned} & 52.6 \\ & 78 \% \end{aligned}$ | $105$ | $\begin{aligned} & 3.3 \\ & 0 \% \end{aligned}$ | $\begin{aligned} & 1.3 \\ & 62 \% \end{aligned}$ | $\begin{gathered} 20.1 \\ 0 \% \end{gathered}$ | $\begin{gathered} 40.3 \\ 1 \% \end{gathered}$ | $\begin{gathered} 28.9 \\ 0 \% \end{gathered}$ |
| - doubled $\mathrm{SO}_{4}$ \& T. Alk. ${ }^{\text {a }}$ |  |  |  |  |  |  |  |  |
|  | 1 -stage RO | $\begin{aligned} & 36.1 \\ & 235 \% \end{aligned}$ | $\begin{aligned} & 129 \\ & -1 \% \end{aligned}$ | $\begin{gathered} 2.0 \\ -33 \% \end{gathered}$ | $\begin{gathered} 0.5 \\ 5 \% \end{gathered}$ | $\begin{aligned} & 18.2 \\ & 46 \% \end{aligned}$ | $\begin{aligned} & 20.8 \\ & -21 \% \end{aligned}$ | $\begin{gathered} 17.7 \\ 1 \% \end{gathered}$ |
|  | 2-stage RO | $\begin{aligned} & 50.4 \\ & 71 \% \end{aligned}$ | $\begin{aligned} & 103 \\ & -3 \% \end{aligned}$ | $\begin{gathered} 2.0 \\ -40 \% \end{gathered}$ | $\begin{gathered} 0.8 \\ 5 \% \end{gathered}$ | $\begin{aligned} & 30.3 \\ & 51 \% \end{aligned}$ | $\begin{gathered} 24.1 \\ -40 \% \end{gathered}$ | $\begin{gathered} 29.7 \\ 3 \% \end{gathered}$ |
| - doubled DOC ${ }^{\text {a }}$ |  |  |  |  |  |  |  |  |
|  | 1-stage RO | $\begin{gathered} 10.8 \\ 0 \% \end{gathered}$ | $\begin{gathered} 132 \\ 0 \% \end{gathered}$ | $\begin{aligned} & 3.0 \\ & 0 \% \end{aligned}$ | $\begin{aligned} & 0.5 \\ & 4 \% \end{aligned}$ | $\begin{gathered} 12.4 \\ 0 \% \end{gathered}$ | $\begin{gathered} 26.3 \\ 0 \% \end{gathered}$ | $\begin{aligned} & 33.1 \\ & 89 \% \end{aligned}$ |
|  | 2-stage RO | $\begin{gathered} 29.5 \\ 0 \% \end{gathered}$ | $\begin{gathered} 106 \\ 1 \% \end{gathered}$ | $\begin{aligned} & 3.3 \\ & 0 \% \end{aligned}$ | $\begin{gathered} 0.8 \\ 2 \% \end{gathered}$ | $\begin{gathered} 20.0 \\ 0 \% \end{gathered}$ | $\begin{gathered} 39.8 \\ 0 \% \end{gathered}$ | $\begin{aligned} & 54.7 \\ & 89 \% \end{aligned}$ |
| - doubled $\mathrm{Cl}^{\text {a }}$ |  |  |  |  |  |  |  |  |
|  | 1-stage RO | $\begin{gathered} 10.8 \\ 0 \% \end{gathered}$ | $\begin{gathered} 132 \\ 0 \% \end{gathered}$ | $3.0$ | $\begin{gathered} 0.5 \\ 4 \% \end{gathered}$ | $\begin{gathered} 12.4 \\ 0 \% \end{gathered}$ | $\begin{aligned} & 33.3 \\ & 27 \% \end{aligned}$ | $\begin{gathered} 17.4 \\ 0 \% \end{gathered}$ |
|  | 2-stage RO | $\begin{gathered} 29.5 \\ 0 \% \end{gathered}$ | $106$ | $\begin{gathered} 3.3 \\ 0 \% \end{gathered}$ | $\begin{gathered} 0.8 \\ 3 \% \end{gathered}$ | $\begin{gathered} 20.1 \\ 0 \% \end{gathered}$ | $\begin{aligned} & 51.6 \\ & 29 \% \end{aligned}$ | $\begin{gathered} 28.8 \\ 0 \% \end{gathered}$ |

[^6]Table 7.6: Changes in treated water quality, in both the 1 -stage and 2 -stage back-end RO, deepwell, ion-exchange configurations, triggered by given alterations in raw water and effluent plant outflow chemistry.

|  |  |  |  |  |
| :---: | :---: | :---: | :---: | :---: |
| Chemical <br> change | Model | Sulphate <br> $(\mathrm{mg} / \mathrm{L})$ | Sodium <br> $(\mathrm{mg} / \mathrm{L})$ | Chloride <br> $(\mathrm{mg} / \mathrm{L})$ | | DOC |
| :---: |
| $(\mathrm{mg} / \mathrm{L})$ |$\quad$| Total CO |
| :---: |
| $(\mathrm{mol} / \mathrm{L})$ |

## No changes

| Current system | $\mathbf{8 1 . 9}$ | $\mathbf{5 0 . 5}$ | $\mathbf{2 . 4}$ | $\mathbf{6 . 1}$ | $\mathbf{3 . 1} \times \mathbf{1 0}^{-4}$ |
| ---: | :---: | :---: | :---: | :---: | :---: |
| 1-stage RO | 43.3 | 37.1 | 7.5 | 7.4 | $4.7 \times 10^{-4}$ |
| 2-stage RO | 42.0 | 40.5 | 12.5 | 10.2 | $4.7 \times 10^{-4}$ |

Raw water

- doubled $\mathrm{Ca} \& \mathrm{Mg}^{\mathrm{b}}$

1-stage RO
2-stage RO

| 43.4 | 104 |
| :---: | :---: |
| $0 \%$ | $180 \%$ |

18.0

$$
\begin{aligned}
& 7.1 \\
& -4 \% \\
& 9.6 \\
& -6 \%
\end{aligned}
$$

$141 \%$
31.4 151\%
$4.7 \times 10^{-4}$
$4.7 \times 10^{-4}$
$0 \%$

- doubled $\mathrm{SO}_{4} \&$ T. Alk. ${ }^{\text {b }}$

| 1-stage RO | 84.4 | 18.9 | 4.6 | 7.4 | $1.4 \times 10^{-3}$ |
| :--- | :--- | :--- | :--- | :--- | :--- |
|  | $95 \%$ | $-49 \%$ | $-38 \%$ | $0 \%$ | $190 \%$ |
| 2-stage RO | 80.0 | 24.6 | 7.9 | 10.1 | $1.4 \times 10^{-3}$ |
|  | $90 \%$ | $-39 \%$ | $-37 \%$ | $-1 \%$ | $187 \%$ |

## E/P outflow

- doubled $\mathrm{Ca} \& \mathrm{Mg}^{\mathrm{b}}$
1-stage RO
2-stage RO
$\mathrm{O}_{4} \&$ T. Alk.b

| 1-stage RO | 44.6 | 28.8 | 6.2 | 7.4 | $4.7 \times 10^{-4}$ |
| :---: | :---: | :---: | :---: | :---: | :---: |
|  | $3 \%$ | $-22 \%$ | $-17 \%$ | $0 \%$ | $0 \%$ |
| 2-stage RO | 44.7 | 24.8 | 7.9 | 10.3 | $2.5 \times 10^{-4}$ |
|  | $7 \%$ | $-39 \%$ | $-37 \%$ | $1 \%$ | $-48 \%$ |

37.3
$0 \%$
40.8
$1 \%$
7.6
$1 \%$
12.6
$1 \%$
7.4
$0 \%$
10.2
$0 \%$
$4.7 \times 10^{-4}$ $0 \%$
$4.7 \times 10^{-4}$

- doubled $\mathrm{SO}_{4} \&$ T. Alk. ${ }^{\text {b }}$
- doubled DOCb
1-stage RO
43.3
37.2
7.5
10.2
$4.7 \times 10^{-4}$

Table 7.6 (con't)
$\left.\begin{array}{lccccc}\hline \begin{array}{c}\text { Chemical } \\ \text { change }\end{array} & \text { Model } & \begin{array}{c}\text { Sulphate } \\ (\mathrm{mg} / \mathrm{L})\end{array} & \begin{array}{c}\text { Sodium } \\ (\mathrm{mg} / \mathrm{L})\end{array} & \begin{array}{c}\text { Treated water chemistrya } \\ \text { Chloride } \\ (\mathrm{mg} / \mathrm{L})\end{array} & \begin{array}{c}\text { DOC } \\ (\mathrm{mg} / \mathrm{L})\end{array}\end{array} \begin{array}{c}\text { Total CO3 } \\ (\mathrm{mol} / \mathrm{L})\end{array}\right]$
a Calcium and magnesium treated water concentrations not shown since they were held at <1.4 and $0.5 \mathrm{mg} / \mathrm{L}$, repectively.
${ }^{\mathrm{b}} \mathrm{Ca}=$ calcium, $\mathrm{Mg}=$ magnesium, $\mathrm{SO}_{4}=$ sulphate, $\mathrm{T} . \mathrm{Alk} .=$ total alkalinity, $\mathrm{Cl}=$ cloride, Total $\mathrm{CO}_{3}=$ total carbonate content, $\mathrm{DOC}=$ dissolved organic carbon

Figure 7.1: An illustration of the projected flow patterns in a composite discharge, reverse osmosis


ZED system with nanofiltration softening


Figure 7.2: An illustration of the changes in equipment and flow patterns in a given ZED system using nanofiltration instead of ion-exchange softening.

Figure 7.3: An illustration of the projected flow patterns in a back-end, BCS-only ZED configuration.

Figure 7.4: An illustration of the projected flow patterns in a back-end, reverse osmosis


Figure 7.5: A simplified illustration of a 1-stage RO, BCS ZED system.


Figure 7.6: A simplified illustration of a 2-stage RO, BCS ZED system.

### 8.0 ZED IMPLEMENTATION AND IMPLICATIONS

The purpose of this section is to describe the sequential changes required at the FNGP to transform it into a ZED facility, and the rationale for, and/or consequences of, each step.

## STEP 1: Check the accuracy of existing flow meters

The accuracy of some of the FNGP's flow meters has been called into question several times in this report. Specifically, the raw water inflow and treated water outflow meters on the FESs appeared to be mis-calibrated (see Section 5.2.1). Several Westcoast personnel have also indicated that the leak detected at the effluent plant may have been caused by incorrect flow readings instead of actual water losses (Bruce Kosugi, personal com.) These problems need to be investigated and, if necessary, corrected. An accurate and precise water monitoring system is vital to the ZED project. Only with such a system can the validity of the findings described herein be authenticated. Furthermore, one has to be able to properly assess water flow changes predicted to occur in Steps 2 through 7.

## STEP 2: Reuse the 8 psi steam venting off of the deaerators

All of the 8 psi steam generated in the deaerators should be captured at its point of origin. If the hot lime treater is on, then the trapped 8 psi steam should be cooled with a counter-current heat exchanger using raw water heading into the hot lime treater. Using a heat exchanger to cool the 8 psi vapor will ensure that raw water temperatures reach required levels without having to inject 8 psi steam directly into the lime treater. When the lime treater is "off-line", the 8 psi steam should be converted to liquid in an air-cooled condenser; the same type that is already in place at the plant (E/F Manual). To maintain current raw water dilution rates, approximately $170 \mathrm{~m}^{3} /$ day of condensed 8 psi steam should flow into the hot lime treater when it is in operation. Remaining water should be returned to the hotwells for reuse as boiler feedwater.

Reusing the 8 psi steam will substantially reduce both the flow of wastewater and the flux of contaminants through the lime ponds, as well as the demand for treated water make-up (see Section
5.2.2). If this action were carried out in isolation and the plant's wastewater continued to be discharged to the FNR, the former result may cause permit non-compliance problems. Effluent from the lime pond is used to dilute outflow waters from the effluent plant. As the volume of available dilutant drops with 8 psi steam reuse, wastewater released to the FNR may no longer satisfy regulatory guidelines. Sealing the 8 psi steam leaks will reduce raw water demands and improve the efficiency of the existing water network, hence its importance to the ZED project. On its own, this action also has the potential to disrupt current discharge practices should the expected decrease in lime pond effluent trigger non-compliance of the plant's wastewater flow, should it continue to be discharged to the FNR.

## STEP 3: Collect water flow and water chemistry data

Once the 8 psi vents have been sealed, plant personnel should collect water flow and chemical data to evaluate the validity of the simulator's predictions. If. the water network changes as expected, the system should continue to respond accordingly as ZED equipment is added. On the other hand, if the resulting flow patterns are significantly different from those predicted by the simulator, the model should be corrected and reworked to reassess the benefits of a ZED program.

## STEP 4: Bench-scale and pilot-plant testing of pertinent ZED equipment

Small scale testing of ZED equipment, specifically RO membranes and a B/C, will allow for a true evaluation of the robustness and resilience of the available technology to produce clean water under field conditions. These experiments will also provide a better understanding of the required pretreatment trains and energy demands of each ZED process. The resulting data can then be used to evaluate the feasibility of proceeding to full scale ZED implementation.

## STEP 5: Installation of a separate domestic feed pipe on the raw water pipeline

To prevent cross-contamination of domestic feedwaters with recycled wastewater a separate pipeline should be attached to the incoming raw water line. Raw water will then travel directly into the domestic system without contacting reclaimed effluent. Plant personnel have already identified
a connection valve on the existing system where the new feedline could be connected and the equipment necessary to set-up an independent domestic feed pipe (ZED Report).

## STEP 6: Attach RO filters to the water network (if applicable)

Regardless of whether the chosen ZED design is a BCS or deep well RO configuration, it may be desirable to install the RO filtration system before either a deep well or BCS assembly is purchased or installed, and to operate the plant with only the RO units attached. RO technology is cheaper than the BCS and possibly a deep well. By running the plant with only the RO vessels attached, the plant can examine the potential effectiveness of a full-scale ZED program before larger amounts of capital are invested in the project. Furthermore, as indicated by the deep well ZED simulations, a smaller waste disposal system will be required if wastewater recycling starts with only the RO filters in place, since wastewater production rates will drop as soon as reclaimed wastewater begins to cycle back through the plant (Tables $7.1 \& 7.4$ ). . Starting the ZED process after installing only the RO technology should, therefore, highlight any potential limitations of a full-scale ZED program prior to purchasing the more costly final disposal equipment, as well as reduce the size and cost of these expensive systems.

Operating the FNGP without a means of disposing of RO concentrates on site will necessitate their discharge to the FNR. As the RO waste stream will be more concentrated than the current FNR discharge wastewater, a special temporary discharge permit may be required. Alternatively, RO concentrates could be diluted with raw water to meet current regulations, if they indeed prove to be too contaminated to be released on their own. The advantages of beginning the ZED program with only the RO equipment in place must be considered with due respect to potential RO concentrate disposal problems.

## STEP 7: Install a BCS or deep well (whichever is appropriate)

Constructing a deep well will complete the transformation of the FNGP to a ZED system. Concerns about wastewater disposal now shift from the FNR to subsurface migration. ZED implies that no liquid waste will leave the FNGP. The deep well, therefore, has to be built to
ensure that contaminants transferred to the subsurface do not migrate off-site, a difficult and potentially expensive task.

Similarly, although a BCS assembly will produce a solid, rather than a liquid, waste, the waste stream will still require appropriate disposal. The generated solids will be water soluble; if they are left exposed to the elements, they will readily re-dissolve. Should the solid waste liquefy and drain into the FNR, the primary objective of the ZED project (to eliminate the flow of contaminants from the FNGP into the FNR) will not have been met. Furthermore, the discharge of previously solidified contaminants into the FNR may prove to be more harmful to the river's ecosystem than the plant's current, relatively inert outflow, since the collected waste would be carried into the river during infrequent rainfalls or snow melts. The flux of material entering the river would be much higher during these storm events than it is now. For these reasons, solid waste produced from a BCS ZED program cannot be stored in exposed above ground stock-piles.

Burying the solid waste may not, on its own, constitute a sufficient disposal mechanism. Water percolating through the cover layer will cause some of the underlying solids to dissolve. Unless the disposal pit has been adequately lined with an impermeable barrier, the nowcontaminated water will continue its downward migration until it encounters a groundwater flow and subsequently migrates off-site. As previously indicated, off-site migration of wastewater violates the principal of a ZED facility. Although, by regulatory standards, the solid waste generated in a BCS ZED system would not be a hazardous or special waste, it clearly must be handled and disposed of correctly because, if it re-dissolves and exits the FNGP compound, none of the original objectives of the ZED project will have been met.

Transforming the FNGP to a ZED facility is a relatively straight forward process. There are, however, consequences associated with such a conversion which must be considered. Final waste management is a critical issue. All the plant's wastewater is currently treated on-site and released to the FNR. Waste products from either a BCS or deep well ZED design would, on the other hand, need a far more stringent disposal system. Waste management would require the constant
attention of plant personnel as these wastes would be retained on site indefinitely, unlike currently treated wastewater.

### 9.0 POTENTIAL LIMITATIONS

The analyses described herein, from constructing water balances and identifying opportunities for optimization to developing a computer model and choosing the most suitable ZED design from the resulting simulations, are all ultimately based on the original water flow and water chemistry data collected at the plant. There are potential limitations to these data and the way in which they have been used which need to be considered. First and foremost, the data may not be totally representative of the true situation at the FNGP. Some of the data were derived from assumptions and calculations rather that from actual measurements. Furthermore, only average values were utilized in this investigation. These limitations, as well as those associated with the computer simulator and the ZED design selection process, are discussed below.

### 2.1 Non-Representative Numbers

### 9.1.1 Flow data

The flow data used to characterize the FNGP initially appeared to contain some unrealistic or non-representative values, as illustrated by large discrepancies between some metered and calculated flow numbers (e.g. Table 5.10). Replacing problematic metered data with calculated values eliminated most of the inconsistencies. Plant personnel have repeatedly reviewed the constructed water balances, and it is unlikely that any serious flaws continue to exist.

### 9.1.2 Chemical data

The inclusion of non-representative chemical values within this study is possible considering that several areas of the plant were incorrectly sampled or described by only one or two samples. A large number of inconsistencies were indeed observed in the initial mass balances (Tables 5.4 through 5.8). A re-examination of the water system indicated that water additives and other contaminating agents had been overlooked and that the characteristics of the mis-sampled waters could be estimated from information in several operational manuals at the plant. When the mass balances were reworked with this new data, a number of the earlier discrepancies disappeared.

The corrected data set seemed to adequately detail this facility's water chemistry. Further sampling and subsequent analyses will, however, be required to fully test this hypothesis.

### 9.2 Sample Variability

### 9.2.1 Water data

All water flows were expressed as daily averages, including inconsistent pathways such as blowdowns from the boilers, ion-exchangers and hot lime treater. This transformation was necessary for the construction of the water and mass balances. Flow variability does, however, become important in sizing the required ZED equipment. The treatment system has to be designed to handle the largest expected flows. One reason for focusing the ZED program around the FNR discharge, instead of individual drainage points, was to avoid using costly, over designed, cleaning processes which would have to be big enough to treat large, but infrequent, feedwater volumes. The lime ponds and effluent plant produce much more consistent outflows. Nevertheless, the volume of wastewater discharged from the plant is not constant (Table 5.9), nor, according to plant personnel, is the amount of treated water lost from the thermo-oxidizer (Shang Su , personal com.). To assess the potential impact flow variation may have on a ZED configuration, a sensitivity analysis was performed using the back-end, 1 -stage RO, BCS/I-X design.

Some flow changes, such as increased treated water losses to the stack, did not significantly alter the size of the required ZED equipment (Table 9.1), although they produced large changes elsewhere in the system (e.g. increased raw water demands and/or solid waste output). On the other hand, increasing other flow paths to account for observed variability did indeed necessitate a bigger ZED treatment train (Table 9.1). As illustrated by these findings, the chosen ZED scenario will have to be designed appropriately to account for water flow variability.

### 9.2.2 Chemical data

Varying contaminant concentration will similarly change ZED equipment sizing requirements. Doubling raw water calcium and magnesium content, for example, resulted in an $11 \%$ increase in

RO feedwater flows in the back-end, RO, BCS/I-X models (Table 7.2). The higher calcium and magnesium levels triggered increased softener and boiler blowdowns, which then had to be treated in larger ZED vessels.

Clearly fluctuations in both water flow and water chemistry must be considered when designing the final ZED program, as they will likely affect the sizing of the required equipment. Apart from some testing on the effects of variability, this study was, in the main, restricted to initial design and comparison work, and, as such, did not incorporate actual plant variability into the analyses.

### 2.3 Computer Model

Although all bugs appeared to have been removed from the computer simulator prior to its use in evaluating the proposed ZED designs, there are numerous assumptions built into the simulator. They range from the obvious (e.g: chloride levels in the lime treater product waters equal those in the raw water) to the potentially erroneous (e.g. treated water from the ion-exchangers will always have a calcium concentration of $1.4 \mathrm{mg} / \mathrm{L}$ ). The value of the model's output is dependent on its ability to mirror the real water system; if the simulator does not react to change the same way the actual water network would, then it has no predictive value. The realism of the model is therefore dependent on the assumptions and calculations used in its construction. If these "building blocks" are valid, then the results are valuable. The validity of the model, its output, and the resulting conclusions remain unproven at this point. In other words, although the simulator reacted as expected to given changes in the virtual water system, there is no way to know if this is how the plant's actual water network will react until the same changes are made at the plant. The predictive ability of the simulator should therefore be further examined after the 8 psi steam vents are closed.

A more obvious limitation of the computer model is that there is no time lag incorporated into the system. The lime ponds, effluent plant and polishing pond have hydraulic retention times of 5, 20, and 1 day(s), respectively (E/P Manual). Changes in the water distribution network would not occur as rapidly as they appear to in the model. Calcium pretreatment systems were also never
fully incorporated into the model. The extent of included pretreatment was the removal of calcium from RO feedwaters, and its inclusion in the RO concentrate stream. Other influences from the removal process were ignored. For example, regeneration wastes from (say) an ion-exchanger or chemical changes resulting from the use of an anti-scalant to immobilize the problematic calcium were not accounted for in the simulator. Although their absence undoubtedly alters the absolute precision of the model, it is unlikely that the inclusion of these small missing flows, and/or chemical changes, would have changed the overall conclusions of this project.

### 2.4 ZED Design Evaluation

The ZED design selection process included a general discussion of system price. The cost of a particular ZED program was limited to the capital investment required to purchase the key pieces of equipment, including the RO, BCS and/or NF units. RO pretreatment costs were limited to mixedmedia filters and ion-exchange softeners (Appendix G.). Although cartridge filters, chlorination and dechlorination stations, as well as pH adjustment stations were included in several ZED diagrams (i.e. Figs. 7.1 \& 7.4), it is unclear at this stage what pretreatment systems will be needed at the FNGP beyond a mixed-media filter and inlet softener ${ }^{1}$. As a result, no cost estimates were made for these other units. They were, however, included in the illustrations to indicate that an RO pretreatment train is likely to include several different processes as suggested by Applegate (1984), Pohland (1987) and Suemoto et. al. (1994).

Aside from certain parts of the hypothesized RO pretreatment trains, deep well installation, solid waste disposal and general system operating costs were never evaluated. Given that this study was initiated as a "first look" into transforming the FNGP into a ŻED facility, the reported costs should be sufficient to allow Westcoast personnel to decide if a ZED program is feasible and which particular option to pursue.

[^7]
### 9.5. Conclusion

There are potential limitations inherent in this investigation. The data base on which this report is founded may contain non-representative information. The computer simulator used to test the 18 proposed ZED designs is a yet untested model of the FNGP's water distribution network which instantaneously adjusts to any change, and a very rudimentary cost analysis was incorporated into the ZED design selection process. On the other hand, plant personnel have reviewed the collected data and seems satisfied with the reported values. The simulator continued to respond as expected to changes in the plant's configuration; the pricing system used herein was sufficient to fulfill the primary objective of this project, which was to show that the FNGP can be transformed into a ZED facility for a given amount of money. Finally, sample variability, although not taken into account during this study for previously mentioned reasons, is bound to be important in ZED equipment sizing, as demonstrated in sensitivity analyses described above. Further research should therefore be performed prior to installing a ZED system at the FNGP to better define water flow and water chemistry fluctuations, as well as to better assess ZED pretreatment requirements.

Table 9.1: Changes to the configuration of the back-end, BCS, 1-stage RO, ion-exchange design with various flow alterations.

| Flow change | $\begin{aligned} & \text { R.W. } \\ & \text { flow } \\ & \left(\mathrm{m}^{3} / \mathrm{day}\right) \end{aligned}$ | Calcium pretreat ${ }^{\text {a }}$ ( $\mathrm{mg} / \mathrm{L}$ ) | $\begin{gathered} \text { RO } \\ \substack{\text { f.water } \\ \left(\mathrm{m}^{3} / \mathrm{day}\right)} \end{gathered}$ | $\begin{gathered} \text { B/C } \\ \text { f.watera } \\ \left(\mathrm{m}^{3} / \text { day }\right) \end{gathered}$ | $\begin{gathered} \text { Waste } \\ \text { water } \\ \left(\mathrm{m}^{3} / \text { day }\right) \end{gathered}$ | oducts solids (kg/day) |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| Initial outlay |  |  |  |  |  |  |
| No change | 407 | 48.2 | 282 | 141 | 14.1 | 415 |
| No 8 psi steam reuse | 110\% | 146\% | 26\% | 26\% | 25\% | 68\% |
| Double T/O losses ${ }^{\text {b }}$ | 48\% | 79\% | 7\% | 7\% | 7\% | 24\% |
| Double E/P ouflow ${ }^{\text {b }}$ | 0\% | -44\% | 39\% | 39\% | 39\% | 36\% |
| Double lime pond | 47\% | - | 81\% | 81\% | 81\% | 77\% |
| Full recycle |  |  |  |  |  |  |
| No change | 138 | 0.0 | 240 | 120 | 12.0 | 234 |
| No 8 psi steam reuse | 275\% | 71.0 | 27\% | 27\% | 27\% | 103\% |
| Double T/O losses ${ }^{\text {b }}$ | 130\% | 23.9 | 9\% | 10\% | 9\% | 42\% |
| Double E/P ouflow ${ }^{\text {b }}$ | -76\% | 0\% | 41\% | 41\% | 41\% | 55\% |
| Double lime pond | -23\% | 0\% | 53\% | 53\% | 53\% | 16\% |

[^8]
### 10.0 CONCLUSIONS

Data collected from the FNGP indicated that operational water losses are greater than domestic wastewater production rates, so the plant can theoretically become a ZED facility. The data also showed that there are 2 leak points in the plant's water distribution network. One was in the steam tracing lines, and the second one was within the effluent plant. It is doubtful that closing off either escape route will be of economic benefit to the ZED project; the leak in the steam tracing pipes is too small to be of any consequence, and recovering wastewater lost from the effluent plant will only increase the flow of domestic wastewater into a ZED system. The increased domestic flow will not jeopardize the success of a ZED program at the plant, but it will necessitate a larger treatment system to purify the water and generate a larger final waste stream than the current FNR discharge flow. Unfortunately the FNGP will only be recognized as a ZED facility once the leak at the effluent plant has been eliminated.

Direct wastewater reuse is limited to the 8 psi steam currently discharged from the deaerators. Collecting, condensing and reusing this water should significantly reduce raw water inflow and wastewater outflow rates, which will simplify the design of the ultimate ZED treatment train. There is, however, an inherent risk that reusing this steam may lead to the non-compliance of the FNR discharge stream, as detailed in Section 9.0-Step 2. Due caution must be exercised if the 8 psi steam is reused while continuing to discharge the plant's wastewaters to the FNR. Although water escaping from the thermo-oxidizer is another potentially reusable waste stream, further research is needed to determine if reusing this water will compromise the plant's gas processing efficiency or the ability to transform this facility over to ZED framework. Rather than tackling the remaining wastewater flows at their points of origin, the ZED project focused on recycling the FNR discharge.

Eighteen prospective ZED designs were developed and subsequently tested with a computer simulator. The best deep well configuration was a back-end, 1 -stage RO unit with an ion-
exchange softener in the plant's front-end softening system. This design was estimated to cost $\$ 101700$ US dollars, excluding the cost of a deep well. The best BCS models were the back-end, 1 -stage RO and BCS-only options, both of which maintained an ion-exchanger in the front-end softening system. They were estimated to cost approximately $\$ 1.61 \& \$ 1.79$ million US dollars, respectively.

Transforming the FNGP into a ZED facility is, on its own, a simple process; the appropriate equipment just needs to be incorporated into the water system. There are, however, consequences to a ZED program. The most important is final waste management. If a deep well solution is used, the well itself will have to be build to prevent off-site migration of disposal wastes, a potentially expensive endeavor. Similarly, the solid waste generated by a BCS assembly will have to be disposed of properly to prevent any of it from resolubilizing and migrating off-site. Once a ZED protocol is initiated, wastewater or waste solids produced at the FNGP will no longer disappear with the FNR; they will remain on-site indefinitely.

### 11.0 RECOMMENDATIONS

1) Close the 8 psi steam loop regardless of whether or not a ZED program is adopted at the plant, so long as this process can occur without disrupting current wastewater disposal practices.
2) Westcoast Energy Inc. needs to assess the viability of transforming the FNGP into a ZED facility in light of the involved costs and final waste disposal requirements.
3) If the plant is to be transformed into a ZED facility, implement the ZED program as specified in Chapter 8.0.

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APPENDIX A

WATER FLOW DATA
Effluent Plant Flows

|  | Date | $\begin{gathered} \text { Oil } \\ \text { pond } \end{gathered}$ | Treated wastewater | Lime pond | Composite discharge | Nov | Date | $\begin{gathered} \text { Oil } \\ \text { pond } \end{gathered}$ | Treated wastewater | Lime pond | Composite discharge |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| 1993 | 1 | 104 | 110 | 157 | 277 |  | 1 | 175 | 113 | 175 | 288 |
| Oct | 2 | 107 | 118 | 124 | 251 |  | 2 | 175 | 119 | 238 | 357 |
|  | 3 | 147 | 104 | 144 | 257 |  | 3 | 174 | 160 | 219 | 387 |
|  | 4 | 169 | 109 | 212 | 330 |  | 4 | 162 | 135 | 190 | 334 |
|  | 5 | 191 | 129 | 162 | 300 |  | 5 | 154 | 121 | 204 | 334 |
|  | 6 | 147 | 124 | 143 | 276 |  | 6 | 150 | 110 | 175 | 294 |
|  | 7 | 145 | 144 | 145 | 299 |  | 7 | 175 | 175 | 175 | 351 |
|  | 8 | 191 | 175 | 119 | 294 |  | 8 | 189 | 113 | 160 | 282 |
|  | 9 | 206 | 142 | 141 | 292 |  | 9 | 242 | 122 | 160 | 291 |
|  | 10 | 208 | 171 | 167 | 349 |  | 10 | 213 | 111 | 257 | 371 |
|  | 11 | 120 | 179 | 186 | 373 |  | 11 | 248 | 95 | 251 | 352 |
|  | 12 |  |  |  |  |  | 12 | 287 | 101 | 218 | 328 |
|  | 13 | 180 | 160 | 211 | 381 |  | 13 | 280 | 100 | 130 | 239 |
|  | 14 | 206 | 147 | 206 | 353 |  | 14 | 215 | 99 | 158 | 265 |
|  | 15 | 222 | 133 | 147 | 280 |  | 15 | 207 | 112 | 158 | 280 |
|  | 16 | 191 | 119 | 107 | 226 |  | 16 | 228 | 143 | 210 | 361 |
|  | 17 | 191 | 133 | 119 | 313 |  | 17 | 90 | 163 | 114 | 271 |
|  | 18 | 191 | 138 | 196 | 342 |  | 18 | 92 | 172 | 88 | 259 |
|  | 19 | 201 | 129 | 176 | 310 |  | 19 | 120 | 157 | 106 | 259 |
|  | 20 | 176 | 123 | 198 | 329 |  | 20 | 121 | 164 | 99 | 262 |
|  | 21 | 177 | 169 | 156 | 336 |  | 21 | 121 | 164 | 99 | 262 |
|  | 22 | 178 | 113 | 189 | 337 |  | 22 | 171 | 150 | 183 | 333 |
|  | 23 | 67 | 120 | 170 | 301 |  | 23 | 199 | 159 | 144 | 303 |
|  | 24 | 74 | 123 | 133 | 264 |  | 24 | 222 | 153 | 170 | 322 |
|  | 25 | 131 | 92 | 169 | 268 |  | 25 | 234 | 151 | 196 | 346 |
|  | 26 | 125 | 109 | 163 |  |  | 26 | 156 | 165 | 168 | 332 |
|  | 27 | 199 | 99 | 227 | 335 |  | 27 | 136 | 169 | 216 | 385 |
|  | 28 | 119 | 110 | 210 | 338 |  | 28 | 99 | 158 | 201 | 359 |
|  | 29 | 197 | 98 | 183 | 290 |  | 29 | 96 | 158 | 184 | 340 |
|  | 30 | 191 | 107 | 107 | 214 |  | 30 | 97 | 174 | 166 | 340 |
|  | 31 | 175 | 100 | 238 | 338 |  |  |  |  |  |  |





Effluent Plant Flows

Effluent Plant Flows

Effluent Plant Flows

| Date | $\begin{gathered} \text { Oil } \\ \text { pond } \end{gathered}$ | Treated wastewater | Lime pond | Composite discharge |  | Date | Oil <br> pond | Treated wastewater | Lime pond | Composite discharge |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| 1 | 146 | 119 | 206 | 325 | 1995 | 1 | 175 | 83 | 175 | 258 |
| 2 | 146 | 119 | 272 | 391 | Jan | 2 | 175 | 94 | 175 | 269 |
| 3 | 160 | 119 | 206 | 325 |  | 3 | 175 | 94 | 206 | 300 |
| 4 | 136 | 92 | 223 | 315 |  | 4 | 94 | 94 | 175 | 269 |
| 5 | 140 | 82 | 222 | 304 |  | 5 | 94.3 | 94.3 | 238 | 332 |
| 6 | 161 | 94 | 290 | 384 |  | 6 | 83 | 94 | 308 | 402 |
| 7 | 154 | 147 | 206 | 353 |  | 7 | 161 | 107 | 206 | 313 |
| 8 | 118 | 129 | 221 | 350 |  | 8 | 100 | 119 | 176 | 295 |
| 9 | 99 | 109 | 233 | 342 |  | 9 | 94 | 94 | 206 | 345 |
| 10 | 119 | 119 | 235 | 239 |  | 10 | 94 | 94 | 205 | 349 |
| 11 | 94 | 147 | 272 | 419 |  | 11 | 119 | 119 | 147 | 266 |
| 12 | 94 | 146 | 206 | 352 |  | 12 | 169 | 107 | 175 | 275 |
| 13 | 94 | 133 | 238 | 371 |  | 13 | 206 | 119 | 222 | 341 |
| 14 | 94 | 139 | 190 | 329 |  | 14 | 206 | 119 | 272 | 391 |
| 15 | 94 | 139 | 307 | 446 |  | 15 | 238 | 119 | 206 | 354 |
| 16 | 120 | 133 | 272 | 408 |  | 16 | 206 | 119 | 222 | 341 |
| 17 | 120 | 133 | 272 | 405 |  | 17 | 191 | 119 | 272 | 391 |
| 18 | 175 | 133 | 222 | 355 |  | 18 | 175 | 107 | 308 | 415 |
| 19 | 146 | 146 | 175 | 322 |  | 19 | 147 | 107 | 326 | 433 |
| 20 | 119 | 119 | 206 | 325 |  | 20 | 147 | 107 | 308 | 415 |
| 21 | 94 | 119 | 272 | 391 |  | 21 | 94 | 133 | 281 | 414 |
| 22 | 146 | 94 | 287 | 332 |  | 22 | 147 | 133 | 383 | 516 |
| 23 | 140 | 119 | 238 | 357 |  | 23 | 175 | 133 | 238 | 371 |
| 24 | 140 | 113 | 206 | 319 |  | 24 | 175 | 133 | 308 | 441 |
| 25 | 119 | 113 | 238 | 351 |  | 25 | 208 | 107 | 206 | 313 |
| 26 | 94 | 113 | 255 | 368 |  | 26 | 308 | 94 | 238 | 332 |
| 27 | 83 | 133 | 238 | 371 |  | 27 | 272 | 72 | 308 | 380 |
| 28 | 238 | 133 | 206 | 339 |  | 28 | 238 | 94 | 206 | 300 |
| 29 | 147 | 107 | 222 | 329 |  | 29 | 94 | 94 | 272 | 366 |
| 30 | 146 | 107 | 206 | 3.13 |  | 30 | 94 | 146 | 214 | 360 |
| 31 | 175 | 83 | 175 | 258 |  | 31 | 83 | 133 | 206 | 339 | ロ


Data from Fort Nelson Gas Plant Powerhouse Morning Reports
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Data from Fort Nelson Gas Plant Powerhouse Morning Reports

| Date | Steam Generation |  |  | Boiler feedwater |  |  | Excess 45 <br> psi steam <br> KLBH | Raw Water IGPH | Treated Water IGPH | Stack Emissions MMSCFD |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  | Boiler 5 | Boilér 6 | Boiler 7 | Boiler 5 | Boiler 6 | Boiler 7 |  |  |  |  |
|  | KLBH | KLBH | KLBH | KLBH | KLBH | KLBH |  |  |  |  |
| 1 | 266.0 | 268.0 | 266.0 | 261.0 | 260.6 | 260.6 | 64.2 | 5005 | 7088 | 6.8 |
| 2 | 261.0 | 264.0 | 261.0 | 239.0 | 239.3 | 246.4 | 64.8 | 5783 | 7875 | 6.2 |
| 3 | 249.0 | 251.0 | 251.5 | 252.0 | 244.8 | 247.2 | 40.1 | 4475 | 5700 | 6.3 |
| 4 | 270.0 | 288.0 | 250.0 | 287.0 | 261.8 | 272.0 | 38.2 | 9820 | 12600 | 7.5 |
| 5 | 286.0 | 276.0 | 276.0 | 301.0 | 285.1 | 294.8 | 34.9 | 9589 | 12600 | 7.5 |
| 6 | 319.0 | 215.5 | 307.0 | 320.0 | 205.2 | 312.4 | 29.8 | 9944 | 12600 | 7.0 |
| 7 | 318.0 | 216.0 | 304.0 | 315.0 | 209.5 | 300.3 | 29.2 | 7078 | 1750 | 7.0 |
| 8 | 294.0 | 217.0 | 294.0 | 291.0 | 207.8 | 289.2 | 43.1 | 0 | 0 | 7.3 |
| 9 | 256.0 | 271.0 | 283.0 | 269.0 | 265.5 | 279.5 | 51.6 | 6194 | 8400 | 7.3 |
| 10 | 290.0 | 286.0 | 291.0 | 287.0 | 277.9 | 294.5 | 58.5 | 10235 | 12600 | 7.0 |
| 11 | 292.0 | 288.0 | 292.0 | 291.0 | 287.8 | 293.3 | 61.3 | 9900 | 12600 | 7.0 |
| 12 | 293.0 | 288.0 | 291.0 | 295.0 | 287.0 | 296.2 | 62.3 | 807 | 600 | 7.3 |
| 13 | 294.0 | 291.0 | 294.0 | 285.0 | 281.9 | 293.6 | 61.7 | 4490 | 6300 | 7.0 |
| 14 | 298.0 | 289.0 | 289.0 | 316.0 | 295.0 | 306.6 | 61.4 | 7692 | 12300 | 7.0 |
| 15 | 311.0 | 298.0 | 296.0 | 308.0 | 290.5 | 296.8 | 74.0 | 8768 | 12300 | 7.4 |
| 16 | 311.0 | 297.0 | 296.0 | 314.0 | 292.0 | 299.0 | 82.1 | 177 | 256 | 7.1 |
| 17 | 311.0 | 297.0 | 296.0 | 315.0 | 290.5 | 295.0 | 83.5 | 4635 | 5638 | 7.0 |
| 18 | 311.0 | 298.0 | 296.0 | 310.0 | 288.8 | 296.0 | 88.9 | 8908 | 12300 | 7.1 |
| 19 | 309.0 | 287.0 | 295.0 | 313.0 | 300.3 | 300.2 | 85.7 | 8702 | 12300 | 7.1 |
| 20 | 308.0 | 295.0 | 293.0 | 296.0 | 275.5 | 291.6 | 84.4 | 588 | 513 | 6.6 |
| 21 | 313.0 | 299.0 | 297.0 | 317.0 | 296.4 | 301.6 | 86.3 | 6414 | 9450 | 7.1 |
| 22 | 325.0 | 312.0 | 309.0 | 320.0 | 302.6 | 311.1 | 88.3 | 9023 | 12600 | 7.4 |
| 23 | 323.0 | 309.0 | 307.0 | 315.0 | 292.0 | 305.7 | 82.8 | 9481 | 12600 | 7.3 |
| 24 | 321.0 | 308.0 | 306.0 | 322.0 | 298.3 | 305.5 | 80.8 | 4171 | 5513 | 7.3 |
| 25 | 322.0 | 308.0 | 306.0 | 322.0 | 303.1 | 310.5 | 80.9 | 3200 | 4200 | 7.5 |
| 26 | 323.0 | 310.0 | 308.0 | 326.0 | 306.0 | 309.2 | 88.3 | 8779 | 12600 | 7.4 |
| 27 | 316.0 | 303.0 | 301.0 | 309.0 | 292.6 | 291.7 | 87.9 | 10022 | 12600 | 7.0 |
| 28 | 308.0 | 296.0 | 293.0 | 319.0 | 303.6 | 310.8 | 81.4 | 6862 | 8925 | 7.4 |

Data from Fort Nelson Gas Plant Powerhouse Morning Reports

|  | Steam Generation |  |  |  | Boiler feedwater |  |  | Excess 45 |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | Raw 를

Data from Fort Nèlson Gas Plant Powerhouse Morning Reports

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Data from Fort Nelson Gas Plant Powerhouse Morning Reports

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|  |  |

Data from Fort Nelson Gas Plant Powerhouse Morning Reports

| Date | Steam Generation |  |  | Boiler feedwater |  |  | Excess 45 psi steam KLBH | Raw <br> Water <br> IGPH | Treated Water IGPH | Stack <br> Emissions MMSCFD |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  | Boiler 5 <br> KLBH | Boiler 6 <br> KLBH | Boiler 7 <br> KLBH | Boiler 5 <br> KLBH | Boiler 6 <br> KLBH | Boiler 7 <br> KLBH |  |  |  |  |
| 1 | 287.0 | 286.0 | 281.0 | 295.0 | 290.3 | 294.3 | 88.0 | 3835 | 7222 | 7.4 |
| 2 | 298.0 | 300.0 | 295.0 | 295.0 | 287.6 | 294.8 | 97.8 | 5375 | 9269 | 7.3 |
| 3 | 298.0 | 300.0 | 294.0 | 299.0 | 290.9 | 291.0 | 107.1 |  |  | 6.9 |
| 4 | 292.0 | 296.0 | 289.0 | 295.0 | 285.3 | 287.6 | 104.0 | 5427 | 8700 | 7.2 |
| 5 | 297.0 | 299.0 | 293.0 | 293.0 | 287.6 | 295.0 | 109.2 | 5671 | 8400 | 7.4 |
| 6 | 296.0 | 299.0 | 293.0 | 291.0 | 281.9 | 291.0 | 110.1 | 5265 | 8400 | 7.3 |
| 7 | 293.0 | 296.0 | 293.0 | 300.0 | 282.7 | 290.0 | 106.8 | 6931 | 8400 | 7.4 |
| 8 | 298.0 | 295.0 | 298.0 | 294.0 | 283.8 | 287.3 | 108.4 | 5850 | 8400 | 7.2 |
| 9 | 298.0 | 295.0 | 290.0 | 298.0 | 281.1 | 293.9 | 111.3 | 5828 | 8400 | 7.3 |
| 10 | 298.0 | 296.0 | 294.0 | 290.0 | 297.0 | 282.2 | 289.8 | 5124 | 8400 | 7.2 |
| 11 | 299.0 | 296.0 | 288.0 | 297.0 | 282.4 | 288.1 | 113.6 | 4363 | 9000 | 7.3 |
| 12 | 295.0 | 293.0 | 284.0 | 293.0 | 281.3 | 287.1 | 114.4 | 3765 | 6600 | 7.1 |
| 13 | 296.0 | 293.0 | 285.0 | 295.0 | 285.3 | 280.9 | 115.9 |  | 6600 | 7.2 |
| 14 | 300.0 | 296.0 | 288.0 | 306.0 | 286.6 | 292.1 | 112.5 | 3999 | 6900 | 7.4 |
| 15 | 303.0 | 299.0 | 291.0 | 299.0 | 292.9 | 288.2 | 110.7 | 3679 | 6600 | 6.8 |
| 16 | 302.0 | 299.0 | 291.0 | 304.0 | 281.2 | 288.3 | 125.2 |  |  | 7.1 |
| 17 | 291.0 | 288.0 | 279.0 | 280.0 | 265.9 | 277.2 | 125.1 | 7977 | 11400 | 7.2 |
| 18 | 288.0 | 284.0 | 276.0 | 285.0 | 272.3 | 269.4 | 122.0 | 5199 | 6563 | 7.2 |
| 19 | 296.0 | 291.0 | 283.0 | 278.0 | 261.9 | 263.1 | 115.0 | 8484 | 11400 | 7.2 |
| 20 | 286.0 | 283.0 | 274.0 | 285.0 | 64.2 | 268.7 | 114.7 | 7784 | 10200 | 7.4 |
| 21 | 286.0 | 282.0 | 274.0 | 285.0 | 271.3 | 274.1 | 94.2 | 7550 | 10500 | 7.4 |
| 22 | 292.0 | 291.0 | 281.0 | 285.0 | 269.6 | 268.5 | 78.3 | 8747 | 11700 | 7.1 |
| 23 | 296.0 | 293.0 | 285.0 | 286.0 | 280.2 | 279.6 | 81.1 | 8591 | 12000 | 7.5 |
| 24 | 295.0 | 293.0 | 284.0 | 283.0 | 269.6 | 274.1 | 78.6 | 9261 | 12000 | 7.6 |
| 25 | 283.0 | 284:0 | 273.0 | 281.0 | 282.1 | 286.9 | 65.9 | 8952 | 11850 | 8.0 |
| 26 | 289.0 | 293.0 | 292.0 | 286.0 | 281.9 | 287.5 | 57.7 | 8182 | 10500 | 7.8 |
| 27 | 283.0 | 290.0 | 288.0 | 288.0 | 273.8 | 285.9 | 68.4 | 8854 | 12000 | 7.5 |
| 28 | 286.0 | 293.0 | 291.0 | 281.0 | 281.1 | 287.5 | 79.9 | 7392 | 10200 | 7.5 |
| 29 | 289.0 | 295.0 | 293.0 | 283.0 | 276.2 | 287.2 | 89.0 | 7321 | 9600 | 7.5 |
| 30 | 290.0 | 296.0 | 293.0 | 293.0 | 290.2 | 297.8 | 89.3 | 7255 | 9600 | 7.9 |

Data from Fort Nelson Gas Plant Powerhouse Morning Reports

| July | Date | Steam Generation |  |  | Boiler feedwater |  |  | Excess 45 psi steam KLBH | Raw Water IGPH | Treated Water IGPH | Stack Emissions MMSCFD |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  |  | Boiler 5 KLBH | Boiler 6 KLBH | Boiler 7 <br> KLBH | Boiler 5 <br> KLBH | Boiler 6 <br> KLBH | Boiler 7 <br> KLBH |  |  |  |  |
|  | 1 | 302.0 | 306.0 | 304.0 | 299.0 | 288.2 | 304.1 | 87.9 | 8321 | 10800 | 5.5 |
|  | 2 | 226.0 | 156.5 | 231.5 | 278.0 | 0.0 | 282.6 | 71.1 | 6000 | 8400 | 5.7 |
|  | 3 | 279.0 | 0.0 | 285.0 | 282.0 | 0.0 | 276.5 | 52.3 | 7851 | 11100 | 5.6 |
|  | 4 | 277.0 | 9.6 | 283.0 | 282.0 | 0.0 | 290.7 | 39.7 | 7496 | 10200 | 5.5 |
|  | 5 | 256.0 | 6.4 | 264.0 | 246.0 | 0.0 | 251.4 | 68.2 | 4402 | 5850 | 3.7 |
|  | 6 | 255.0 | 0.0 | 262.0 | 256.0 | 0.0 | 258.0 | 88.7 | 0 | 0 | 4.5 |
|  | 7 | 268.0 | 0.0 | 271.0 | 260.0 | 0.0 | 263.3 | 66.7 | 6281 | 9000 | 5.7 |
|  | 8 | 261.0 | 0.0 | 265.0 | 262.0 | 0.0 | 266.5 | 42.7 | 7592 | 10200 | 5.7 |
|  | 9 | 260.0 | 0.9 | 268.0 | 252.0 | 0.0 | 261.1 | 29.8 | 6656 | 8400 | 5.7 |
|  | 10 | 258.0 | 0.0 | 265.0 | 264.0 | 0.0 | 263.3 | 23.1 |  |  | 5.6 |
|  | 11 | 261.0 | 0.0 | 268.0 | 251.0 | 0.0 | 264.3 | 19.1 | 7500 | 6307 | 5.8 |
|  | 12 | 255.5 | 0.0 | 263.0 | 251.0 | 0.0 | 255.3 | 15.6 | 4675 | 6900 | 5.7 |
|  | 13 | 240.0 | 6.1 | 247.5 | 243.0 | 0.0 | 253.4 | 36.6 | 3477 | 7248 | 4.1 |
|  | 14 | 224.0 | 16.1 | 232.0 | 220.0 | 0.0 | 223.2 | 50.5 | 4770 | 6669 | 3.1 |
|  | 15 | 216.0 | 14.7 | 224.0 | 229.0 | 0.0 | 234.2 | 76.9 |  |  | 3.4 |
|  | 16 | 234.0 | 21.3 | 241.0 | 232.0 | 0.0 | 237.7 | 88.3 | 1458 | 2479 | 3.7 |
|  | 17 | 232.5 | 6.9 | 239.5 | 230.0 | 0.0 | 237.1 | 93.2 | 8396 | 11400 | 4.0 |
|  | 18 | 223.0 | 8.5 | 230.5 | 221.0 | 0.0 | 222.2 | 102.0 | 7153 | 9600 | 3.9 |
|  | 19 | 229.5 | 8.8 | 237.0 | 267.0 | 0.0 | 271.2 | 104.2 | 2343 | 6750 | 4.7 |
|  | 20 | 243.0 | 85.8 | 249.5 | 206.0 | 175.5 | 207.6 | 93.6 | 3668 | 6900 | 5.3 |
|  | 21 | 241.0 | 235.5 | 237.5 | 260.0 | 246.9 | 245.4 | 64.9 | 5710 | 8100 | 5.9 |
|  | 22 | 276.0 | 269.0 | 266.0 | 286.0 | 271.6 | 275.8 | 42.9 | 6591 | 9000 | 7.3 |
|  | 23 | 264.0 | 260.0 | 256.0 | 246.0 | 236.7 | 239.0 | 32.1 | 9215 | 12000 | 7.4 |
|  | 24 | 240.0 | 239.0 | 236.0 | 245.0 | 236.4 | 241.8 | 64.2 | 9770 | 12600 | 6.4 |
|  | 25 | 239.5 | 238.0 | 234.5 | 237.0 | 227.4 | 223.8 | 79.4 | 9686 | 12600 | 7.1 |
|  | 26 | 254.5 | 251.5 | 249.0 | 250.0 | 236.8 | 239.8 | 77.3 | 9770 | 12600 | 7.3 |
|  | 27 | 253.0 | 249.5 | 248.0 | 273.0 | 258.4 | 257.4 | 78.4 | 8186 | 10800 | 7.0 |
|  | 28 | 260.0 | 254.5 | 252.5 | 249.0 | 247.4 | 238.5 | 568.0 | 7500 | 10200 | 7.1 |
|  | 29 | 253.5 | 249.5 | 247.5 | 249.0 | 245.0 | 243.0 | 38.1 | 8680 | 10800 | 7.1 |
|  | 30 | 248.0 | 244.0 | 241.0 | 240.0 | 231.8 | 236.0 | 27.6 | 8443 | 10800 | 7.0 |
|  | 31 | 244.5 | 240.0 | 238.0 | 237.0 | 226.8 | 227.2 | 25.5 | 6965 | 8400 | 7.1 |

Data from Fort Nelson Gas Plant Powerhouse Morning Reports

Data from Fort Nelson Gas Plant Powerhouse Morning Reports

| Date | Steam Generation |  |  | Boiler feedwater |  |  | Excess 45 psi steam KLBH | Raw <br> Water <br> IGPH | Treated Water IGPH | Stack Emissions MMSCFD |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  | Boiler 5 <br> KLBH | Boiler 6 KLBH | Boiler 7 <br> KLBH | Boiler 5 KLBH | Boiler 6 <br> KLBH | Boiler 7 <br> KLBH |  |  |  |  |
| 1 | 249.0 | 249.5 | 249.0 | 237.0 | 235.4 | 233.6 | 23.5 | 7672 | 8900 | 6.9 |
| 2 | 244.5 | 245.0 | 246.5 | 236.0 | 225.8 | 236.3 | 27.2 | 2195 | 3125 | 7.0 |
| 3 | 271.0 | 273.0 | 272.0 | 292.0 | 292.8 | 297.3 | 37.1 | 5333 | 6200 | 7.3 |
| 4 | 296.0 | 298.0 | 298.0 | 287.0 | 284.5 | 292.7 | 60.4 | 7926 | 10500 | 7.4 |
| 5 | 293.0 | 297.0 | 296.0 | 288.0 | 291.0 | 290.9 | 86.1 | 8117 | 10500 | 7.3 |
| 6 | 293.0 | 296.0 | 296.0 | 289.0 | 287.5 | 290.7 | 104.5 | 4503 | 9600 | 7.3 |
| 7 | 293.0 | 296.0 | 295.0 | 289.0 | 286.0 | 292.6 | 113.0 |  |  | 7.4 |
| 8 | 293.0 | 296.0 | 296.0 | 287.0 | 285.9 | 296.7 | 113.2 | 2600 | 3400 | 7.4 |
| 9 | 294.0 | 297.0 | 296.0 | 294.0 | 294.3 | 296.8 | 105.2 | 7161 | 9000 | 7.6 |
| 10 | 296.0 | 299.0 | 299.0 | 292.0 | 290.3 | 293.2 | 108.8 | 2983 | 7200 | 7.2 |
| 11 | 296.0 | 299.0 | 299.0 | 288.0 | 290.1 | 290.8 | 110.7 | 4007 | 6731 | 7.4 |
| 12 | 293.0 | 296.0 | 295.0 | 289.0 | 289.3 | 297.8 | 109.7 |  |  | 7.4 |
| 13 | 294.0 | 296.0 | 296.0 | 284.0 | 285.6 | 294.1 | 114.9 | 4411 | 5638 | 7.1 |
| 14 | 291.0 | 295.0 | 294.0 | 284.0 | 289.4 | 289.1 | 126.7 | 8942 | 12063 | 7.0 |
| 15 | 288.0 | 293.0 | 291.0 | 286.0 | 287.9 | 292.3 | 126.9 | 422 | 513 | 7.0 |
| 16 | 291.0 | 296.0 | 295.0 | 282.0 | 288.4 | 289.8 | 125.8 | 3944 | 5250 | 7.1 |
| 17 | 294.0 | 300.0 | 298.0 | 285.0 | 290.5 | 296.1 | 120.5 | 6066 | 8700 | 7.1 |
| 18 | 293.0 | 298.0 | 297.0 | 284.0 | 294.7 | 292.0 | 115.2 | 5053 | 8700 | 7.1 |
| 19 | 298.0 | 302.0 | 301.0 | 301.0 | 300.3 | 303.7 | 105.2 | 5540 | 7688 | 7.2 |
| 20 | 299.0 | 301.0 | 299.0 | 221.0 | 217.6 | 212.2 | 94.5 | 3339 | 4800 | 5.1 |
| 21 | 219.0 | 220.5 | 216.0 | 280.0 | 277.8 | 278.2 | 66.8 | 4645 | 7288 | 6.0 |
| 22 | 283.0 | 285.0 | 283.0 | 273.0 | 270.2 | 273.0 | 43.5 | 8483 | 12413 | 7.1 |
| 23 | 275.0 | 276.0 | 274.0 | 276.0 | 272.4 | 277.0 | 56.3 | 2130 | 3000 | 7.0 |
| 24 | 274.0 | 276.0 | 275.0 | 269.0 | 269.2 | 268.0 | 65.1 | 0 | 0 | 6.9 |
| 25 | 263.0 | 265.0 | 264.0 | 246.0 | 242.4 | 244.2 | 61.9 | 0 | 0 | 6.7 |
| 26 | 261.0 | 241.5 | 260.0 | 242.0 | 239.6 | 241.5 | 51.8 | 5966 | 7800 | 6.8 |
| 27 | 246.5 | 248.0 | 246.0 | 237.0 | 239.5 | 242.3 | 37.1 | 5282 | 5900 | 7.0 |
| 28 | 256.0 | 259.0 | 256.0 | 245.0 | 237.7 | 244.9 | 35.1 | 1030 | 1200 | 6.8 |
| 29 | 246.5 | 248.0 | 246.0 | 243.0 | 246.5 | 247.0 | 25.3 | 6551 | 8719 | 7.1 |
| 30 | 249.0 | 250.0 | 249.0 | 240.0 | 244.7 | 245.1 | 19.8 | 2374 | 3200 | 6.9 |苞

Data from Fort Nelson Gas Plant Powerhouse Morning Reports

| Oct |  | Steam Generation |  |  | Boiler feedwater |  |  | Excess 45 <br> psi steam <br> KLBH | Raw <br> Water IGPH | Treated Water IGPH | Stack <br> Emissions <br> MMSCFD |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  | Date | Boiler 5 KLBH | Boiler 6 <br> KLBH | Boiler 7 <br> KLBH | Boiler 5 <br> KLBH | Boiler 6 <br> KLBH | Boiler 7 <br> KLBH |  |  |  |  |
|  | 1 | 247.5 | 249.0 | 247.5 | 236.0 | 233.5 | 236.8 | 16.3 | 5296 | 8450 | 6.7 |
|  | 2 | 241.0 | 242.5 | 241.0 | 232.0 | 231.7 | 237.3 | 14.4 | 9018 | 11700 | 6.9 |
|  | 3 | 245.0 | 246.0 | 244.5 | 246.0 | 243.9 | 246.1 | 12.8 | 7583 | 10800 | 7.1 |
|  | 4 | 250.0 | 250.0 | 249.0 | 243.0 | 241.8 | 245.2 | 11.5 | 400 | 625 | 6.5 |
|  | 5 | 247.0 | 248.0 | 246.5 | 245.0 | 245.3 | 250.0 | 10.7 | 5389 | 6800 | 6.3 |
|  | 6 | 259.0 | 259.0 | 258.0 | 244.0 | 245.3 | 251.2 | 10.8 | 9480 | 10200 | 6.2 |
|  | 7 | 248.0 | 249.5 | 246.5 | 231.0 | 230.1 | 233.4 | 12.5 | 3766 | 5250 | 12.5 |
|  | 8 | 243.0 | 241.5 | 239.5 | 237.0 | 226.3 | 233.7 | 13.6 | 7580 | 10200 | 7.1 |
|  | 9 | 248.0 | 236.0 | 234.0 |  |  |  | <<sos | §\%\%\% | \%\#\#\#\#\#\# | 7.1 |
|  | 10 | 252.0 | 239.0 | 241.0 |  |  |  |  |  |  | 7.1 |
|  | 11 | 252.0 | 241.0 | 241.0 |  |  |  |  |  |  | 7.1 |
|  | 12 | 333.0 | 320.0 | 320.0 |  |  |  |  |  |  | 8.1 |
|  | 13 | 288.0 | 279.0 | 277.0 | 308.0 | 288.5 | 296.1 | 43.6 |  |  | 7.8 |
|  | 14 | 308.0 | 298.0 | 297.0 | 315.0 | 291.0 | 298.7 | 46.2 | 8836 | 9900 | 7.6 |
|  | 15 | 228.0 | 333.0 | 328.0 | 186.0 | 337.5 | 3431.0 | 42.8 | 7161 | 7500 | 7.4 |
|  | 16 | 197.5 | 340.0 | 333.0 | 186.0 | 346.0 | 342.5 | 48.2 | 4266 | 3300 | 7.5 |
|  | 17 | 197.0 | 338.0 | 332.0 | 194.0 | 320.6 | 334.1 | 54.6 | 2121 | 0 | 7.4 |
|  | 18 | 201.0 | 355.0 | 347.0 | 190.0 | 373.9 | 375.9 | 76.5 | 0 | 10500 | 7.6 |
|  | 19 | 199.5 | 354.0 | 350.0 | 199.0 | 335.0 | 343.3 | 84.4 | 10600 | 12400 | 7.5 |
|  | 20 | 208.0 | 350.0 | 345.0 | 194.0 | 358.0 | 361.4 | 89.1 | 9898 | 11700 | 7.3 |
|  | 21 | 202.0 | 350.0 | 343.0 | 189.0 | 336.1 | 337.6 | 94.2 | 9280 | 9900 | 7.2 |
|  | 22 | 185.5 | 341.0 | 333.0 | 128.0 | 329.2 | 330.8 | 81.0 | 2450 | 2370 | 7.2 |
|  | 23 | 136.0 | 352.0 | 345.0 | 0.0 | 414.4 | 403.9 | 54.5 |  |  | 7.2 |
|  | 24 | 45.1 | 412.0 | 405.0 | 0.0 | 395.8 | 392.0 | 36.0 | 10464 | 11200 | 7.3 |
|  | 25 | 57.9 | 387.0 | 381.0 | 288.0 | 280.5 | 279.7 | 26.0 | 10231 | 11400 | 7.4 |
|  | 26 | 301.0 | 301.0 | 296.0 | 298.0 | 301.9 | 289.5 | 44.7 | 9783 | 10800 | 7.4 |
|  | 27 | 299.0 | 298.0 | 293.0 | 297.0 | 290.6 | 290.1 | 57.8 | 5400 | 5542 | 7.5 |
|  | 28 | 298.0 | 292.0 | 291.0 | 290.0 | 281.1 | 290.8 | 65.1 | 10079 | 10200 | 7.4 |
|  | 29 | 298.0 | 293.0 | 291.0 | 300.0 | 282.7 | 290.5 | 73.3 | 9881 | 11181 | 7.4 |
|  | 30 | 295.0 | 291.0 | 288.0 | 290.0 | 279.8 | 284.7 | 75.2 | 10311 | 11100 | 7.2 |
|  | 31 | 295.0 | 291.0 | 88.0 | 290.0 | 280.6 | 289.0 | 75.6 | 8881 | 9637 | 7.3 |

Data from Fort Nelson Gas Plant Powerhouse Morning Reports

|  | Date | Steam Generation |  |  | Boiler feedwater |  |  | Excess 45 psi steam KLBH | Raw <br> Water <br> IGPH | Treated Water IGPH | Stack Emissions MMSCFD |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  |  | Boiler 5 <br> KLBH | Boiler 6 <br> KLBH | Boiler 7 <br> KLBH | Boiler 5 KLBH | Boiler 6 <br> KLBH | Boiler 7 <br> KLBH |  |  |  |  |
| Nov | 1 | 291.0 | 287.0 | 284.0 | 288.0 | 281.2 | 283.5 | 71.8 | 6973 | 6587 | 7.2 |
|  | 2 | 282.0 | 276.0 | 274.0 | 297.0 | 289.0 | 294.9 | 66.0 | 0 | 0 | 6.9 |
|  | 3 | 302.0 | 295.0 | 293.0 | 11: m | \%\%\% 1 | MataM | 4 | 5799 | 6012 | 7.5 |
|  | 4 | 300.0 | 293.0 | 292.0 |  |  |  |  | 8330 | 9300 | 7.4 |
|  | 5 | 303.0 | 296.0 | 296.0 |  |  |  |  | 7804 | 9375 | 7.2 |
|  | 6 | 303.0 | 296.0 | 296.0 | 288.0 | 292.2 | 289.1 | 102.1 | 7198 | 9000 | 7.6 |
|  | 7 | 304.0 | 298.0 | 297.0 | 302.0 | 290.3 | 294.1 | 108.9 | 5526 | 5950 | 7.4 |
|  | 8 | 306.0 | 297.0 | 296.0 | 302.0 | 281.4 | 289.3 | 109.4 | 0 | 0 | 7.6 |
|  | 9 | 305.0 | 298.0 | 297.0 | 298.0 | 293.5 | 293.9 | 105.4 | 6547 | 7315 | 7.4 |
|  | 10 | 307.0 | 301.0 | 300.0 | 314.0 | 300.9 | 306.7 | 110.0 |  |  | 7.5 |
|  | 11 | 302.0 | 298.0 | 297.0 | 302.0 | 290.8 | 300.3 | 108.0 |  |  | 7.5 |
|  | 12 | 299.0 | 296.0 | 295.0 | 288.0 | 286.7 | 290.7 | 101.5 |  |  | 7.5 |
|  | 13 | 299.0 | 296.0 | 295.0 | 299.0 | 289.3 | 293.0 | 99.3 |  |  | 7.4 |
|  | 14 | 299.0 | 295.0 | 293.0 | 293.0 | 291.2 | 296.4 | 95.2 | 5605 | 6574 | 7.5 |
|  | 15 | 299.0 | 295.0 | 295.0 | 290.0 | 284.0 | 290.9 | 94.1 | 0 | 0 | 7.6 |
|  | 16 | 301.0 | 297.0 | 296.0 | 299.0 | 288.8 | 294.8 | 92.3 | 6538 | 7500 | 7.4 |
|  | 17 | 306.0 | 285.0 | 301.0 | 305.0 | 264.4 | 304.8 | 91.5 | 8486 | 11181 | 7.4 |
|  | 18 | 303.0 | 289.0 | 293.0 | 300.0 | 288.4 | 298.5 | 88.8 | 6064 | 8250 | 7.3 |
|  | 19 | 296.0 | 298.0 | 291.0 | 292.0 | 291.5 | 288.0 | 86.0 | 6673 | 8475 | 7.3 |
|  | 20 | 295.0 | 291.0 | 290.0 | 291.0 | 280.6 | 297.8 | 77.4 | 0 | 0 | 7.1 |
|  | 21 | 297.0 | 293.0 | 292.0 | 299.0 | 290.7 | 291.2 | 79.4 | 3142 | 4937 | 7.2 |
|  | 22 | 298.0 | 291.0 | 293.0 | 285.0 | 283.0 | 283.3 | 79.8 | 8269 | 9900 | 7.4 |
|  | 23 | 296.0 | 288.0 | 290.0 | गи:Min | \%10:M | : | זer: | 7236 | 9300 | 7.2 |
|  | 24 | 295.0 | 291.0 | 291.0 |  |  |  |  | 0 | 0 | 7.3 |
|  | 25 | 291.0 | 287.0 | 286.0 |  | . |  |  | 6042 | 5037 | 7.2 |
|  | 26 | 300.0 | 296.0 | 295.0 |  |  |  |  | 7359 | 9600 | 6.9 |
|  | 27 | 296.0 | 291.0 | 290.0 |  |  |  |  | 6365 | 9600 | 7.0 |
|  | 28 | 297.0 | 292.0 | 291.0 |  |  |  |  | 1954 | 2800 | 7.1 |
|  | 29 | 301.0 | 296.0 | 291.0 |  |  |  |  | 530 | 850 | 7.2 |
|  | 30 | 308.0 | 303.0 | 300.0 |  |  |  |  | 8435 | 11100 | 7.2 |

Data from Fort Nelson Gas Plant Powerhouse Morning Reports

| ceam Generatio |  |  |  | $$ |  |  | $\begin{gathered} \text { Excess } 45 \\ \text { psi steam } \\ \text { KLBH } \end{gathered}$ | RawWaterIGPH | $\begin{gathered} \substack{\text { Treaededed } \\ \text { Wacer } \\ \text { che }} \end{gathered}$ | $\begin{gathered} \text { Stack } \\ \text { Emissions } \\ \text { MMSCFD } \\ \hline \end{gathered}$ |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| Date | Boiler 5 | $\begin{gathered} \text { Boiler } 6 \\ \text { KLBH } \\ \hline \end{gathered}$ | $\underset{\substack{\text { Boile } 7 \\ \text { K183 }}}{\substack{\text { a }}}$ |  |  |  |  |  |  |  |
| 1 |  | 300.0 | 298.0 |  |  |  |  |  |  |  |
| ${ }^{2}$ | 3060 | 301.0 | 298.0 |  |  |  |  | 4940 | 7400 | ${ }^{7} 3$ |
| 3 | 306.0 | 301.0 | ${ }_{290}^{2980}$ |  |  |  |  | 0 | 0 | 72 |
| 5 | 304.0 | $\xrightarrow{299.0} \mathbf{\substack { 2 9 0 . 0 }}$ | ${ }_{2930}^{2900}$ |  |  |  |  | ${ }_{7774}^{4174}$ | ${ }_{1081}^{6090}$ | ${ }_{7}^{7.0}$ |
| 6 | 300.0 | 29.0 | 292.0 |  |  |  |  | 7889 | 10500 | 7.1 |
| 7 | 301.0 | 294.0 | 29.10 |  |  |  |  | 1717 | 1800 | ${ }^{73}$ |
| 8 | 33660 | ${ }^{300.0}$ | 296.0 |  |  |  |  | ${ }^{7489}$ | 9500 | ${ }^{7.3}$ |
|  | 3010 | $\underset{\substack{2950 \\ 2900}}{ }$ | ${ }_{20}^{29.0}$ |  |  |  |  | ${ }^{3480}$ | ${ }^{5230}$ | -6.8 |
| 11 | ${ }^{399} 0$ | ${ }_{2940}$ | ${ }_{290.0}^{29.0}$ |  |  |  |  | 7044 | 10200 | ${ }^{6.1}$ |
| 12 | 299.0 | 296.0 | 292.0 |  |  |  |  |  | 10200 | ${ }^{7} .0$ |
| 11 14 14 | 年迆, | - | $\underset{\substack{2950 \\ 2950}}{ }$ |  |  |  |  |  | 10200 6600 | 7.1 70 7 |
| 15 | 302.0 | 298.0 | ${ }_{293.0}^{293.0}$ |  |  |  |  | ${ }^{30} 0$ | ${ }^{600}$ | 7.0 |
| 16 | 298.0 | 294.0 | 291.0 |  |  |  |  | 4554 | 4325 | ${ }_{6} 6$ |
|  | 296.0 | 229.0 | ${ }^{288.0}$ |  |  |  |  |  | 10200 |  |
| ${ }_{19}^{18}$ | 2990.0 | ${ }_{294.0}^{29.0}$ | ${ }_{291.0}^{288.0}$ |  |  |  |  | ${ }_{\substack{1168 \\ 1190}}^{7}$ | 10200 <br> 1700 | ${ }_{7}^{6} 7$ |
| 20 | 3090 | 305.0 | 29.0 |  |  |  |  | ${ }^{1495}$ | 3000 | 7.5 |
| 21 | 301.0 | 296.0 | 280.0 |  |  |  |  | ${ }^{8596}$ | 11400 | ${ }_{6}^{6.8}$ |
| ${ }^{22}$ | 295.0 | ${ }^{2889.0}$ | ${ }^{274.0}$ |  |  |  |  | ${ }_{8821} 8$ | 12000 | 7.1 |
| ${ }_{24}^{23}$ | 5050 | 30.0) | ${ }_{288}^{290}$ |  |  |  |  | 807 | 11900 | ${ }_{6} .5$ |
| ${ }_{25}^{24}$ | 300.0 | ${ }_{299.0}$ | ${ }_{294.0}^{288.0}$ |  |  |  |  | 4168 | 6662 | 6.9 6.9 |
| 26 | 303.0 | 298.0 | 295.0 |  |  |  |  | ${ }_{8841}$ | 12300 | ${ }^{7.0}$ |
| 27 | 304.0 | 297.0 | 294.0 |  |  |  |  | 4679 | 6300 | 7.0 |
| ${ }_{29}^{28}$ | 306.0 | ${ }_{297.0}^{298.0}$ | ${ }_{2950}^{29.0}$ |  |  |  |  | ${ }_{8}^{1335}$ | 2100 <br> 11400 | ${ }_{6}^{6.9}$ |
| 30 | 304.0 | 297.0 | 29.0 |  |  |  |  | 6099 | 8750 | ${ }_{6} 6$ |

Data from Fort Nelson Gas Plant Powerhouse Morning Reports

Data from Fort Nelson Gas Plant Powerhouse Morning Reports

Data from Fort Nelson Gas Plant Process Morning Reports

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Data from Fort Nelson Gas Plant Process Morning Reports

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Data from Fort Nelson Gas Plant Process Morning Reports

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Data from Fort Nelson Gas Plant Process Morning Reports

| Date | Train C |  |  |  | Train D |  |  |  | Train E |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  | Carb unit KLBH | MEA KLBH | Total steam KLBH | Gas prod MMSCFD | Carb unit KLBH | MEA <br> KLBH | Total Steam KLBH | Gas prod MMSCFD | $\begin{aligned} & \text { DEA } \\ & \text { KLBH } \\ & \hline \end{aligned}$ | Gas prod MMSCFD |
| Overall trend: |  |  |  |  |  |  |  |  |  |  |
| Average | 105.59 | 31.95 | 137.54 | 100.51 | 107.02 | 33.08 | 140.09 | 105.01 | 127.66 | 89.89 |
| Maximum | 144 | 46 | 185 | 145 | 146 | 42 | 187 | 146 | 155 | 114 |
| Minimum | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 93 | 56 |
| Standard dev. | 18.77 | 5.68 | 23.58 | 19.19 | 14.47 | 4.05 | 17.36 | 14.36 | 10.18 | 7.29 |
| Coef. of var. | 0.18 | 0.18 | 0.17 | 0.19 | 0.14 | 0.12 | 0.12 | 0.14 | 0.08 | 0.08 |
| Count | 151 | 151 | 151 | 151 | 151 | 151 | 151 | 151 | 151 | 151 |

Data from Fort Nelson Gas Plant Process Morning Reports

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Data from Fort Nelson Gas Plant Process Morning Reports

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Data from Fort Nelson Gas Plant Process Morning Reports

|  | Train F |  | Train G |  | Train H |  | Total | Total |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  | DEA | Gas prod | DEA | Gas prod | DEA | Gas prod | Steam Use | Gas prod <br> Date |
| KLBH | MMSCFD | KLBH | MMSCFD | KLBH | MMSCFD | KLBH | MMSCFD |  |
| Overall trends |  |  |  |  |  |  |  |  |
| Average | 126.39 | 91.72 | 129.17 | 91.55 | 127.65 | 90.35 | 788.50 | 569.02 |
| Maximum | 157 | 116 | 151 | 109 | 157 | 106 | 880 | 625 |
| Minimum | 93 | 60 | 88 | 0 | 90 | 45 | 703 | 430 |
| Standard dev. | 10.28 | 6.77 | 10.06 | 10.62 | 8.80 | 6.20 | 28.39 | 23.27 |
| Coef. of var. | 0.08 | 0.07 | 0.08 | 0.12 | 0.07 | 0.07 | 0.04 | 0.04 |
| Count | 151 | 151 | 151 | 151 | 151 | 151 | 151 | 151 |

Data from Fort Nelson Gas Plant Computer monitoring system

|  |  |  | 0 psi st | ransfer |  | 450 to 45 | reakdo |  | d Water |  | Conti | Boiler | down | X-cess 45\# |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  |  | C/D | E/F | G/H | S/P | E/F | G/H | C/D | EFGH | S/P | Blr. 5 | Blr. 6 | Blr. 7 | steam-S/P |
|  | Date | KLBH | KLBH | KLBH | KLBH | KLBH | KLBH | IGPM | IGPM | IGPM | LBH | LBH | LBH | KLBH |
| 1994 | 1 |  |  |  |  |  |  |  |  |  |  |  |  |  |
| Dec | 2 | 169.9 | 124.2 | 100.9 | 56.1 | 2.1 | 11.8 | 50.8 | 0.0 | 132.6 | 779 | 2761 | 2549 | 59.1 |
|  | 3 | 167.1 | 123.9 | 99.1 | 57.6 | 2.5 | 14.2 | 50.9 | 0.0 | 150.4 | 2301 | 2478 | 2549 | 61.8 |
|  | 4 | 167.8 | 122.1 | 0.0 | 58.3 | 1.9 | 16.8 | 50.7 | 0.0 | 152.5 | 2265 | 2478 | . 2655 | 60.9 |
|  | 5 | 165.0 | 120.7 | 0.0 | 55.6 | 1.6 | 14.4 | 50.5 | 0.0 | 142.0 | 2761 | 2690 | 2195 | 57.3 |
|  | 6 | 167.8 | 119.3 | 99.1 | 58.3 | 0.3 | 13.0 | 51.4 | 0.0 | 138.9 | 3080 | 2619 | 1947 | 57.3 |
|  | 7 | 166.4 | 116.1 | 99.5 | 57.9 | 2.4 | 13.5 | 50.2 | 0.0 | 110.7 | 2973 | 2655 | 2124 | 62.7 |
|  | 8 | 172.7 | 117.5 | 99.1 | 54.8 | 0.6 | 28.6 | 51.7 | 0.0 | 137.8 | 2832 | 2655 | 2336 | 73.0 |
|  | 9 | 172.7 | 114.3 | 97.0 | 54.8 | 0.9 | 16.1 | 52.7 | 0.0 | 142.0 | . 2478 | 2301 | 2301 | 72.1 |
|  | 10 | 175.6 | 114.0 | 99.1 | 55.4 | 0.4 | 8.8 | 52.5 | 0.0 | 142.0 | 2832 | 1770 | 2478 | 73.0 |
|  | 11 | 177.7 | 112.2 | 0.0 | 55.0 | 0.8 | 9.7 | 54.1 | 0.0 | 137.8 | 2655 | 2478 | 2478 | 69.4 |
|  | 12 | 179.1 | 110.8 | 0.0 | 55.9 | 0.0 | 11.1 | 53.7 | 0.0 | 143.1 | 2814 | 2938 | 2655 | 73.0 |
|  | 13 | 183.4 | 115.4 | 0.0 | 54.5 | 0.5 | 15.3 | 53.7 | 0.0 | 148.3 | 2832 | 3009 | 2655 | 72.1 |
|  | 14 | 178.4 | 117.2 | 98.2 | 54.5 | 0.0 | 14.0 | 53.8 | 0.0 | 146.2 | 2230 | 2832 | 2442 | 72.1 |
|  | 15 | 165.7 | 127.1 | 104.4 | 54.8 | 0.2 | 12.1 | 55.2 | 0.0 | 149.3 | 885 | 2832 | 2301 | 71.2 |
|  | 16 | 153.6 | 123.9 | 102.7 | 56.4 | 0.1 | 17.5 | 56.3 | 0.0 | 148.3 | 1593 | 1274 | 2088 | 67.2 |
|  | 17 | 152.2 | 118.9 | 98.4 | 57.6 | 0.6 | 11.3 | 54.7 | 27.7 | 145.2 | 2124 | 2832 | 2832 | 73.4 |
|  | 18 | 154.3 | 117.9 | 95.6 | 57.0 | 0.4 | 7.8 | 54.7 | 0.0 | 144.1 | 2053 | 3858 | 2549 | 78.8 |
|  | 19 | 155.0 | 120.7 | 99.8 | 56.7 | 0.0 | 12.3 | 54.6 | 0.0 | 138.9 | 1947 | 3858 | 2655 | 83.3 |
|  | 20 | 164.2 | 121.4 | 102.7 | 56.7 | 0.4 | 7.8 | 57.3 | 0.0 | 147.2 | 1947 | 3363 | 2655 | 91.8 |
|  | 21 | 145.8 | 122.5 | 103.0 | 55.4 | 2.8 | 40.8 | 58.8 | 13.2 | 135.8 | 2230 | 3363 | 2655 | 133.4 |
|  | 22 | 118.9 | 122.5 | 103.7 | 49.5 | 2.2 | 29.1 | 54.6 | 28.9 | 129.5 | 3080 | 3009 | 2761 | 88.2 |
|  | 23 | 184.8 | 121.8 | 100.5 | 54.2 | 0.1 | 18.2 | 51.5 | 0.0 | 138.9 | 3469 | 3575 | 4319 | 83.7 |
|  | 24 | 177.0 | 117.2 | 93.8 | 54.8 | 0.2 | 9.9 | 51.8 | 0.0 | 144.1 | 3363 | 4425 | 4460 | 89.6 |
|  | 25 | 172.7 | 114.0 | 98.4 | 53.4 | 1.1 | 12.0 | 50.8 | 0.0 | 142.0 | 3469 | 3929 | 3681 | 99.0 |
|  | 26 | 167.8 | 116.1 | 98.4 | 53.4 | 0.3 | 15.8 | 50.1 | 0.0 | 142.0 | 3327 | 3965 | 4000 | 109.7 |
|  | 27 | 169.9 | 116.8 | 97.7 | 53.0 | 0.7 | 12.1 | 49.9 | 0.0 | 145.2 | 3150 | 3717 | '3965 | 115.5 |
|  | 28 | 171.3 | 117.5 | 98.8 | 53.4 | 0.4 | 13.2 | 50.8 | 0.0 | 147.2 | 2938 | 3752 | 3575 | 124.5 |
|  | 29 | 164.2 | 118.2 | 97.0 | 54.4 | 0.1 | 21.9 | 50.8 | 0.0 | 138.9 | 2407 | 3257 | 2478 | 125.8 |
|  | 30 | 166.4 | 118.6 | 97.7 | 55.3 | 1.4 | 12.5 | 46.9 | 0.0 | 139.9 | 1027 | 2619 | 2301 | 118.2 |
|  | 31 | 155.8 | 121.4 | 100.5 | 54.2 | 0.1 | 15.3 | 45.3 | 0.0 | 117.0 | 1416 | 2655 | 2478 | 110.2 |

Data from Fort Nelson Gas Plant Computer monitoring system

| Date | . 450 pris seam tansfered t |  |  |  | 450 to 45 pib traakdow |  | $\underset{\text { cop }}{\substack{\text { cid }}}{ }^{\text {Th}}$ | $\begin{aligned} & \text { Watarer } \\ & \text { EFFFH} \end{aligned}$ IGPM | $\underset{\substack{\text { SRP } \\ \text { IGPM }}}{ }$ | $\begin{gathered} \text { contin } \\ \text { cint } \\ \text { Lir } \\ \hline \end{gathered}$ |  | $\begin{gathered} \text { divn } \\ \text { Brr } \end{gathered}$ |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| $\begin{array}{ll}1995 \\ \text { Jan } & 1 \\ 2\end{array}$ | 174.2 | 118.2 | 94.5 | 54.8 | 0.87 | 17.9 | 42.7 | 0.00 | 144.1 | 1062 | 3080 | 3186 | 101.2 |
|  | 169.9 | 117.9 | 95.9 | 55.4 | 0.92 | 10.8 | 493 | 0.00 | 1420 | 1062 | 3327 | 3363 | 1025 |
| 5 | 172.0 | 117.9 | 100.2 | 59.3 | 0.01 | 19.4 | 48.6 | 0.00 | 150.4 | 1062 | 2832 | 2513 | 90.5 |
| 6 | 173.5 | 116.5 | 95.6 | 58.4 | 0.00 | 10.9 | 48.8 | 0.00 | 151.4 | 1062 | 3611 | 2407 | 87.3 |
|  | 165.7 | 1129 | 95.6 | 57.5 | 0.06 | 17.7 | 49.2 | 9.63 | 144.1 | 1593 | 3434 | 3346 | ${ }^{87.8}$ |
| 8 | .164.2 | 113.6 | 19.5 | 58.6 | 0.00 | 15.4 | 50.1 | 0.00 | 141.0 | 1558 | 3009 | 2301 | 87.3 |
| 9 | 163.5 | 113.6 | 0.0 | 49.1 | 0.09 | 21.9 | 49.3 | 0.00 | ${ }^{133.8}$ | ${ }^{1168}$ | ${ }^{2832}$ | ${ }^{2195}$ | 86.0 |
| 10 | 160.0 | 114.0 | 0.0 | ${ }^{53,0}$ | 0.12 | 10.9 | 49.3 | 0.00 | 138.9 | 1062 | 2832 | ${ }^{2124}$ | 90.9 1021 |
| 11 | 163.5 | 111.2 | 0.0 | 48.9 | 0.09 | 12.0 | 49.3 | 0.00 | ${ }_{13,14}^{14.0}$ | ${ }_{1062}^{1062}$ | 2655 | 2124 <br> 1286 <br> 18 | 20.1 |
| 12 | 162.1 | 112.6 | 0.0 | ${ }^{48,3}$ | 0.06 | 14.2 | 48.3 | 0.00 | ${ }_{1}^{13,9}$ | 1062 | ${ }_{2}^{2478}$ | ${ }^{1827}$ | ${ }^{8.4} 8$ |
| 13 | 16.9 | 113.3 | -0.0 | ${ }_{4}^{48,0}$ | 0.00 | -11.3 | 48.6 | 0.00 | 13.9 | 2832 | 2055 | 3009 <br> 2832 <br> 1 | ${ }_{1113.7}^{112.8}$ |
| 14 | 16.1 | 1112.9 | 34.7 623 | ${ }_{475}^{47.9}$ | -0.16 | 11.4 <br> 12.5 | ${ }_{425}^{42.1}$ | ${ }_{0} 0.00$ | 134.1 <br> 14.1 <br> 14.1 | 2790 2655 | ${ }_{2478}^{2796}$ | $\underset{\substack{2832 \\ 2832}}{ }$ | 111.9 |
| 16 |  | 115.0 | 101.6 | 48.3 | 0.05 | 19.8 | 425 | 0.00 | ${ }_{146.2}$ | ${ }^{2584}$ | 2655 | ${ }_{2938}$ |  |
| 17 | 172.0 | 113.3 | 96.3 | 48.8 | 0.01 | 32.4 | 42.8 | 0.00 | 15.2 | 2513 | 2655 | 3363 |  |
| 18 | 6 |  | 98.4 | 49.1 | 0.42 | 32.1 | 43.5 | 7.22 | ${ }^{148.3}$ | 2478 |  | 3504 | 1.2 |
| 19 | 169.9 | 121.1 | 99.1 | 49.2 | 0.04 | 40.1 | 43.5 | 0.00 | 137.8 | 2301 | 3292 | 3363 | 16.6 |
| 20 | 165.0 | 116.8 | 98.1 | 49.2 | 0.17 | 54.3 | 43.5 | 0.00 | 133.7 | 2195 | 4071 | 2832 | 13.7 |
| 21 | 170.6 | 118.9 | 97. | 49.5 | 0.07 | 38.0 | 43.5 | 0.00 | ${ }^{132.6}$ | 1487 | 3752 | 1841 | 6.0 |
| 22 | 173.5 | 119.6 | 99.8 | 49.1 | 0.43 | 10.2 | 42.8 | 7.22 | 138.9 | 1274 | 3150 | 1912 | 20.1 |
|  | 174.2 | 118.6 | 100.9 | 48.9 | 0.13 | 19.3 | 42.4 | 0.00 | ${ }^{13226}$ | 1841 | 3398 | 2249 | 4.9 |
|  | 177.0 | 117.2 | 99.8 | 49.2 | 0.22 | 17.2 | 43.4 | 0.00 | 129.5 | 2033 | 3292 | ${ }^{2372}$ | 94.5 |
| 25 | 172.0 | 118.6 | 99.1 | ${ }^{50.3}$ | 0.34 | 12.2 | 44.1 | 0.00 | ${ }_{12136}^{131.6}$ | ${ }_{2}^{2301}$ |  | 2301 | 91.8 |
| ${ }_{27}^{26}$ | 177.0 | 116.8 | 99.8 | 49.5 | 0.11 | 11.2 | 44.3 | 0.00 | ${ }_{13,5}^{133.8}$ | ${ }^{2336}$ | 2195 | 1912 | ${ }^{95.8}$ |
| 27 | 174.9 | 114.7 | -99.8 | 49.5 | 0.05 | 15.6 | 44.0 | 0.00 | ${ }^{133.6}$ | 2301 | 7487 | 2263 |  |
| 28 | 相 | 14.0 | 97.3 | 50.2 | 0.20 | 14.3 | 42.8 | 0.00 |  |  | 595 | 退 | \% |
|  | 7 | 2 | 9,96 | 488 | 0.00 | 13, | 4.4 | 0.00 | S358 | 2088 | S300 | 1953 |  |
| 31 | 1777 | 1115.0 | 94.5 | ${ }_{49.5}^{49.8}$ | ${ }_{0}^{0.00}$ | 11.9 11.8 | ${ }_{44.3}^{44.1}$ | ${ }_{0} 0.00$ | ${ }_{123.8}^{129.5}$ | 0 | ${ }_{3065}$ | 1064 2159 | ${ }_{918}^{96.1}$ |



## APPENDIX B

WATER BALANCE EQUATIONS AND ASSUMPTIONS

## GENERAL

## 8 psị steam balance:



Given: Amount of 8 psi steam condensed in Hot Lime Treater $=20 \%$ of raw water volume
Volume of 8 psi steam generated by deaerators $=50.2 \mathrm{KLBH}$ or $547 \mathrm{~m}^{3} /$ day
Treated water losses from plant (excluding 8 psi steam) $=44.1 \mathrm{KLBH}$ or $480 \mathrm{~m}^{3} /$ day
Hot Lime Treater blowdown $=1.5 \mathrm{KLBH}$ or $16.3 \mathrm{~m}^{3} /$ day
$\mathrm{x}=$ volume of 8 psi steam condensing in teh Hot Lime Treater
Then:

$$
\begin{aligned}
\text { Raw water flow } & =44.1+(50.2-\mathrm{x}) \\
\text { Venting } & =50.2-\mathrm{x} \\
\text { Treated water flow } & =(44.1+(50.2-\mathrm{x})) * 1.2-1.5
\end{aligned}
$$

Since,
Total incoming flow $=$ Total outgoing flow

$$
\begin{aligned}
(44.1+50.2-x)+50.2 & =(50.2-x)+((44.1+50.2-x) * 1.2-1.5)+1.5 \\
94.3 & =(94.3-x) * 1.2 \\
1.2 x & =18.9 \\
x & =15.7
\end{aligned}
$$

So,

$$
\begin{aligned}
& 8 \text { psi condensate }=15.7 \mathrm{KLBH} \text { or } \mathbf{1 7 1} \mathbf{~ m}^{3} / \text { day } \\
& \text { Raw water flow }=78.9 \mathrm{KLBH} \text { or } \mathbf{8 5 9} \mathbf{~ m}{ }^{3} / \text { day } \\
& \text { Treated water flow }=93.2 \mathrm{KLBH} \text { or } 1.01 \mathrm{e} 3 \mathrm{~m}^{3} / \text { day } \\
& \text { Venting } 8 \text { psi steam }=34.5 \mathrm{KLBH} \text { or } \mathbf{3 7 6} \mathbf{~ m}^{3 /} \text { day }
\end{aligned}
$$

## \#3 Hot Lime Treater blowdown:

The Water Treatment Manual (WTM) indicates that the blowdown rate $=8-10 \%$ of the recirculating sludge. By design, the sludge should be recycled at a rate of 30 gpm. Blowdown from the treater $=10 \%$ of $30 \mathrm{gpm}=3 \mathrm{gpm}$ or $\mathbf{1 6 . 3} \mathbf{~ m}{ }^{3} /$ day .

## Softener regeneration brine:

Regenerating a zeolite softener requires the following water volumes (from WTM):

| Operation | Flow rates <br> (gpm) | Time <br> (minutes) | Total flow <br> (gallons) |
| :--- | :---: | :---: | :---: |
| Backwash | 495 | 10 | 4950 |
| Brine injection | 26 | 12 | 312 |
| - dilution water | 36.5 |  | 438 |
| Brine displacement | 36.5 | 20 | 730 |
| Fast Rinse | 200 | 30 | 6000 |
| Service Rinse | 200 | 15 | $\underline{3000}$ |
|  |  | Grand total: | 15430 |

These waters are discharged to the lime pond and the \#3 Hot Lime Treater as detailed below:
Given: Capacity of each softener $=20000$ grains $/ \mathrm{ft}^{3}$
Volume of each softener $=132 \mathrm{ft}^{3}$
1 grain is = equivalent to 17.1 ppm of hardness*US gal
Inlet water contains $40.5 \mathrm{mg} / \mathrm{L}$ of $\mathrm{Ca} \& 1.2 \mathrm{mg} / \mathrm{L}$ of Mg
Flow through all three filters $=9.31 \mathrm{e} 3 \mathrm{Igal} / \mathrm{h}$.
Each unit can filter:
$=\left(20000\right.$ grains $\left./ \mathrm{ft}^{3} \times 132 \mathrm{ft}^{3}\right) \times 17.1 \mathrm{ppm}^{*} \mathrm{US}$ gal/grain $\times 106 \mathrm{mg} / \mathrm{L}$ as $\mathrm{CaCO}_{3}$
$=4.25 \mathrm{e} 5 \mathrm{USgal}$ of water before regeneration
Time until softener needs regenerating:

$$
\begin{aligned}
& =4.25 \mathrm{e} 5 \mathrm{USgal} /\left(9.31 \mathrm{e} 3 \mathrm{Igal} / \mathrm{h} \mathrm{x} 1.2 \mathrm{USgal} / \mathrm{Igal} \times 1 / 2^{*}\right) \\
& =76.1 \text { hours } \\
& (* \text { assuming only } 2 \text { filters running at any given time) }
\end{aligned}
$$

Assuming:

- backwash water returned to \#3 treater
- all other regeneration water (listed above) discharge to lime pond

Amount of water going back to \#3 treater:

$$
\begin{aligned}
& =4950 \mathrm{USgal} / \text { regeneration } / \text { exchanger } \\
& =65.0 \mathrm{USgal} / \mathrm{h} / \text { exchanger } \\
& =195 \mathrm{USgal} / \mathrm{h} \text { or } \mathbf{1 7 . 7} \mathbf{~ m}^{3} / \mathrm{day}
\end{aligned}
$$

Amount of water to lime pond:

$$
\begin{aligned}
& =10480 \mathrm{USgal} / \text { regeneration/exchanger } \\
& =138 \mathrm{USgal} / \mathrm{h} / \text { exchanger } \\
& =413 \mathrm{USgal} / \mathrm{h} \text { or } 37.6 \mathbf{~ m}^{3} / \mathbf{d a y}
\end{aligned}
$$

But using a sodium and chloride balance on the lime pond:

| brine concentration $=$ | Chloride | Sodium |  |
| ---: | :---: | :---: | :--- |
| 5.49 e 3 | 3.56 e 3 | $\mathrm{mg} / \mathrm{L}($ see Appendix D) |  |
| brine mass needed balance $=$ | $\underline{2.15 \mathrm{e} 8}$ | $\underline{1.39 \mathrm{e} 8}$ | $\mathrm{mg} / \mathrm{day}$ |
| require flow $=$ | 3.91 e 4 | 3.91 e 4 | $\mathrm{~L} /$ day |
|  |  | $\mathbf{3 9 . 1}$ | $\mathbf{m}^{3 / \mathbf{d a y}}$ |

## Domestic filter backwash:

It was assumed that the equivalent of 3 volumes of treated water are used to backwash each of the 2 domestic filters. They measure $30^{\prime \prime}$ (ID) by $60^{\prime \prime}$ (height), and are cleaned every 24 hours.

## Turbines:

It was assumed that all of the turbines within the FNGP were loosing $0.1 \%$ of their incoming steam through venting.

## Steam tracing:

It was assumed that $1 \%$ of the steam used for steam tracing escaped through leaks in the piping.

## Treated water cleaning:

Assumed that a total of 2 KLBH or $21.8 \mathrm{~m}^{3} /$ day of treated water were used for cleaning vessels, floors ... in the 6 process trains.

## Water lost in sweet gas:

Assumed a loss of 4 lb of water $/ \mathrm{mmscf}$ of processed gas. The average gas production for January 1995 was 201.4 and $367.8 \mathrm{mmscf} /$ day for the C/D \& E/F/G/H trains, respectively.

## Flare water:

Assumed a loss of 80 lb of water/mmscf of processed gas. The average gas production for January 1995 was 201.4 and $367.8 \mathrm{mmscf} /$ day for the $\mathrm{C} / \mathrm{D} \& \mathrm{E} / \mathrm{F} / \mathrm{G} / \mathrm{H}$ trains, respectively.

## Water vapour lost with the acid gas:

Given: acid gas mixture contains $\mathrm{H}_{2} \mathrm{~S}, \mathrm{CO}_{2}$ and water

$$
\% \text { of } \mathrm{H}_{2} \mathrm{~S} \text { in acid gas from process }=11.2 \%
$$

Gas law:

$$
\mathrm{P}_{\mathrm{AG}} \times \mathrm{V}=\left(\mathrm{n}_{\mathrm{H} 2 \mathrm{~S}}+\mathrm{n}_{\mathrm{CO} 2}+\mathrm{n}_{\mathrm{H} 2 \mathrm{O}}\right) \times \mathrm{R} \times \mathrm{T}
$$

where: $\quad \mathrm{P}_{\mathrm{AG}}=$ acid gas pressure (atm)
$\mathrm{V}=\operatorname{acid}$ gas volume ( L )
$\mathrm{n}=$ moles of gas
$\mathrm{R}=$ gas constant
$=0.08206(\mathrm{~L} * \mathrm{~atm}) /(\mathrm{mol} * \mathrm{~K})$
$\mathrm{T}=$ temperature (Kelvins)

So

$$
\mathrm{n}_{\mathrm{H} 2 \mathrm{O}}=\left(\mathrm{P}_{\mathrm{AG}} \mathrm{~V} / \mathrm{RT}\right)-\left(\mathrm{n}_{\mathrm{H} 2 \mathrm{~S}}+\mathrm{n}_{\mathrm{CO} 2}\right)
$$

But

$$
\mathrm{n}_{\mathrm{H} 2 \mathrm{~S}} / \mathrm{n}_{\mathrm{AG}}=\% \mathrm{H}_{2} \mathrm{~S} \text { content }
$$

and

$$
\mathrm{n}_{\mathrm{AG}}=\mathrm{P}_{\mathrm{AG}} \mathrm{~V} / \mathrm{RT}
$$

So

$$
\mathrm{n}_{\mathrm{H} 2 \mathrm{~S}}=\left(\% \mathrm{H}_{2} \mathrm{~S}\right)\left(\mathrm{P}_{\mathrm{AG}} \mathrm{~V} / \mathrm{RT}\right)
$$

Since $V$ is measured in scf - actual volume has been transformed into equivalent volume at 1 atm. of pressure and a temperature $=293$ kelvins.

$$
\begin{aligned}
\mathrm{n}_{\mathrm{H} 2 \mathrm{~S}} & =11.2 \% \times\left(1 \mathrm{~atm} . \times 113.2 \mathrm{mmscfd} \times 28.3168 \mathrm{ft}^{3} / \mathrm{L}\right) / \\
& \left(0.08206 \mathrm{~L} * \mathrm{~atm} / \mathrm{mol}^{*} \mathrm{~K} \times 293 \mathrm{~K}\right) \\
& =1.47 \mathrm{e} 7 \mathrm{~mol} / \mathrm{day}
\end{aligned}
$$

Assuming:

- $\mathrm{CO}_{2}$ content in raw gas $=13.5 \%$ (as indicated in Sulphur Plant Operating Manual)
- All of the $\mathrm{CO}_{2}$ in the raw gas ends up on the acid gas stream

By similar steps:

$$
\begin{aligned}
\mathrm{n}_{\mathrm{CO} 2} & =\% \mathrm{CO}_{2} \times\left(\mathrm{P}_{\mathrm{RG}} \mathrm{~V}_{\mathrm{RG}} / \mathrm{RT}\right) \\
& =13.5 \% \times\left(1 \mathrm{~atm} . \times 682.5 \mathrm{mmscfd} \times 28.3168 \mathrm{ft}^{3} / \mathrm{L}\right) / \\
& (0.08206 \mathrm{~L} * \mathrm{~atm} / \mathrm{mol} * \mathrm{~K} \times 293 \mathrm{~K}) \\
& =1.09 \mathrm{e} 8 \mathrm{~mol} / \mathrm{d}
\end{aligned}
$$

Now

$$
\begin{aligned}
\mathrm{n}_{\mathrm{H} 2 \mathrm{O}} & =\left[\left(1 \mathrm{~atm} . \times 113.2 \mathrm{mmscfd} \times 28.3168 \mathrm{ft}^{3} / \mathrm{L}\right) /\left(0.08206 \mathrm{~L}^{*} \mathrm{~atm} / \mathrm{mol}^{*} \mathrm{~K} \mathrm{x}\right.\right. \\
& 293 \mathrm{~K})]-(1.47 \mathrm{e} 7 \mathrm{~mol} / \mathrm{d}+1.09 \mathrm{e} 8 \mathrm{~mol} / \mathrm{d}) \\
& =9.87 \mathrm{e} 6 \mathrm{~mol} / \mathrm{d}
\end{aligned}
$$

Mass of water:

$$
\begin{aligned}
& =\mathrm{n}_{\mathrm{H} 2 \mathrm{O}} \times 18 \mathrm{~g} / \mathrm{mol} \times \mathrm{kg} / 1000 \mathrm{~g} \times 2.2046 \mathrm{lb} / \mathrm{kg} \times \text { day } / 24 \mathrm{~h} \\
& =16300 \mathrm{lb} / \mathrm{h} \text { or } \mathbf{1 7 8} \mathbf{~ m}^{3} / \text { day }
\end{aligned}
$$

## Water formed within the Sulphur Plant:

Given: mol of incoming $\mathrm{H}_{2} \mathrm{~S}=1.49 \mathrm{e} 7 \mathrm{~mol} / \mathrm{d}$
(see "water vapour in acid gas")
average mass of sulphur leaving stack $=15.4$ tonnes $/ \mathrm{d}$
Mol of sulphur leaving the stack:

$$
\begin{aligned}
& =\text { mass } S \times 1000 \mathrm{~kg} / \text { tonne } \times 1000 \mathrm{~g} / \mathrm{kg} \times 1 / 32 \mathrm{~mol} / \mathrm{g} \\
& =15.4 \times 1000 \times 1000 \times 1 / 32 \\
& =4.81 \mathrm{e} 5 \mathrm{~mol} / \mathrm{d}
\end{aligned}
$$

Mol of $\mathrm{H}_{2} \mathrm{~S}$ converted into elemental sulphur:

$$
\begin{aligned}
& =\text { incoming mol }- \text { outgoing mol } \\
& =1.49 \mathrm{e} 7 \mathrm{~mol} / \mathrm{d}-4.81 \mathrm{e} 5 \mathrm{~mol} / \mathrm{d} \\
& =1.45 \mathrm{e} 7 \mathrm{~mol} / \mathrm{d}
\end{aligned}
$$

Conversion reaction:

$$
2 \mathrm{H}_{2} \mathrm{~S}+\mathrm{SO}_{2}=3 \mathrm{~S}+2 \mathrm{H}_{2} \mathrm{O}
$$

So:

$$
\begin{aligned}
\mathrm{mol} \mathrm{H}_{2} \mathrm{O} & =\mathrm{mol} \mathrm{H}_{2} \mathrm{~S} \\
& =1.45 \mathrm{e} 7 \mathrm{~mol} / \mathrm{d}
\end{aligned}
$$

Mass of water formed:
$=1.45 \mathrm{e} 7 \mathrm{~mol} / \mathrm{d} \mathrm{x} 18 \mathrm{~g} / \mathrm{mol} \mathrm{x} \mathrm{kg} / 1000 \mathrm{~g} \times 2.2046$
lbs/kg x klb/1000lb
$=23.9 \mathrm{KLBH}$ or $\mathbf{2 6 0} \mathrm{m}^{3} /$ day

## UNIT SPECIFIC

## Powerhouse

## Treated water to process:

This value is the sum of all of the treated water demands from the 6 process trains.

## Process

Cleaning solution make-up:
Sum of water lost in sweet gas, acid gas and to the flare pits

## Sulphur Plant

Individual blowdowns:
Values taken off of a schematic diagram

## Boiler feedwater distribution:

The volumes of BFW going to the individual condensers and the reaction furnace were calculated as follows:

Given: $\quad$ Total volume of BFW to Sulphur Plant $=160 \mathrm{KLBH}$ or $1.74 \mathrm{e} 3 \mathrm{~m}^{3} /$ day 5 condensers produce 37.2 KLBH or $378 \mathrm{~m}^{3} /$ day of condensate
15 psi condensers produce 34.7 KLBH or $378 \mathrm{~m}^{3} /$ day of condensate
Reaction furnace blowdown $=2.0 \mathrm{KLBH}$ or $21.8 \mathrm{~m}^{3} /$ day
Blowdown from \#1 condensers $=1.5 \mathrm{KLBH}$ or $16.3 \mathrm{~m}^{3} /$ day
Blowdown from \#2 condensers $=3.0 \mathrm{KLBH}$ or $32.7 \mathrm{~m}^{3} /$ day
150 psi steam to Petrosul $=1.8 \mathrm{KLBH}$ or $19.7 \mathrm{~m}^{3} /$ day
45 psi steam from \#2 condenser $=41.0 \mathrm{KLBH}$ or $446 \mathrm{~m}^{3} /$ day
BFW to reaction furnace:

$$
\begin{aligned}
& =5 \text { psi condensate }+ \text { Petrosul }+ \text { blowdown }+ \text { venting } \\
& =37.2+2.0+1.8 \\
& =41.0 \mathrm{KLBH} \text { or } \mathbf{4 4 6} \mathbf{~ m}^{3} / \mathbf{d a y}
\end{aligned}
$$

BFW to \#2 condensers:

$$
\begin{aligned}
& =15 \mathrm{psi} \text { condensate }+ \text { blowdown }+45 \mathrm{psi} \text { steam } \\
& =34.7+3.0+41.0 \\
& =78.7 \mathrm{KLBH} \text { or } \mathbf{8 5 7} \mathbf{~ m}^{3} / \mathbf{d a y}
\end{aligned}
$$

BFW to \#1 condensers:

$$
=160-\text { (B.F.W. to \#2 condensers }+ \text { B.F.W. to reaction }
$$

furnace)

$$
=160-(78.7+41.0)
$$

## $=40.3 \mathrm{KLBH}$ or $\mathbf{4 3 8} \mathbf{~ m}^{3} / \mathbf{d a y}$

45 psi steam from \#1 condensers:

$$
\begin{aligned}
& =\text { B.F.W. to \#1 condensers - blowdown } \\
& =40.3-1.5 \\
& =38.8 \mathrm{KLBH} \text { or } \mathbf{4 2 2} \mathbf{~ m}^{3} / \mathbf{d a y}
\end{aligned}
$$

## Effluent Plant:

Dirty raw water:
Assumed that 137 of the 159 incoming units was dirty raw water used either for cleaning or within the domestic system.
Sour water:
Assumed that the amount of sour water traveling to the effluent plant is negligible

## APPENDIX C

## WATER CHEMISTRY DATA

| Parameter | Date of sample collection |  |  |  |  |  |  |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  | March | May | June * | July | Avg. | April | July | Avg. |
| General |  |  |  |  |  |  |  |  |
| pH | 7.8 | 7.3 | 7.6 | 7.7 | 7.6 | 7.6 | 7.8 | 7.7 |
| Conductivity | 571 | 300 | 430 | 358 | 415 | 393 | 405 | 399 |
| Alkalinity |  |  |  |  |  |  |  |  |
| P | - | - | - | - |  | - | - |  |
| M | 141 | 76.0 | 114 | 106 |  | 93.0 | 155 |  |
| Total | 141 | 76.0 | 114 | 106 | 109 | 93.0 | 155 | 116 |
| Solids |  |  |  |  |  |  |  |  |
| Total | 380 | 266 | 5365 | 734 | 1686 | 659 | 442 | 880 |
| Suspended | 1.25 | 174 | 4316 | 512 | 1251 | 395 | 206 | 591 |
| Dissolved | 379 | 92.2 | 1049 | 222 | 435 | 265 | 236 | 289 |
| Carbon content |  |  |  |  |  |  |  |  |
| Total | 35.9 | 30.9 | 32.8 | 34.4 | 33.5 | 28.4 | 63.7 | 40.0 |
| Organic | 0.5 | 12.1 | 3.8 | 8.3 | 6.2 | 5.5 | 25.2 | 11.3 |
| Inorganic | 35.4 | 18.8 | 29.0 | 26.1 | 27.3 | 22.9 | 38.5 | 28.7 |
| Metals |  |  |  |  |  |  |  |  |
| Calcium | 73.8 | 38.2 | 114 | 54.9 | 70.1 | 45.0 | 61.1 | 57.8 |
| Magnesium | 23.6 | 8.2 | 29.5 | 13.6 | 18.7 | 10.8 | 14.2 | 14.4 |
| Sodium | 11.4 | 7.6 | 6.1 | 5.1 | 7.6 | 9.0 | 5.4 | 6.8 |
| Iron | 0.13 | 2.46 | 28.0 | 5.03 | 8.90 | 6.36 | 3.14 | 5.86 |
| Inorganics |  |  |  |  |  |  |  |  |
| Phosphates | 0.05 | 0.05 | 0.07 | 0.08 | 0.06 | 0.05 | 0.05 | 0.06 |
| Chlorides | 2.2 | 1.7 | 1.0 | 1.3 | 1.6 | 1.6 | 2.4 | 1.7 |
| Sulphates | 129 | 43.6 | 82.8 | 49.0 | 76.0 | 62.9 | 21.0 | 52.2 |
| Silica | 4.1 | 2.8 | 3.8 | 3.6 | 3.6 | 0.6 | 7.5 | 3.8 |
| Units : $\mathrm{pH}=\mathrm{pH}$ units |  |  |  |  | Phosphates $=\mathrm{mg}$ of $\mathrm{P} / \mathrm{L}$ |  |  |  |
| Conductivity $=\mu \mathrm{S} / \mathrm{cm}$ @ 25 C |  |  |  |  | Chlorides $=\mathrm{mg}$ of $\mathrm{Cl} / \mathrm{L}$ |  |  |  |
| Alkalinity $=\mathrm{mg} / \mathrm{L}$ as CaCO 3 |  |  |  |  | Sulphates $=\mathrm{mg}$ of SO4/L |  |  |  |
|  | Solids Carbon c Metals | tent | - mg/L |  | Silica $=$ | of SiO |  |  |

## Raw water header

Sample dropped from final data set
Date of sample collection

| Parameter | 3-May | 8-May | 10-May | 15-May | 17-May | 24-May | 12 ll | Avg. |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| General |  |  |  |  |  |  |  |  |
| pH | 8.0 | 7.8 | 8.0 | 7.9 | 7.9 | 8.0 | 76 | 7.9 |
| Conductivity | 571 | 666 | 658 | 476 | 489 | 518 | 792 | 563 |

Alkalinity

| P | - | - | - |  | - |  |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| M | 202 | 201 | 200 | 169 | 158 | 155 | 139 |
| Total | 202 | 201 | 200 | 169 | 158 | 155 | 139 |

Solids

| Total | 452 | 419 | 241 | 564 | 345 | 354 | 709 | 396 |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| Suspended | 9.88 | 4.02 | 4.67 | 4.60 | 14.33 | 0.56 | 1.91 | 6.34 |
| Dissolved | 443 | 414 | 237 | 559 | 331 | 353 | 707 | 389 |

Carbon content
Tota
$60.1 \quad 61.3 \quad 62$
$62.0 \quad 4$
46.4
72.0
52.6

170
59.1

| Organic | 9.2 | 9.1 | 9.8 | 8.3 | 9.5 | 9.8 | 135 | 9.3 |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |


| Inorganic | 50.9 | 52.2 | 52.2 | 38.1 | 62.5 | 42.8 | $\cong$ | 350 | $\mathbf{4 9 . 8}$ |
| :--- | :--- | :--- | :--- | :--- | :--- | :--- | :--- | :--- | :--- |

Metals

| Calcium | 97.3 | 90.3 | 92.2 | 70.9 | 68.4 | 68.4 | 66.7 | 81.3 |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| Magnesium | 21.4 | 21.0 | 20.0 | 16.7 | 15.1 | 15.5 | 16.9 | 18.3 |
| Sodium | 17.7 | 12.1 | 11.6 | 10.6 | 11.2 | 11.8 | 37.4 | 12.5 |
| Iron | 0.39 | 0.40 | 0.38 | 0.31 | 0.67 | 0.21 | 0.42 | 0.39 |

Inorganics

| Phosphates | 0.05 | 0.05 | 0.05 | 0.05 | 0.05 | 0.05 | 0.11 | 0.05 |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| Chlorides | 3.5 | 2.1 | 2.0 | 3.6 | 3.9 | 2.1 | 82.7 | 2.9 |
| Sulphates | 81.6 | 100.3 | 100.8 | 81.1 | 77.5 | 79.3 | 85.9 | 86.8 |
| Silica | 5.3 | 5.3 | 5.9 | 5.0 | 5.3 | 4.5 | 3.1 | 5.2 |

Units: $\mathrm{pH}=\mathrm{pH}$ units
Conductivity $=\mu \mathrm{S} / \mathrm{cm} @ 25 \mathrm{C}$
Alkalinity $=\mathrm{mg} / \mathrm{L}$ as CaCO 3
Solids
$\begin{array}{lll}\text { Carbon content } \\ \text { Metals } & \\ & \end{array}$
Phosphates $=m g$ of $\mathrm{P} / \mathrm{L}$
Chlorides $=\mathrm{mg}$ of $\mathrm{Cl} / \mathrm{L}$
Sulphates $=\mathrm{mg}$ of SO4/L
Silica $=m g$ of $\mathrm{SiO} 2 / \mathrm{L}$

## Reservoir water

| Sample dropped from final data set |  |  |  |  |  |  |  |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| Parameter | 3 May | 8-May | 10-May | 15-May | 17-May | 24-May | 12-Jul | Avg. |
| General |  |  |  |  |  |  |  |  |
| pH | 7.5 | No | No | No | No | 8.2 | 7.9 | 8.1 |
| Conductivity | 896 | Sample | Sample | Sample | Sample | 504 | 523 | 514 |
| Alkalinity |  |  |  |  |  |  |  |  |
|  | » |  |  |  |  | - | - |  |
| M | 370 |  |  |  |  | 151 | 136 |  |
| Total | 370 |  |  |  |  | 151 | 136 | 144 |
| Solids |  |  |  |  |  |  |  |  |
| Total | 92.5 |  |  |  |  | 331 | 297 | 314 |
| Suspended | 1.27 |  |  |  |  | 4.79 | 1.85 | 3.32 |
| Dissolved: | 91.2 |  |  |  |  | 327 | 296 | 311 |
| Carbon content |  |  |  |  |  |  |  |  |
| Total | 10.6 |  |  |  |  | 51.3 | 45.1 | 48.2 |
| Organic | 2.8 |  |  |  |  | 10.4 | 11.2 | 10.8 |
| Inorganic | 78 |  |  |  |  | 40.9 | 33.9 | 37.4 |
| Metals |  |  |  |  |  |  |  |  |
| Calcium: | 15.4. |  |  |  |  | 65.2 | 62.6 | 63.9 |
| Magnesium) | 28 |  |  |  |  | 15.2 | 16.3 | 15.8 |
| Sodium | 2.6 |  |  |  |  | 12.0 | 11.1 | 11.5 |
| Iron» | 010 |  |  |  |  | 0.14 | 0.20 | 0.17 |
| Inorganics |  |  |  |  |  |  |  |  |
| Phosphates) | 005 |  |  |  |  | 0.05 | 0.08 | 0.06 |
| Chlorides | 0.5 |  |  |  |  | 2.3 | 1.3 | 1.8 |
| Sulphates | 13.2 |  |  |  |  | 87.8 | 69.8 | 78.8 |
| Silica | 1.0 |  |  |  |  | 4.5 | 3.6 | 4.0 |

Units: $\mathrm{pH}=\mathrm{pH}$ units
Conductivity $=\mu \mathrm{S} / \mathrm{cm} @ 25 \mathrm{C}$
Alkalinity $=\mathrm{mg} / \mathrm{L}$ as CaCO 3
Solids
Carbon content $-\mathrm{mg} / \mathrm{L}$ Metals

Phosphates $=\mathrm{mg}$ of $\mathrm{P} / \mathrm{L}$
Chlorides $=\mathrm{mg}$ of $\mathrm{Cl} / \mathrm{L}$
Sulphates $=\mathrm{mg}$ of SO4/L
Silica $=\mathrm{mg}$ of $\mathrm{SiO} 2 / \mathrm{L}$

## Treater system outlet

| Parameter | Date of sample collection |  |  |  |  | 24-May | Avg. |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  | 3-May | 8-May | 10-May | 15-May | 17-May |  |  |
| General |  |  |  |  |  |  |  |
| pH | 9.4 | 10.1 | 9.8 | 10.4 | No | 11.0 | 9.9 |
| Conductivity | 229 | 365 | 324 | 327 | Sample | 1124 | 311 |
| Alkalinity |  |  |  |  |  |  |  |
| P | 7.0 | 16.5 | 10.5 | 23.5 |  | 186 |  |
| M | 14.5 | 7.5 | 20.5 | 11.0 |  | 140 |  |
| Total | 21.5 | 24.0 | 31.0 | 34.5 |  | 326 | 27.8 |
| Solids |  |  |  |  |  |  |  |
| Total | 222 | 201 | 54 | 420 |  | 541 | 288 |
| Suspended | 23.4 | 3.14 | 1.52 | 2.19 |  | 14.5 | 8.95 |
| Dissolved | 199 | 198 | 52 | 418 |  | 527 | 279 |
| Carbon content |  |  |  |  |  |  |  |
| Total | 9.8 | 8.8 | 11.4 | 24.4. |  | 52.8 | 20.7 |
| Organic | 5.8 | 5.2 | 5.4 | 5.3 |  | 8.1 | 6.1 |
| Inorganic | 4.0 | 3.6 | 6.0 | ٪. |  | 44.7 | 14.6 |
| Metals |  |  |  |  |  |  |  |
| Calcium | 1.5 | 0.2 | 0.1 | 0.1 |  | 5.0 | 1.4 |
| Magnesium | 1.2 | 0.3 | 0.1 | 0.3 |  | 0.5 | 0.5 |
| Sodium | 45.2 | 51.5 | 54.8 | 50.6 |  | 180 . | 50.5 |
| Iron | 0.04 | 1.78 | 0.35 | 0.86 |  | 0.04 | 0.61 |
| Inorganics |  |  |  |  |  |  |  |
| Phosphates | 0.05 | 0.11 | 0.05 | 0.06 |  | 0.06 | 0.07 |
| Chlorides | 1.7 | 2.1 | 2.1 | 2.4 |  | 3.7 | 2.4 |
| Sulphates | 102 | 84.0 | 79.6 | 68.8 |  | 74.9 | 81.9 |
| Silica | 0.5 | 0.2 | 1.8 | 0.2 |  | 1.8 | 0.9 |

$$
\begin{aligned}
\text { Units : } & \mathrm{pH}=\mathrm{pH} \text { units } \\
& \text { Conductivity }=\mu \mathrm{S} / \mathrm{cm} @ 25 \mathrm{C} \\
& \text { Alkalinity }=\mathrm{mg} / \mathrm{L} \text { as } \mathrm{CaCO} 3 \\
& \text { Solids } \\
& \text { Carbon content }-\mathrm{mg} / \mathrm{L} \\
& \text { Metals } \quad \text { and }
\end{aligned}
$$

Phosphates $=\mathrm{mg}$ of $\mathrm{P} / \mathrm{L}$
Chlorides $=\mathrm{mg}$ of $\mathrm{Cl} / \mathrm{L}$
Sulphates $=\mathrm{mg}$ of $\mathrm{SO} 4 / \mathrm{L}$
Silica $=\mathrm{mg}$ of $\mathrm{SiO} 2 / \mathrm{L}$
٪. = dropped samples

## Softener regeneration brine

Hot Lime Treater


## Condensate return

## Date of sample collection

| Parameter | 3-May | 8-May | 10-May | 15-May | 17-May | 24-May | Avg. |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| General |  |  |  |  |  |  |  |
| pH | 8.9 | 8.0 | 8.9 | 9.0 | 9.2 | 8.1 | 8.7 |
| Conductivity | 13.0 | 12.4 | 18.0 | 18.3 | 22.4 | 11.8 | 16.0 |

Alkalinity

| P | 1.1 | - | 0.7 | 1.5 | 1.8 | - |  |
| ---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| M | 4.4 | 4.0 | 4.0 | 4.3 | 4.5 | 4.8 |  |
| Total | 5.5 | 4.0 | 4.8 | 5.8 | 6.3 | 4.8 | $\mathbf{5 . 2}$ |

Solids

| Total | 66.3 | 27.5 | 0.0 | 238.4 | 63.8 | 30.5 | $\mathbf{7 1 . 1}$ |
| ---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| Suspended | 0.17 | 0.00 | 0.00 | 0.40 | 0.63 | 0.00 | $\mathbf{0 . 2 0}$ |
| Dissolved | 66.1 | 27.5 | 0.0 | 238.0 | 63.1 | 30.5 | $\mathbf{7 0 . 9}$ |

Carbon content

| Total | 5.8 | 5.7 | 4.2 | 0.0 | 8.4 | 5.6 |
| ---: | :--- | :--- | :--- | :--- | :--- | :--- |
| Organic | 4.3 | 4.3 | 3.2 | 4.3 | 4.2 | 4.2 |
| Inorganic | 1.5 | 1.4 | 1.0 | - | 4.2 | 1.4 |
| $\mathbf{4 . 1}$ |  |  |  |  |  |  |

Metals

| Calcium | 0.4 | 0.3 | 0.3 | 0.3 | 0.1 | 0.2 | $\mathbf{0 . 2}$ |
| ---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| Magnesium | 0.3 | 0.2 | 0.1 | 0.2 | 0.1 | 0.1 | $\mathbf{0 . 2}$ |
| Sodium | 1.1 | 0.0 | 3.1 | 1.5 | 0.7 | 1.6 | $\mathbf{1 . 3}$ |
| Iron | 0.04 | 0.22 | 0.59 | 0.15 | 0.18 | 0.04 | $\mathbf{0 . 2 0}$ |

Inorganics

| Phosphates | 0.05 | 0.05 | 0.05 | 0.05 | 0.05 | 0.05 | $\mathbf{0 . 0 5}$ |
| ---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| Chlorides | 0.1 | 0.1 | 0.1 | 0.1 | 0.1 | 0.1 | $\mathbf{0 . 1}$ |
| Sulphates | 2.1 | 1.9 | 1.7 | 1.0 | 1.9 | 1.0 | $\mathbf{1 . 6}$ |
| Silica | 0.1 | 0.1 | 0.1 | 0.1 | 0.1 | 0.1 | $\mathbf{0 . 1}$ |

Units : $\mathrm{pH}=\mathrm{pH}$ units
Conductivity $=\mu \mathrm{S} / \mathrm{cm}$ @ 25 C
Alkalinity $=\mathrm{mg} / \mathrm{L}$ as CaCO 3
Solids
Carbon content
Metals $\quad \mathrm{mg} / \mathrm{L}$

Phosphates $=\mathrm{mg}$ of $\mathrm{P} / \mathrm{L}$
Chlorides $=\mathrm{mg}$ of $\mathrm{Cl} / \mathrm{L}$
Sulphates $=\mathrm{mg}$ of $\mathrm{SO} / \mathrm{L}$
Silica $=\mathrm{mg}$ of $\mathrm{SiO} 2 / \mathrm{L}$

## Boiler blowdown

## Date of sample collection

| Parameter | 3-May | 8-May | 10-May | 15-May | 17-May | 24-May | Avg. |
| :--- | ---: | ---: | ---: | ---: | ---: | ---: | ---: |
| General |  |  |  |  |  |  |  |
|  | pH | 11.4 | 11.5 | 11.6 | 11.5 | 11.6 | 11.4 |
| Conductivity | 2084 | 3853 | 4724 | 2666 | 3602 | 2799 | $\mathbf{3 2 8 8}$ |

Alkalinity

| P | 167 | 230 | 306 | 222 | 292 | 224 |
| ---: | :---: | :---: | :---: | :---: | :---: | :---: |
| M | 45.0 | 78.0 | 72.0 | 58.0 | 68.0 | 64.0 |
| Total | 212 | 308 | 378 | 280 | 360 | 288 |

304

Solids

| Total | 1265 | 1910 | 2080 | 1558 | 1702 | 1320 | $\mathbf{1 6 3 9}$ |
| ---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| Suspended | 4.29 | 19.6 | 43.9 | 1.18 | 0.30 | 18.5 | $\mathbf{1 4 . 6}$ |
| Dissolved | 1261 | 1890 | 2036 | 1557 | 1702 | 1301 | $\mathbf{1 6 2 5}$ |

Carbon content

| Total | 48.2 | 77.8 | 93.4 | 44.2 | 73.8 | 63.5 | $\mathbf{6 6 . 8}$ |
| ---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| Organic | 37.8 | 57.6 | 60.9 | 37.6 | 46.6 | 44.5 | $\mathbf{4 7 . 5}$ |
| Inorganic | 10.4 | 20.2 | 32.5 | 6.6 | 27.2 | 19.0 | $\mathbf{1 9 . 3}$ |

Metals

| Calcium | 2.9 | 1.2 | 2.0 | 0.1 | 0.1 | 1.0 | $\mathbf{1 . 2}$ |
| ---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| Magnesium | 1.0 | 1.0 | 1.2 | 0.4 | 0.2 | 0.7 | $\mathbf{0 . 8}$ |
| Sodium | 358 | 603 | 676 | 417 | 503 | 397 | $\mathbf{4 9 2}$ |
| Iron | 0.95 | 1.08 | 5.55 | 0.08 | 0.17 | 0.48 | $\mathbf{1 . 3 9}$ |

Inorganics

| Phosphates | 6.24 | 10.4 | 7.02 | 7.65 | 8.32 | 8.59 | $\mathbf{8 . 0 4}$ |
| ---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| Chlorides | 15.5 | 21.8 | 24.5 | 15.1 | 17.8 | 15.6 | $\mathbf{1 8 . 4}$ |
| Sulphates | 613 | 963 | 1134 | 656 | 756 | 579 | $\mathbf{7 8 3}$ |
| Silica | 5.5 | 6.7 | 8.0 | 7.2 | 8.6 | 6.5 | $\mathbf{7 . 1}$ |

Units : $\mathrm{pH}=\mathrm{pH}$ units
Conductivity= $\mu \mathrm{S} / \mathrm{cm}$ @ 25 C
Alkalinity $=\mathrm{mg} / \mathrm{L}$ as CaCO 3
Solids
Carbon content
Metals

Phosphates $=\mathrm{mg}$ of $\mathrm{P} / \mathrm{L}$ Chlorides $=\mathrm{mg}$ of $\mathrm{Cl} / \mathrm{L}$ Sulphates $=\mathrm{mg}$ of $\mathrm{SO} 4 / \mathrm{L}$ Silica $=m g$ of $\mathrm{SiO} 2 / \mathrm{L}$

## Sulphur Plant blowdown

| Parameter | Date of sample collection |  |  |  |  |  |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  | 3-May | 8-May | 10-May | 15-May | 17-May | 24-May | Avg. |
| General |  |  |  |  |  |  |  |
| pH | 11.0 | 11.0 | 11.0 | 11.2 | 11.2 | 11.1 | 11.1 |
| Conductivity | 1120 | 1665 | 1451 | 1399 | 1466 | 1478 | 1430 |
| Alkalinity |  |  |  |  |  |  |  |
| P | 94.5 | 106 | 105 | 128 | 127 | 122 |  |
| M | 34.5 | 35.0 | 36.0 | 38.0 | 39.0 | 40.0 |  |
| Total | 129 | 141 | 141 | 166 | 166 | 162 | 151 |
| $\underline{\text { Solids }}$ |  |  |  |  |  |  |  |
| Total | 849 | 843 | 609 | 1013 | 764 | 679 | 793 |
| Suspended | 0.33 | 0.00 | 0.89 | 1.21 | 1.41 | 0.32 | 0.69 |
| Dissolved | 848 | 843 | 608 | 1011 | 762 | 678 | 792 |
| Carbon content |  |  |  |  |  |  |  |
| Total | 40.5 | 48.9 | 49.3 | 40.7 | 46.3 | 48.5 | 45.7 |
| Organic | 33.0 | 36.3 | 32.9 | 35.2 | 35.1 | 34.5 | 34.5 |
| Inorganic | 7.5 | 12.6 | 16.4 | 5.5 | 11.2 | 14.0 | 11.2 |
| Metals |  |  |  |  |  |  |  |
| Calcium | 1.3 | 0.3 | 0.3 | 0.3 | 0.3 | 0.2 | 0.4 |
| Magnesium | 0.5 | 0.4 | 0.3 | 0.4 | 0.4 | 0.5 | 0.4 |
| Sodium | 207 | 241 | 223 | 234 | 218 | 194 | 219 |
| Iron | 0.31 | 0.25 | 0.68 | 0.23 | 0.17 | 0.07 | 0.29 |
| Inorganics |  |  |  |  |  |  |  |
| Phosphates | 4.88 | 5.47 | 4.95 | 6.52 | 6.57 | 7.09 | 5.91 |
| Chlorides | 8.8 | 8.7 | 8.3 | 9.6 | 8.9 | 8.2 | 8.8 |
| Sulphates | 360 | 441 | 418 | 362 | 355 | 276 | 369 |
| Silica | 3.1 | 3.4 | 3.4 | 4.0 | 3.5 | 3.7 | 3.5 |
| Units : $\mathrm{pH}=\mathrm{pH}$ units |  |  |  |  | Phosphates $=\mathrm{mg}$ of $\mathrm{P} / \mathrm{L}$ |  |  |
| Conductivity $=\mu \mathrm{S} / \mathrm{cm}$ @ 25 C |  |  |  |  | Chlorides $=\mathrm{mg}$ of $\mathrm{Cl} / \mathrm{L}$ |  |  |
| Alkalinity $=\mathrm{mg} / \mathrm{L}$ as CaCO 3 |  |  |  |  | Sulphates | $=\mathrm{mg}$ of S |  |
|  | Solids Carbon cont Metals |  | - mg/L |  | Silica $=\mathrm{m}$ | $\mathrm{g} \text { of } \mathrm{SiO} 2 /$ |  |

## Lime Pond discharge

## Date of sample collection

| Parameter | 3-May | 8-May | 10-May | 15-May | 17-May | 24-May | Avg. |
| :--- | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| General |  |  |  |  |  |  |  |
|  | pH | 9.6 | 9.9 | 9.9 | 10.1 | 10.4 | 10.4 |
| Conductivity | 4256 | 431 | 5327 | 4652 | 4478 | 5543 | $\mathbf{4 1 1 4}$ |

Alkalinity

| P | 21.8 | 33.0 | 27.5 | 35.5 | 36.0 | 46.0 |  |
| ---: | ---: | ---: | ---: | ---: | ---: | ---: | ---: |
| M | 45.5 | 50.0 | 33.0 | 41.0 | 30.0 | 35.5 |  |
| Total | 67.3 | 83.0 | 60.5 | 76.5 | 66.0 | 81.5 | $\mathbf{7 2 . 5}$ |

Solids

| Total | 2663 | 2205 | 2515 | 2896 | 2343 | 2739 | $\mathbf{2 5 6 0}$ |
| ---: | ---: | ---: | ---: | ---: | ---: | ---: | ---: |
| Suspended | 2.44 | 4.67 | 12.5 | 2.65 | 11.4 | 7.84 | $\mathbf{6 . 9 1}$ |
| Dissolved | 2660 | 2200 | 2503 | 2894 | 2331 | 2731 | $\mathbf{2 5 5 3}$ |

Carbon content
$\begin{array}{llllllll}\text { Total } & 24.9 & 31.8 & 26.5 & 20.4 & 25.7 & 30.7 & \mathbf{2 6 . 7}\end{array}$
$\begin{array}{llllllll}\text { Organic } & 13.7 & 15.7 & 15.2 & 15.2 & 18.2 & 19.4 & \mathbf{1 6 . 2}\end{array}$
$\begin{array}{llllllll}\text { Inorganic } & 11.2 & 16.1 & 11.3 & 5.2 & 7.5 & 11.3 & \mathbf{1 0 . 4}\end{array}$
Metals

| Calcium | 260 | 159 | 261 | 241 | 198 | 207 | $\mathbf{2 2 1}$ |
| ---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| Magnesium | 22.2 | 17.5 | 19.5 | 26.5 | 17.8 | 15.8 | $\mathbf{1 9 . 9}$ |
| Sodium | 622 | 570 | 672 | 639 | 568 | 683 | $\mathbf{6 2 6}$ |
| Iron | 0.17 | 0.16 | 0.40 | 0.04 | 0.20 | 0.04 | $\mathbf{0 . 1 7}$ |

Inorganics

| Phosphates | 0.16 | 0.24 | 0.05 | 0.05 | 0.05 | 0.05 | $\mathbf{0 . 1 0}$ |
| ---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| Chlorides | 918 | 682 | 914 | 1034 | 898 | 1017 | $\mathbf{9 1 1}$ |
| Sulphates | 792 | 615 | 601 | 578 | 579 | 550 | $\mathbf{6 1 9}$ |
| Silica | 4.3 | 4.5 | 4.4 | 3.5 | 3.0 | 2.8 | $\mathbf{3 . 8}$ |

Units: $\mathrm{pH}=\mathrm{pH}$ units
Conductivity $=\mu \mathrm{S} / \mathrm{cm} @ 25 \mathrm{C}$
Alkalinity $=\mathrm{mg} / \mathrm{L}$ as CaCO 3
Solids
Carbon content $-\mathrm{mg} / \mathrm{L}$
Metals

Phosphates $=\mathrm{mg}$ of $\mathrm{P} / \mathrm{L}$
Chlorides $=\mathrm{mg}$ of $\mathrm{Cl} / \mathrm{L}$
Sulphates $=\mathrm{mg}$ of $\mathrm{SO} 4 / \mathrm{L}$
Silica $=\mathrm{mg}$ of $\mathrm{SiO} 2 / \mathrm{L}$

## Effluent Plant discharge

Date of sample collection

|  | Date of sample collection |  |  |  |  |  |  |
| ---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| Parameter | 3-May | 8-May | 10-May | 15-May | 17-May | 24-May | Avg. |
| General |  |  |  |  |  |  |  |
| PH | 7.4 | 7.9 | 7.8 | 7.7 | 7.5 | 6.0 | $\mathbf{7 . 4}$ |
| Conductivity | 3316 | 4456 | 4390 | 3999 | 4513 | 4694 | $\mathbf{4 2 2 8}$ |
| Alkalinity |  |  |  |  |  |  |  |
| P | - | - | - | - | - | - |  |
| M | 121 | 274 | 249 | 241 | 161 | 23.0 |  |
| Total | 121 | 274 | 249 | 241 | 161 | 23.0 | $\mathbf{1 7 8}$ |
|  |  |  |  |  |  |  |  |
| Solids |  |  |  |  |  |  |  |
| Total | 2003 | 2849 | 2828 | 3280 | 3009 | 3201 | $\mathbf{2 8 6 2}$ |
| Suspended | 14.9 | 6.15 | 8.55 | 5.96 | 6.44 | 24.7 | $\mathbf{1 1 . 1}$ |
| Dissolved | 1988 | 2843 | 2819 | 3274 | 3002 | 3177 | $\mathbf{2 8 5 0}$ |
|  |  |  |  |  |  |  |  |
| Carbon content |  |  |  |  |  |  |  |
| Total | 197 | 220 | 212 | 192 | 182 | 209 | $\mathbf{2 0 2}$ |
| Organic | 156 | 131 | 128 | 123 | 129 | 192 | $\mathbf{1 4 3}$ |
| Inorganic | 41.3 | 89.1 | 84.5 | 69.7 | 53.0 | 17.0 | $\mathbf{5 9 . 1}$ |
|  |  |  |  |  |  |  |  |
| Metals |  |  |  |  |  |  |  |
| Calcium | 56.6 | 53.4 | 54.4 | 54.9 | 53.4 | 62.7 | $\mathbf{5 5 . 9}$ |
| Magnesium | 17.5 | 16.8 | 16.7 | 17.1 | 16.8 | 18.0 | $\mathbf{1 7 . 2}$ |
| Sodium | 499 | 596 | 599 | 580 | 561 | 516 | $\mathbf{5 5 8}$ |
| Iron | 3.14 | 3.11 | 3.31 | 3.28 | 3.42 | 4.88 | $\mathbf{3 . 5 2}$ |
| Inorganics |  |  |  |  |  |  |  |
| Phosphates | 0.69 | 0.05 | 0.05 | 0.05 | 0.05 | 0.05 | $\mathbf{0 . 1 6}$ |
| Chlorides | 209 | 194 | 187 | 217 | 199 | 211 | $\mathbf{2 0 3}$ |
| Sulphates | 323 | 309 | 340 | 326 | 326 | 326 | $\mathbf{3 2 5}$ |
| Silica | 7.3 | 6.9 | 7.5 | 7.2 | 6.6 | 5.8 | $\mathbf{6 . 9}$ |

Units : $\mathrm{pH}=\mathrm{pH}$ units
Conductivity $=\mu \mathrm{S} / \mathrm{cm} @ 25 \mathrm{C}$
Alkalinity $=\mathrm{mg} / \mathrm{L}$ as CaCO 3
Solids
Carbon content
Metals $\quad \mathrm{mg} / \mathrm{L}$

Phosphates $=\mathrm{mg}$ of $\mathrm{P} / \mathrm{L}$
Chlorides $=\mathrm{mg}$ of $\mathrm{Cl} / \mathrm{L}$ Sulphates $=\mathrm{mg}$ of SO4/L Silica $=\mathrm{mg}$ of $\mathrm{SiO} 2 / \mathrm{L}$

## Polishing Pond discharge

| Date of sample collection |  |  |  |  |  |  |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| Parameter | 3-May | 8-May | 10-May | 15-May | 17-May | 24-May | Avg. |
| General |  |  |  |  |  |  |  |
| pH | 7.3 | 7.8 | 7.9 | 7.8 | 7.6 | 7.1 | 7.6 |
| Conductivity | 3846 | 4471 | 4519 | 4341 | 4868 | 4985 | 4505 |
| Alkalinity |  |  |  |  |  |  |  |
| P | - | - | - | - | - | - |  |
| M | 73.0 | 119 | 125 | 119 | 116 | 51.0 |  |
| Total | 73.0 | 119 | 125 | 119 | 116 | 51.0 | 101 |
| Solids |  |  |  |  |  |  |  |
| Total | 2655 | 2585 | 2480 | 3074 | 2919 | 2914 | 2771 |
| Suspended | 18.0 | 10.3 | 33.0 | 41.7 | 12.2 | 47.1 | 27.1 |
| Dissolved | 2637 | 2575 | 2447 | 3032 | 2907 | 2867 | 2744 |
| Carbon content |  |  |  |  |  |  |  |
| Total | 95.6 | 105 | 107 | 90.6 | 93.3 | 94.8 | 97.7 |
| Organic | 73.7 | 74.7 | 69.3 | 61.3 | 59.2 | 75.5 | 69.0 |
| Inorganic | 21.9 | 30.7 | 37.3 | 29.3 | 34.1 | 19.3 | 28.8 |
| Metals |  |  |  |  |  |  |  |
| Calcium | 146 | 134 | 137 | 157 | 157 | 146 | 146 |
| Magnesium | 21.7 | 19.9 | 19.0 | 22.2 | 21.5 | 19.1 | 20.6 |
| Sodium | 567 | 578 | 580 | 604 | 596 | 578 | 584 |
| Iron | 1.89 | 1.66 | 1.68 | 1.51 | 1.48 | 1.98 | 1.70 |
| Inorganics |  |  |  |  |  |  |  |
| Phosphates | 0.51 | 0.47 | 0.35 | 0.11 | 0.11 | 0.06 | 0.27 |
| Chlorides | 606 | ? | 538 | 703 | 686 | 602 | 627 |
| Sulfates | 539 | 515 | 526 | 558 | 574 | 546 | 543 |
| Silica | 5.2 | 5.5 | 5.8 | 5.4 | 5.2 | 4.4 | 5.2 |

Units: $\mathrm{pH}=\mathrm{pH}$ units
Conductivity $=\mu \mathrm{S} / \mathrm{cm} @ 25 \mathrm{C}$
Alkalinity $=\mathrm{mg} / \mathrm{L}$ as CaCO 3
$\begin{aligned} & \text { Solids } \\ & \text { Carbon content } \\ & \text { Metals }\end{aligned} \quad \mathrm{mg} / \mathrm{L}, ~$

Phosphates $=\mathrm{mg}$ of $\mathrm{P} / \mathrm{L}$
Chlorides $=\mathrm{mg}$ of $\mathrm{Cl} / \mathrm{L}$ Sulphates $=\mathrm{mg}$ of $\mathrm{SO} 4 / \mathrm{L}$ Silica $=\mathrm{mg}$ of $\mathrm{SiO} 2 / \mathrm{L}$

## Flared water

| Date of sample collection |  |  |  |  |  |  |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| Parameter | 3-May | 8-May | 10-May | 15-May | 17-May | 24-May | Avg. |
| General |  |  |  |  |  |  |  |
| pH | 6.0 | 5.4 | 5.4 | 5.2 | 4.8 | No | 5.4 |
| Conductivity | 60.2 | Not measured |  |  |  | Sample | 60.2 |
| Alkalinity |  |  |  |  |  |  |  |
| P | - | - | - | - | - |  |  |
| M | 3.9 | 1.5 | 1.8 | 1.8 | 1.8 |  |  |
| Total | 3.9 | 1.5 | 1.8 | 1.8 | 1.8 |  | 2.1 |
| Solids |  |  |  |  |  |  |  |
| Total | 125 | 12.5 | $\square$ | 301 | 22.8 |  | 115 |
| Suspended | 8.75 | 0.00 | 7.31 | 58.3 | 1.68 |  | 68.8 |
| Dissolved | 116 | 12.5 | ? | 243 | 21.1 |  | 98 |
| Carbon content |  |  |  |  |  |  |  |
| Total | 94.0 | 1481 | 433 | 187 | 205 |  | 700 |
| Organic | 97.7 | 1255 | 330 | 91.5 | 211 |  | 559 |
| Inorganic | \% | 226 | 103 | 95.5 | $\stackrel{ }{ }$ |  | 142 |
| Metals |  |  |  |  |  |  |  |
| Calcium No |  | 0.3 | 0.2 | 2.8 | 0.1 |  | 0.9 |
| Magnesium | Sample | 0.1 | 0.1 | 0.3 | 0.1 |  | 0.1 |
| Sodium |  | 0.9 | 0.1 | 4.6 | 0.6 |  | 1.5 |
| Iron |  | 1.93 | 2.05 | 7.14 | 4.51 |  | 3.91 |
| Inorganics |  |  |  |  |  |  |  |
| Phosphates | 0.05 | 0.05 | 0.05 | 0.05 | 0.05 |  | 0.05 |
| Chlorides | 3.2 | 1.9 | 0.8 | 1.2 | 1.8 |  | 1.8 |
| Sulphates | 4.1 | 3.8 | 6.4 | 5.6 | 2.0 |  | 4.4 |
| Silica | 0.7 | 0.6 | 0.4 | 17.2 | 1.8 |  | 4.1 |
| Units : $\mathrm{pH}=\mathrm{pH}$ units |  |  |  |  | Phosphates $=\mathrm{mg}$ of $\mathrm{P} / \mathrm{L}$ |  |  |
| Conductivity $=\mu \mathrm{S} / \mathrm{cm}$ @ 25 C |  |  |  |  | Chlorides $=\mathrm{mg}$ of $\mathrm{Cl} / \mathrm{L}$ |  |  |
| Alkalinity = mg/L as CaCO 3 |  |  |  |  | Sulphates $=\mathrm{mg}$ of SO4/L |  |  |
| SolidsCarbon content |  |  |  |  | Silica $=\mathrm{mg}$ of $\mathrm{SiO} 2 / \mathrm{L}$ |  |  |
| Carbon content |  |  | - mg/L |  | \%... dropped samples |  |  |

APPENDIX D

MASS BALANCE EQUATIONS AND ASSUMPTIONS

## Treater System:

The following streams were used in the mass balance:
Incoming
-Raw water header
-Fresh brine solution
-8 psi steam

The fresh brine solution was never sampled. It was assumed to be identical in chemical composition to the treated water, except for sodium and chloride levels. These values were calculated independently from figures given in the Water Treatment Manual:

Use 792 lb . of NaCl
\& 10480 USgal of water to regenerate 1 softener
Concentration of:

$$
\begin{aligned}
\mathrm{NaCl}= & (792 \mathrm{lb} \times 0.4536 \mathrm{lb} / \mathrm{kg} \times 1000000 \mathrm{mg} / \mathrm{kg}) /(10480 \mathrm{USgal} \mathrm{x} \\
& 3.7854 \mathrm{~L} / \mathrm{gal}) \\
= & 9055.8 \mathrm{mg} / \mathrm{L} \\
\mathrm{Na}= & {[\mathrm{NaCl}] \times 23 / 58.45 } \\
= & 3563 \mathrm{mg} / \mathrm{L} \\
\mathrm{Cl}= & {[\mathrm{NaCl}]-[\mathrm{Na}] } \\
= & 5492 \mathrm{mg} / \mathrm{L}
\end{aligned}
$$

The 8 psi steam entering the system and the vented vapour leaving it were assumed to be identical not only to one another, but also to the condensate sampled elsewhere in the FNGP (see Appendix C).

## Boiler System

The following streams were used in the mass balance:

| Incoming | Outgoing |
| :--- | :--- |
| - Boiler feedwater | -450 psi steam |
|  | - Boiler blowdown |

Chemical characteristics of the boiler feedwater (BFW) were derived from the treated water and condensate return flows, which mix in the deaerators prior to entry into the boilers. A sample calculation is shown below:

Total solids in BFW:

$$
\begin{aligned}
& =\left(\left(\text { Concentration }_{\mathrm{TW}} \times \text { Flow }_{\mathrm{TW}}\right)+\left(\text { Concentration }_{\text {Cond }} \times \text { Flow }_{\text {Cond }}\right)\right) / \\
& \left(\text { Flow }_{\mathrm{TW}}+\text { Flow }_{\text {Cond }}\right)
\end{aligned}
$$

The 450 psi steam leaving the boilers was assumed to have the same chemistry as the condensate return samples collected elsewhere in the FNGP.

The flow of boiler feedwater into the boilers was assumed to be $=$ to the flow of 450 psi steam and blowdown exiting the boiler.

## Lime Pond

The following streams were used in the mass balance:

| Incoming | Outgoing |
| :--- | :--- |
| -\#3 Hot lime treater blowdown | -Lime pond discharge |
| -Boiler blowdown |  |
| -Softener regeneration blowdown |  |
| -Sulphur plant blowdown |  |

Water samples were collected from each of these streams, so it was simply a case of multiplying know concentrations by the respective flow rates.

## Composite Discharge:

The following streams were used in the mass balance:

Incoming
-Effluent plant discharge

Outgoing
-Polishing pond discharge
-Lime pond discharge

Water samples were collected from each of these streams, so it was simply a case of multiplying know concentrations by the respective flow rates.

## ESTIMATED HOT LIME TREATER AND ION-EXCHANGER BLOWDOWN CHARACTERISTICS

## Hot Lime Treater Blowdown:

The water treatment manual stipulates that the hot lime treater removes calcium, magnesium, silica and carbonate. It also indicates that if the total hardness of the inlet water is larger than the inlet M alkalinity, the lime treater effluent will have the following characteristics:

$$
\mathrm{CO}_{3} \quad=25 \mathrm{mg} / \mathrm{L} \text { as } \mathrm{CaCO}_{3}
$$

$$
\begin{array}{ll}
\mathrm{OH} & =5 \mathrm{mg} / \mathrm{L} \text { as } \mathrm{CaCO}_{3} \\
\text { Total alk. } & =30 \mathrm{mg} / \mathrm{L} \text { as } \mathrm{CaCO}_{3} \\
\mathrm{Mg} & =5 \mathrm{mg} / \mathrm{L} \text { as } \mathrm{CaCO}_{3} \\
\mathrm{Ca} & =\text { (raw water hardness - hardness reduction) - final }[\mathrm{Mg}], \\
& \text { where hardness reduction = alkalinity reduction }
\end{array}
$$

Given that the inlet and outlet waters from the front-end softening system were observed to contain:

|  | Influent <br> Flow (m3/day) | Outlet <br> $\mathrm{Ca}(\mathrm{mg} / \mathrm{L})$ |
| :--- | :---: | :---: |
| $\mathrm{Mg}(\mathrm{mg} / \mathrm{L})$ | 1030 | 975 |
| $\mathrm{Fe}(\mathrm{mg} / \mathrm{L})$ | 67.8 | 1.4 |
| $\mathrm{SO}_{4}(\mathrm{mg} / \mathrm{L})$ | 15.3 | 0.5 |
| $\mathrm{SiO}_{2}(\mathrm{mg} / \mathrm{L})$ | 0.4 | 0.6 |
| $\mathrm{PO}_{4}(\mathrm{mg} / \mathrm{L})$ | 72.7 | 81.9 |
| $\mathrm{Cl}(\mathrm{mg} / \mathrm{L})$ | 4.4 | 0.9 |
| $\mathrm{Na}(\mathrm{mg} / \mathrm{L})$ | 0.06 | 0.07 |
| $\mathrm{DOC}(\mathrm{mg} / \mathrm{L})$ | 2.4 | 2.4 |
| pH | 10.6 | 50.5 |
| $\mathrm{M} . \mathrm{Alk}\left(\mathrm{mg} / \mathrm{L}\right.$ as $\left.\mathrm{CaCO}_{3}\right)$ | 133 | 6.1 |
| $\mathrm{~T} . \mathrm{Alk}\left(\mathrm{mg} / \mathrm{L}\right.$ as $\left.\mathrm{CaCO}_{3}\right)$ | 133 | 9.9 |
| $\mathrm{CO}(\mathrm{mol} / \mathrm{L})$ | $1.6 \times 10-5$ | 18.4 |
| $\mathrm{C}_{\mathrm{T}}(\mathrm{mol} / \mathrm{L})$ | 3.1 | 27.8 |

Product water from the hot lime treater should have contained:

| Ca | $=42.5 \mathrm{mg} / \mathrm{L}$ | - hardness - alk. reduction $-[\mathrm{Mg}]$ |
| :--- | :--- | :--- |
| Mg | $=1.2 \mathrm{mg} / \mathrm{L}$ | $-5 \mathrm{mg} / \mathrm{L}$ as $\mathrm{CaCO}_{3}=1.2 \mathrm{mg} / \mathrm{L}$ |
| Fe | $=0.6 \mathrm{mg} / \mathrm{L}$ | - assumed to be identical to the outlet water |
| $\mathrm{SO}_{4}=81.9 \mathrm{mg} / \mathrm{L}$ | - assumed to be identical to the outlet water |  |
| $\mathrm{SiO}_{2}=0.9 \mathrm{mg} / \mathrm{L}$ | - assumed to be identical to the outlet water |  |
| $\mathrm{PO}_{4}=0.07 \mathrm{mg} / \mathrm{L}$ | - assumed to be identical to the outlet water |  |
| Cl | $=2.4 \mathrm{mg} / \mathrm{L}$ | - assumed to be identical to the outlet water |
| Na | $=10.1 \mathrm{mg} / \mathrm{L}$ | - assumed to be identical to the [Na] observed in the single |
|  |  | blowdown sample |
| DOC | $=6.1 \mathrm{mg} / \mathrm{L}$ | - assumed to be identical to the outlet water |

$$
\begin{array}{lll}
\mathrm{pH} & =10.0 & -[\mathrm{OH}] \text { of } 5 \mathrm{mg} / \mathrm{L} \text { as } \mathrm{CaCO}_{3}=\mathrm{pH} \text { of } 10.0 \\
\mathrm{~T} . \text { alk. } & =30 \mathrm{mg} / \mathrm{L} \mathrm{CaCO}_{3} & - \text { as stipulated by water treatment manual } \\
\mathrm{CO}_{3} & =1.2 \times 10^{-4} \mathrm{~mol} / \mathrm{L} & -25 \mathrm{mg} / \mathrm{L} \text { as } \mathrm{CaCO}_{3}=1.2 \times 10^{-4} \mathrm{~mol} / \mathrm{L} \\
\mathrm{C}_{\mathrm{T}} & =3.8 \times 10^{-4} \mathrm{~mol} / \mathrm{L} & \text { - calculated from total alkalinity and } \mathrm{pH}
\end{array}
$$

And the blowdown stream which flowed at $16.3 \mathrm{~m}^{3} /$ day, would have had the following characteristics:

| Ca | $=1640 \mathrm{mg} / \mathrm{L}$ | - (1030 m3/day x $67.8 \mathrm{mg} / \mathrm{L}-42.5 \mathrm{mg} / \mathrm{L} \times 1014 \mathrm{~m} 3 /$ day $)$ |
| :---: | :---: | :---: |
| Mg | $=887 \mathrm{mg} / \mathrm{L}$ | - (1030 m³/day x $15.3 \mathrm{mg} / \mathrm{L}-1.2 \mathrm{mg} / \mathrm{L} \times 1014 \mathrm{~m} 3 /$ day $)$ |
| Fe | $=0.3 \mathrm{mg} / \mathrm{L}$ | - observed [] in single blowdown sample |
| $\mathrm{SO}_{4}$ | $=74.2 \mathrm{mg} / \mathrm{L}$ | - observed [] in single blowdown sample |
| $\mathrm{SiO}_{2}$ | $=219 \mathrm{mg} / \mathrm{L}$ | - ( $1030 \mathrm{~m} 3 /$ day $\times 4.4 \mathrm{mg} / \mathrm{L}-1014 \mathrm{~m} /$ day x $0.9 \mathrm{mg} / \mathrm{L}$ ) |
| $\mathrm{PO}_{4}$ | $=0.05 \mathrm{mg} / \mathrm{L}$ | - observed [] in single blowdown sample |
| Cl | $=2.8 \mathrm{mg} / \mathrm{L}$ | - observed [] in single blowdown sample |
| Na | $=10.1 \mathrm{mg} / \mathrm{L}$ | - observed [] in single blowdown sample |
| DOC | $=28.5 \mathrm{mg} / \mathrm{L}$ | - ( $1030 \mathrm{~m} 3 /$ day $\times 8.4 \mathrm{mg} / \mathrm{L}-1014 \mathrm{~m} 3 /$ day $\times 6.1 \mathrm{mg} / \mathrm{L}$ ) |
| pH | $=10.0$ | - assumed to be the same as the lime treater effluent |
| $\mathrm{C}_{\mathrm{T}}$ | $=1.7 \times 10^{-1} \mathrm{~mol} / \mathrm{L}$ | $\begin{aligned} & -\left(1030 \mathrm{~m}^{3} / \mathrm{day} \times 3.1 \times 10^{-3} \mathrm{~mol} / \mathrm{L}-1014 \mathrm{~m}^{3} / \text { day } \times 3.8 \times\right. \\ & \left.\quad 10^{-4} \mathrm{~mol} / \mathrm{L}\right) \end{aligned}$ |
| $\mathrm{CO}_{3}$ | $=1.7 \times 10^{-1} \mathrm{~mol} / \mathrm{L}$ | $-\left[\mathrm{H}_{2} \mathrm{CO}_{3}\right]+\left[\mathrm{HCO}_{3}\right] \lll\left[\mathrm{CO}_{3}\right]$, so $\left[\mathrm{C}_{\mathrm{T}}\right]=\left[\mathrm{CO}_{3}\right]$ |

## Ion-exchange blowdown:

Given the following data:

|  | Lime treater effluent | Outlet water |
| :--- | :---: | :---: |
| Flow (m3/day) | 1014 | 975 |
| $\mathrm{Ca}(\mathrm{mg} / \mathrm{L})$ | 42.5 | 1.4 |
| $\mathrm{Mg}(\mathrm{mg} / \mathrm{L})$ | 1.2 | 0.5 |
| $\mathrm{Fe}(\mathrm{mg} / \mathrm{L})$ | 0.6 | 0.6 |
| $\mathrm{SO}_{4}(\mathrm{mg} / \mathrm{L})$ | 81.9 | 81.9 |
| $\mathrm{SiO}_{2}(\mathrm{mg} / \mathrm{L})$ | 0.9 | 0.9 |
| $\mathrm{PO}_{4}(\mathrm{mg} / \mathrm{L})$ | 0.07 | 0.07 |
| $\mathrm{Cl}(\mathrm{mg} / \mathrm{L})$ | 2.4 | 2.4 |
| $\mathrm{Na}(\mathrm{mg} / \mathrm{L})$ | 10.1 | 50.5 |
| $\mathrm{DOC}(\mathrm{mg} / \mathrm{L})$ | 6.1 | 6.1 |
| $\mathrm{pH}(\mathrm{mg} / \mathrm{L})$ | 10.0 | 10.1 |
| $\mathrm{~T} . \mathrm{Alk}\left(\mathrm{mg} / \mathrm{L}\right.$ as $\left.\mathrm{CaCO}_{3}\right)$ | 30 | 27.8 |


| $\mathrm{CO}_{3}(\mathrm{~mol} / \mathrm{L})$ | $1.2 \times 10^{-4}$ | $1.0 \times 10^{-4}$ |
| :--- | :--- | :--- |
| $\mathrm{C}_{\mathrm{T}}(\mathrm{mol} / \mathrm{L})$ | $3.8 \times 10^{-4}$ | $3.7 \times 10^{-4}$ |

And that the brine used to regenerate the ion-exchangers consisted of 792 lbs of NaCl dissolved in 10480 USgal of treated/outlet water.

The ion-exchanger blowdown should have had the following characteristics:

| Ca | $=1070 \mathrm{mg} / \mathrm{L}$ | - ( $1014 \mathrm{~m} 3 /$ day $\times 42.5 \mathrm{mg} / \mathrm{L}-975 \mathrm{~m} 3 /$ day $\times 1.4 \mathrm{mg} / \mathrm{L})$ |
| :---: | :---: | :---: |
| Mg | $=19.9 \mathrm{mg} / \mathrm{L}$ | - ( $1014 \mathrm{~m}^{3} /$ day $\times 1.2 \mathrm{mg} / \mathrm{L}-975 \mathrm{~m}^{3} /$ day $\left.\times 0.5 \mathrm{mg} / \mathrm{L}\right)$ |
| Fe | $=0.6 \mathrm{mg} / \mathrm{L}$ | - assumed to be identical to the outlet water |
| $\mathrm{SO}_{4}$ | $=81.9 \mathrm{mg} / \mathrm{L}$ | - assumed to be identical to the outlet water |
| $\mathrm{SiO}_{2}$ | $=0.9 \mathrm{mg} / \mathrm{L}$ | - assumed to be identical to the outlet water |
| $\mathrm{PO}_{4}$ | $=0.07 \mathrm{mg} / \mathrm{L}$ | - assumed to be identical to the outlet water |
| Cl | $=5490 \mathrm{mg} / \mathrm{L}$ | - [Cl] in brine used to regenerate the ion-exchangers |
| Na | $=2331 \mathrm{mg} / \mathrm{L}$ | $-[\mathrm{Na}]$ in brine used to regenerate the ion-exchanger $+2 x$ (equivalent mass of calcium magnesium removed by the ion-exchangers) |
| DOC | $=6.1 \mathrm{mg} / \mathrm{L}$ | - assumed to be identical to the outlet water |
| pH | $=11.0$ | - average value from 5 blowdown samples |
| T. alk. | $=318 \mathrm{mg} / \mathrm{L} \mathrm{CaCO}_{3}$ | - average value from 5 blowdown samples |
| $\mathrm{CO}_{3}$ | $=2.4 \times 10^{-3} \mathrm{~mol} / \mathrm{L}$ | - calculated from pH and total alkalinity readings |
| $\mathrm{C}_{\mathrm{T}}$ | $=2.9 \times 10^{-3} \mathrm{~mol} / \mathrm{L}$ | - calculated from pH and total alkalinity readings. |

## APPENDIX E

## EQUATIONS USED IN THE COMPUTER SIMULATOR

FNGP WATER SYSTEM OVERVIEW SPREADSHEET

| Cell(s) | Equations / values and explanation thereof |
| :---: | :---: |
| C16 | $=\mathrm{IF}$ (D14="Composite", IF (K9 $=$ "no", $(273+\mathrm{H} 7+\mathrm{I} 16+\mathrm{C} 39+\mathrm{D} 39+$ 'Reverse Osmosis'!\$P\$43)-D16,(273+H7+I16+N16+C39+D39+Reverse Osmosis'!\$P\$43)-D16),(273+H7+I16+N16+C39+D39)-D16) <br> $=$ Structure of if/then statements indicates that raw water flow into the process system has one of three possible values depending on the configuration of the ZED system: i - if using a direct composite recycle scenario without a softener, then raw water flow $=273$ (unrecoverable losses) + boiler, sulphur plant, hot lime treater and RO filter blowdowns +8 psi steam loss (kept at 0 in this study) - recycled water ii - if a softener is added to the direct composite discharge scenario, then raw water flow is the same as in (i) + softener blowdown flow iii - if using a back-end model, then raw water flow is = same as (i) + softener blowdown + without RO blowdown |
| C18-C28 | $=$ specific values from original data set |
| C29 | $\begin{aligned} & =\left(50000^{*}\left(\mathrm{C} 28 / 50000+\left(10^{\wedge}-\mathrm{C} 27\right)-\left(10^{\wedge}-14 / 10^{\wedge}-\mathrm{C} 27\right)\right) /\left(1+\left(10^{\wedge}-\mathrm{C} 27 /\left(2^{*} 4.7^{*} 10^{\wedge}-\right.\right.\right.\right. \\ & 11)))) * 0.6 / 61000 \\ & =50000 \times\left(\mathrm{T} . \text { Alk } / 50000+10-\mathrm{pH}-10^{-14} / 10-\mathrm{pH}\right) /\left(1+10-\mathrm{pH} /\left(2 \times \mathrm{H}_{2} \mathrm{CO}_{3}-\mathrm{CO}_{3}\right.\right. \\ & \quad \text { equilibrium constant })) \times 0.6 / 60000 \\ & - \text { from Sawyer \& McCarty } 1978 \end{aligned}$ |
| C30 | $\begin{aligned} &=\left(50000 *\left(\mathrm{C} 28 / 50000+\left(10^{\wedge}-\mathrm{C} 27\right)-\left(10^{\wedge}-14 / 10^{\wedge}-\mathrm{C} 27\right)\right) /\left(1+\left(\left(2 * 4.7^{*} 10^{\wedge}-11\right) / 10^{\wedge}-\right.\right.\right. \\ &\mathrm{C} 27)))^{*} 1.22 / 61000 \\ &= 50000 \times\left(\mathrm{T} . \mathrm{Alk} / 50000+10^{-} \mathrm{pH}-10^{-14} / 10^{-}-\mathrm{pH}\right) /\left(1+\left(2 \mathrm{xH}_{2} \mathrm{CO}_{3}-\mathrm{CO}_{3}\right.\right. \\ &\quad \text { equilibrium constant }) / 10-\mathrm{pH}) \times 1.22 / 61000 \\ & \text { from Sawyer \& McCarty } 1978 \end{aligned}$ |
| C31 | $\begin{aligned} & =\left(\left(10^{\wedge}-\mathrm{C} 27\right)^{*} \mathrm{C} 30\right) /\left(4.2^{*} 10^{\wedge}-7\right) \\ & =\left(10^{-} \mathrm{pH}\right) \times\left[\mathrm{HCO}_{3}\right] / \mathrm{H}_{2} \mathrm{CO}_{3}-\mathrm{HCO}_{3} \text { equilibrium constant } \end{aligned}$ |
| C32 | $\begin{aligned} & =\text { Sum (C29:C31) } \\ & =\text { Sum of Cell C29 through C31 } \end{aligned}$ |


FNGP WATER SYSTEM OVERVIEW SPREADSHEET
D16-D32
號
Equations/values and explanation thereof
$=\mathrm{IF}(\mathrm{F} 4=$ "no", 0 , 'Reverse Osmosis'! $\$ \mathrm{~N} \$ 43 \$ \mathrm{~N} \$ 62$ )
= If reclaimed wastewater collected from the RO and BCS units is recycled through
the system, then these cells have the same values as cells N43 through N62,
respectively, in the "reverse osmosis" spreadsheet

[^9]
FNGP WATER SYSTEM OVERVIEW SPREADSHEET (con't)
Cell(s)
F18-F27
F29 \& F30
F32
G16
G18-G27


FNGP WATER SYSTEM OVERVIEW SPREADSHEET (con't)

| Cell(s) | Equations / values and explanation thereof |
| :---: | :---: |
| I16 | $=\mathrm{IF}((\mathrm{G} 18 * 50 / 20+\mathrm{G} 19 * 50 / 12.2)<25,0$, ' $[$ FNGP water system] Lime treater calc.'!\$K\$81) <br> = This cell has the same value as cell H81 in the "Lime Treater Calc." spreadsheet, unless total incoming hardness is less than $25 \mathrm{mg} / \mathrm{L}$ as $\mathrm{CaCO}_{3}$. At below $25 \mathrm{mg} / \mathrm{L}$, the lime treater goes off-line, as it cannot lower water hardness below this level (water treatment manual) |
| I18, I29 \& I27 | $=$ If the lime treater is on-line (i.e. blowdown flow is $>0.001 \mathrm{~m}^{3} /$ day $)$, then these cells have the same value of cells K83, K84 \& G78 in the "Lime Treater Calc." spreadsheet. Otherwise, they have no value ( 0 is displayed) (e.g. $=\mathrm{IF}(\mathrm{I} 16<0.001$, 0 , '[FNGP water system]Lime treater calc.'! $\$ \mathrm{~K} \$ 83$ )) |
| I20 \& I23 | $\begin{aligned} & =\text { Incoming mass } x(1+\% \text { gain or loss across the lime treater }) / \text { blowdown flow, } \\ & \text { unless lime treater is off-line, in which case cell value }=0(e . g .=\operatorname{IF}(\mathrm{I} 16<0.001, \\ & \left.0, \mathrm{G} 20^{*}(1+\mathrm{H} 20)\right) \end{aligned}$ |
| I21, I24 \& I 25 | $=$ Values in cells G21, G24 and G25, respectively, if lime treater is on-line. If it is off-line, these cells have a value $=0$ (e.g. $=\mathrm{IF}(\mathrm{I} 16<0.001,0, \mathrm{G} 24))$ |
| I22 | $=\mathrm{IF}(\mathrm{I} 16<0.001,0,(\mathrm{G} 26 * \mathrm{G} 16-\mathrm{J} 26 * \mathrm{~J} 16) / \mathrm{I} 16)$ <br> $=\mathrm{IF}(\mathrm{I} 16<0.001,0$, '[FNGP water system]Lime treater calc.'!\$G\$78) <br> - this equation is based on the assumption that the lime treater can only lower silica concentrations to current treater water levels (i.e. $0.9 \mathrm{mg} / \mathrm{L}$ ), so if its feedwaters contain more than $0.9 \mathrm{mg} / \mathrm{L}$, all of the excess mass will be drained away in the blowdown waters; on the other hand, if they do not contain more than $0.9 \mathrm{mg} / \mathrm{L}$ of silica, silica concentrations in the blowdown water will be equal to incoming levels. |
| I26 | $\begin{aligned} & =\mathrm{IF}(\mathrm{I} 16<0.001,0,(\mathrm{G} 26 * \mathrm{G} 16-\mathrm{J} 26 * \mathrm{~J} 16) / \mathrm{I} 16) \\ & =\mathrm{IF}(\mathrm{I} 16<0.001,0, \text { '[FNGP water system]Lime treater calc.' } \$ \$ \mathrm{G} \$ 78) \end{aligned}$ |
| I29 | $\begin{aligned} & =\mathrm{IF}(\mathrm{I} 16<0.001,0, \mathrm{I} 32-(\mathrm{I} 31+\mathrm{I} 30)) \\ & =\left[\mathrm{C}_{\mathrm{T}}\right]-\left[\mathrm{HCO}_{3}\right]-\left[\mathrm{H}_{2} \mathrm{CO}_{3}\right] \text { or } 0, \text { depending if the lime treater is on or off-line } \end{aligned}$ |
| I30 \& I31 | $=\left[\mathrm{HCO}_{3}\right] \&\left[\mathrm{H}_{2} \mathrm{CO}_{3}\right] \text { in the lime treater product waters }(\mathrm{e} . \mathrm{g} .=\mathrm{IF}(\mathrm{I} 16<0.001,0,$ |
| I32 | $=\mathrm{IF}(\mathrm{I} 16<0.001,0,(\mathrm{G} 32 * \mathrm{G} 16-\mathrm{J} 32 * \mathrm{~J} 16) / \mathrm{I} 16)$ |

$\substack{\text { Lime Blowdown } \\ \text { Flow }}$
$\mathrm{Ca}, \mathrm{Mg} \& \mathrm{pH}$
${\mathrm{Fe} \& \mathrm{PO}_{4}}^{\text {Title }}$
$\mathrm{SO}_{4}, \mathrm{Na}_{\mathrm{Na}} \mathrm{Cl}$
$\mathrm{SiO}_{2}$
$\mathrm{DOC}^{\mathrm{CO}_{3}}$
$\mathrm{HCO}_{3} \& \mathrm{H}_{2} \mathrm{CO}_{3}$
$\mathrm{C}_{\mathrm{T}}$
FNGP WATER SYSTEM OVERVIEW SPREADSHEET (con't)

| Title | Cell(s) | Equations/values and explanation thereof |
| :---: | :---: | :---: |
| Lime Product - Cels |  |  |
| Flow | J16 | $\begin{aligned} & =\mathrm{G} 16-\mathrm{I} 16 \\ & =\text { Incoming flow }- \text { blowdown flow } \end{aligned}$ |
| - if lime treater is off-line (i.e. incoming hardness $<25 \mathrm{mg} / \mathrm{L}$ as $\mathrm{CaCO}_{3}$ ) |  |  |
| Ca-DOC \& T.Alk | J18-J26 \& J28 | = Identical to inflow waters |
| pH | J27 | $\begin{aligned} & \text { = Value of cell G78 in "Lime Treater Calc." } \\ & \text { - based on assumption that process waters always raised to } \mathrm{pH} 10 \end{aligned}$ |
| $\mathrm{CO}_{3}$ | J29 | $\begin{aligned} & =50000 \times\left(\mathrm{T} . \mathrm{Alk} / 50000+10-\mathrm{pH}-10^{-14} / 10^{-\mathrm{pH}}\right) /\left(1+10-\mathrm{pH} /\left(2 \times \mathrm{H}_{2} \mathrm{CO}_{3}-\mathrm{CO}_{3}\right.\right. \\ & \text { equilibrium constant })) \times 0.6 / 60000 \\ & - \text { from Sawyer \& McCarty } 1978 \end{aligned}$ |
| $\mathrm{HCO}_{3}$ | J30 | $\begin{aligned} & =50000 \times\left(\mathrm{T} . \mathrm{Alk} / 50000+10-\mathrm{pH}-10^{-14} / 10-\mathrm{pH}\right) /\left(1+\left(2 \mathrm{xH}_{2} \mathrm{CO}_{3}-\mathrm{CO}_{3}\right.\right. \\ & \quad \text { equilibrium constant) } / 10-\mathrm{pH}) \times 1.22 / 61000 \\ & - \text { from Sawyer \& McCarty } 1978 \end{aligned}$ |
| $\mathrm{H}_{2} \mathrm{CO}_{3}$ | J31 | $=(10-\mathrm{pH}) \times\left[\mathrm{HCO}_{3}\right] / \mathrm{H}_{2} \mathrm{CO}_{3}-\mathrm{HCO}_{3}$ equilibrium constant |
| $\mathrm{C}_{\text {T }}$ | J32 | $\begin{aligned} & =\operatorname{Sum}(\mathrm{J} 29: \mathrm{J} 31) \\ & =\left[\mathrm{CO}_{3}\right]+\left[\mathrm{HCO}_{3}\right]+\left[\mathrm{H}_{2} \mathrm{CO}_{3}\right] \end{aligned}$ |
| - if lime treater is on-line (i.e. influent hardness $>25 \mathrm{mg} / \mathrm{L}$ as $\mathrm{CaCO}_{3}$ ) |  |  |
| $\mathrm{Ca}, \mathrm{Mg}$ \& pH | J18, J19 \& J27-J32 | = Values in cells G82, G79, G78, G83 \& K71-73 in the "Lime Treater Calc." spreadsheet, respectively |
| $\mathrm{Fe}, \mathrm{SO}_{4} \& \mathrm{PO}_{4}-\mathrm{DOC}$ | J20, J21 \& J23-J26 | $=$ Influent [ ] x ( $1+\%$ gain or loss across the lime treater |
| $\mathrm{SiO}_{2}$ | J22 | $=($ Influent mass - blowdown mass) / product water flow |
| RO Product Water <br> - if using a composite discharge reuse model: |  |  |
| Flow-DOC | K16-26 | = Values specified in cells N43 through N53 in "Reverse Osmosis" spreadsheet, respectively (e.g. = IF(D14= "Composite", 'Reverse Osmosis'!\$N\$45:\$N\$53, "N/A")) |
| pH | K27 | $=\mathrm{pH}$ of 10.0 <br> - RO product water raised to pH 10.0 to match lime treater product water |
| $\mathrm{CO}_{3}$ | K29 | $=\mathrm{CO}_{3}-\mathrm{HCO}_{3}$ equilibrium constant $\mathrm{x}\left[\mathrm{HCO}_{3}\right] /(10-\mathrm{pH})$ |

FNGP WATER SYSTEM OVERVIEW SPREADSHEET (con't)
Equations / values and explanation thereof

| Cell(s) | Equations / values and explanation thereof |
| :---: | :---: |
| K30 | ```\(=\left[\mathrm{C}_{\mathrm{T}}\right] \times 1 /\left(10-\mathrm{pH} / \mathrm{H}_{2} \mathrm{CO}_{3}\right.\) equilibrium constant \(+1+\mathrm{CO}_{3}-\mathrm{HCO}_{3}\) equilibrium constant / \(10-\mathrm{pH}\) ) - from Benefield et al. 1982``` |
| K31 | $=(10-\mathrm{pH}) \times\left[\mathrm{HCO}_{3}\right] / \mathrm{H}_{2} \mathrm{CO}_{3}-\mathrm{HCO}_{3}$ equilibrium constant |
| K32 | = Value of cell N62 in the "Reverse Osmosis" spreadsheet |

[^10]FNGP WATER SYSTEM OVERVIEW SPREADSHEET (con't)
Equations / values and explanation thereof

| Cell(s) | Equations/values and explanation thereof |
| :--- | :--- |
| N32 M32 |  |
|  | - Self-explanatory |

discharge reuse scenario:
 = Cell C95 in "Softener Brine Calc." spreadsheet
$=(\mathrm{K} 18:$ K19 * K16-M18: M19 * M16) / N16
= (mass in - mass out) / blowdown volume
= K20 - K23 \& K26

- [ ]'s in RO product flow
= Cell H17 and C44, respectively, in the "Softener Brine Calc." spreadsheet M20 - M24 \& M26
M25
M27
M29
M30

M31
M32
 $\frac{\text { Title }}{\mathrm{C}_{\mathrm{T}}}$

- If using an ion-exchan
$\begin{aligned} & \text { Softener Product } \\ & \text { Flow } \\ & \mathrm{Ca} \& \mathrm{Mg}\end{aligned}$ $\begin{array}{lc}\text {-If using an ion-exchanger in a direct composit } \\ \begin{array}{l}\text { Softener Product } \\ \text { Flow }\end{array} & \text { M16 } \\ \text { Ca \& Mg } & \text { M18 \& M19 }\end{array}$ $\begin{array}{lc}\text {-If using an ion-exchanger in a direct composit } \\ \begin{array}{l}\text { Softener Product } \\ \text { Flow }\end{array} & \text { M16 } \\ \text { Ca \& Mg } & \text { M18 \& M19 }\end{array}$
M31
M32

N16
N18 \& N19 $\frac{\text { Title }}{\mathrm{C}_{\mathrm{T}}}$

- If using an ion-exchanger in
$\mathrm{Fe}-\mathrm{Cl} \& \mathrm{DOC}$
Na
pH
$\mathrm{CO}_{3}$
$\mathrm{HCO}_{3}$

$\mathrm{H}_{2} \mathrm{CO}_{3}$
$\mathrm{C}_{\mathrm{T}}$ Ion-X Blowdown
Flow
$\mathrm{Ca} \& \mathrm{Mg}$
$\mathrm{Fe}-\mathrm{PO}_{4} \& \mathrm{DOC}$
$\mathrm{Cl} \& \mathrm{Na}$
FNGP WATER SYSTEM OVERVIEW SPREADSHEET (con't)

Equations / values and explanation thereof

- No change in pH from the hot lime treater
$=\mathrm{CO}_{3}-\mathrm{HCO}_{3}$ equilibrium constant $\mathrm{x}\left[\mathrm{HCO}_{3}\right] /(10-\mathrm{pH})$ $=\left[\mathrm{C}_{\mathrm{T}}\right] \times 1 /\left(10-\mathrm{pH} / \mathrm{H}_{2} \mathrm{CO}_{3}\right.$ equilibrium constant $+1+\mathrm{CO}_{3}-\mathrm{HCO}_{3}$ equilibrium


## constant $/ 10-\mathrm{pH}$ ) - from Benefield et <br> - from Benefield et al. 1982 <br> $=(10-\mathrm{pH}) \times\left[\mathrm{HCO}_{3}\right] / \mathrm{H}_{2} \mathrm{CO}_{3}$ <br> $=\left[\mathrm{C}_{\mathrm{T}}\right]$ in softener product

- if a Nanofilter is used in place of an ion-exchanger
Title

| Title |
| :--- |
| pH |
| $\mathrm{CO}_{3}$ |
| $\mathrm{HCO}_{3}$ |
|  |
| $\mathrm{H}_{2} \mathrm{CO}_{3}$ |
| C | Softener Product

Flow - $\mathrm{C}_{\mathrm{T}}$
NF Blowdown
Flow - $\mathrm{C}_{\mathrm{T}}$

- if no softener is used in the treatment system:
M16-M32
N16-N32
Ion-X or NF Blowdown
Flow - $C_{T}$
Treated Water
Flow


## 016

without a softener
ii) softener product water flow if using a nanofilter
iii) (softener product water - ion-x blowdown) - if using an ion-exchanger
FNGP WATER SYSTEM OVERVIEW SPREADSHEET (con't)

$=$ ' $[$ FNGP water system]Blowdown calc.' $1 \$ \mathrm{Q} \$ 83$

pH
$\mathrm{CO}_{3}$
$\mathrm{HCO}_{3}$

$\mathrm{CO}_{3} \& \mathrm{HCO}_{3}$

Sulphur Plant Blowdown
Flow


$$
\begin{aligned}
& \text { = I16 } \\
& \text { - same value as cell I16 } \\
& \text { = I18: I27 } \\
& \text { - same value as cells I18 through I27, respectively } \\
& \text { = I 29: I32 } \\
& \text { - same value as cells I29 through I32, respectively }
\end{aligned}
$$ $=\mathrm{IF}(\mathrm{D} 14=$ "Composite", $\mathrm{IF}(\mathrm{K} 9=$ "yes", $\mathrm{IF}(\mathrm{O} 10=$ "Lime pond", $\mathrm{N} 16,0), 0), \mathrm{N} 16)$

- Structure of the if/then statements indicates that, regardless of the configuration of
the water treatment system, if a softener is used then the amount of blowdown
water reaching the lime pond is equal to the value of cell N16. If a water softener is
not included in the system, then this cell has a value of 0 .
- same basic format as cell F39, with the resulting value of each of these cells being
equal to 0 or N18 through N32, respectively
 F39

$$
\begin{aligned}
& \text { Ion - X Blowdown } \\
& \text { Flow }
\end{aligned}
$$

F41-F55

| Domestic Backwash \& Lab water <br> Flow |  |
| :--- | :---: |
| $\mathrm{Ca}-\mathrm{pH}$ | $\mathrm{G} 41-\mathrm{G} 50$ |
| $\mathrm{CO}_{3} \& \mathrm{HCO}_{3}$ | $\mathrm{G} 52-\mathrm{G} 53$ |
| $\mathrm{C}_{\mathrm{T}}$ | G 55 |


$\mathrm{Ca}-\mathrm{C}_{\mathrm{T}}$
${ }_{r}$

$$
\begin{aligned}
& =\mathrm{IF}(\mathrm{D} 14=\text { "Composite", } \mathrm{IF}(\mathrm{~K} 9=\text { "yes", } \mathrm{IF}(\mathrm{O} 10=\text { "Lime pond", N16, } 0), 0), \mathrm{N} 16) \\
& \text { - Structure of the if/then statements indicates that, regardless of the configuration of } \\
& \text { the water treatment system, if a softener is used then the amount of blowdown } \\
& \text { water reaching the lime pond is equal to the value of cell N16. If a water softener is } \\
& \text { not included in the system, then this cell has a value of } 0 \text {. } \\
& \text { - same basic format as cell F39, with the resulting value of each of these cells being } \\
& \text { equal to } 0 \text { or N18 through N32, respectively }
\end{aligned}
$$

$$
\begin{aligned}
& =(1.5+0.8) *(1000 / 10.01439) *(1.2 / 264.172) * 24 \\
& \text { - water volume in KLBH x conversion factors changing numbers into } \mathrm{m}^{3} / \text { day } \\
& =\text { O18:O27 } \\
& \text { - same value as cells } 018 \text { through } 027 \text {, respectively } \\
& =\text { O29:O27 } \\
& \text { - same value as cells } 029 \text { and } 030 \text {, respectively } \\
& =\mathrm{O} 32
\end{aligned}
$$


Title
$\underset{\text { Plow }}{\text { Ploals }}$
$\mathrm{Ca}-\mathrm{pH}$
$\mathrm{CO}_{3}-\mathrm{HCO}_{3}$
$\mathrm{C}_{\mathrm{T}}$
Net Inflow
Flow
$\mathrm{Ca}-\mathrm{pH}$
$\mathrm{CO}_{3}$
$\mathrm{HCO}_{3}$
$\mathrm{C}_{\mathrm{T}}$
Sources \& Sinks
$\mathrm{Ca}-\mathrm{C}_{\mathrm{T}}$

pH
$\mathrm{CO}_{3}$
$\mathrm{HCO}_{3}$

$\mathrm{HCO}_{3}$
SYSTEM OVERVIEW SPREADSHEET (con't)
Equations / values and explanation thereof
$=\left(\left(10^{\wedge}-\mathrm{L} 50\right)^{*} \mathrm{~L} 53\right) /\left(4.2^{*} 10^{\wedge}-7\right)$
$=\left(10^{-\mathrm{pH}}\right) \times\left[\mathrm{HCO}_{3}\right] / \mathrm{H}_{2} \mathrm{CO}_{3}-\mathrm{CO}_{3}$ equilibrium constant
$=\mathrm{Sum}\left(\mathrm{L}_{5}: \mathrm{L} 54\right)$
$=\left[\mathrm{CO}_{3}\right]+\left[\mathrm{HCO}_{3}\right]+\left[\mathrm{H}_{2} \mathrm{CO}_{3}\right]$


$=$ Specific values observed in mass balances constructed with the original water data $\begin{aligned} & \text { M39 } \\ = & \text { same value as cell M39 } \\ = & \text { IF(N41:N49= } "-", \text { M41:M49, }((\mathrm{M} 41: \mathrm{M} 49 * \mathrm{M} 39 * 1000) *(1+\mathrm{N} 41: \mathrm{N} 49)) / \\ & (\mathrm{O} 39 * 1000))\end{aligned}$ | Title | Cell(s) |
| :--- | :---: |
| $\mathrm{H}_{2} \mathrm{CO}_{3}$ | L 54 |
| $\mathrm{C}_{\mathrm{T}}$ | L 55 |
| Polishing Pond - Initial Feed |  |
| Flow |  |
| $\mathrm{Ca}-\mathrm{pH}$ | M 39 |
| $\mathrm{CO}_{3}$ | $\mathrm{M} 41-\mathrm{M} 50$ |
| $\mathrm{HCO}_{3}$ | M 52 |
|  |  |
| $\mathrm{H}_{2} \mathrm{CO}_{3}$ | M 53 |
| $\mathrm{C}_{\mathrm{T}}$ | M 54 |
|  | M 55 |
| Sources \& Sinks |  |
| Ca - C |  |

FNGP WATER SYSTEM OVERVIEW SPREADSHEET (con't)
 outgoing water volume. If there is no recognized gain or loss across the polishing pond, then the outgoing contaminant concentration is identical to incoming levels 0
1
1 - polishing pond effluent assumed to continually have a pH of 7.6
$=\left(\left(4.8 * 10^{\wedge}-11\right)^{*} \mathrm{O} 53\right) /\left(10^{\wedge}-\mathrm{O} 50\right)$ $=\mathrm{CO}_{3}-\mathrm{HCO}_{3}$ equilibrium constant $\mathrm{x}\left[\mathrm{HCO}_{3}\right] /(10-\mathrm{pH})$ $\left.=055^{*}\left(1 /\left(\left(10^{\wedge}-\mathrm{O} 50\right) /\left(4.2^{*} 10^{\wedge}-7\right)\right)+1+\left(\left(4.8^{*} 10^{\wedge}-11\right) /\left(10^{\wedge}-\mathrm{O} 50\right)\right)\right)\right)$ $=\left[\mathrm{C}_{\mathrm{T}}\right] \times 1 /\left(10-\mathrm{pH} / \mathrm{H}_{2} \mathrm{CO}_{3}\right.$ equilibrium constant $+1+\mathrm{CO}_{3}-\mathrm{HCO}_{3}$ equilibrium constant $/ 10-\mathrm{pH}$ )

- from Benefield et a
$=\left(\left(10^{\wedge}-\mathrm{O} 50\right)^{*} \mathrm{O} 53\right) /\left(4.2 * 10^{\wedge}-7\right)$
$=\left(10^{-} \mathrm{pH}\right) \times\left[\mathrm{HCO}_{3}\right] / \mathrm{H}_{2} \mathrm{CO}_{3}-\mathrm{HCO}_{3}$ equilibrium constant
$=$ M55
- same value as cell M55

O54 O55
Title
pH
$\mathrm{CO}_{3}$
$\mathrm{HCO}_{3}$

$\mathrm{H}_{2} \mathrm{CO}_{3}$
$\mathrm{C}_{\mathrm{T}}$
BLOWDOWN CALCULATION SPREADSHEET

| Title | Cell(s) | Equations/values and explanation thereof |
| :---: | :---: | :---: |
| CURRENT BOILER SYSTEM |  |  |
| -This top portion of the "Blowdown Calc." spreadsheet essentially details a mass balance on the boilers and sulphur plant to deter there are any sinks or sources not previously detected which need to be included in the computer simulator. |  |  |
| Boiler Feedwater Ouality |  |  |
| Treated Water |  |  |
| Flow | C8 | $\begin{aligned} & =74.2 *(1000 / 10.1439) *(1.2 / 264.172) * 24 \\ & =\text { Flow expressed in KLBH converted to } \mathrm{m}^{3} / \text { day } \end{aligned}$ |
| TDS - T. Alk | C10-C21 | = Specific values from original data set |
| $\mathrm{CO}_{3}$ | C22 | $\begin{aligned} &=\left(50000 *\left(\mathrm{C} 21 / 50000+\left(10^{\wedge}-\mathrm{C} 20\right)-\left(10^{\wedge}-14 / 10^{\wedge}-\mathrm{C} 20\right)\right) /\left(1+\left(10^{\wedge}-\mathrm{C} 20 /\left(2 * 4.7 * 10^{\wedge}-\right.\right.\right.\right. \\ &11)))) * 0.6 / 60000 \\ &= 50000 \times\left(\mathrm{T} . \mathrm{Alk} / 50000+10-\mathrm{pH}-10^{-14} / 10^{-}-\mathrm{pH}\right) /\left(1+10^{-} \mathrm{pH}_{/} /\left(2 \times \mathrm{H}_{2} \mathrm{CO}_{3}-\mathrm{CO}_{3}\right.\right. \\ &\text { equilibrium constant })) \times 0.6 / 60000 \\ &- \text { From Sawyer \& McCarty } 1978 \end{aligned}$ |
| $\mathrm{HCO}_{3}$ | C23 | $\begin{aligned} = & \left(50000^{*}\left(\mathrm{C} 21 / 50000+\left(10^{\wedge}-\mathrm{C} 20\right)-\left(10^{\wedge}-14 / 10^{\wedge}-\mathrm{C} 20\right)\right) /\left(1+\left(\left(2^{*} 4.7^{*} 10^{\wedge}-11\right) / 10^{\wedge}-\right.\right.\right. \\ & \mathrm{C} 20)))^{*} 1.22 / 61000 \\ = & 50000 \times\left(\mathrm{T} . \mathrm{Alk} / 50000+10^{-} \mathrm{pH}-10^{-14} / 10^{-} \mathrm{pH}\right) /\left(1+\left(2 \times \mathrm{H}_{2} \mathrm{CO}_{3}-\mathrm{CO}_{3}\right.\right. \\ & \text { equilibrium constant } \left.) / 10^{-\mathrm{pH}}\right) \times 1.22 / 61000 \end{aligned}$ |
| $\mathrm{H}_{2} \mathrm{CO}_{3}$ | C24 | $\begin{aligned} & =\left(\left(10^{\wedge}-\mathrm{C} 20\right)^{*} \mathrm{C} 23\right) /\left(4.2^{*} 10^{\wedge}-7\right) \\ & =(10-\mathrm{pH}) \times\left[\mathrm{HCO}_{3}\right] / \mathrm{H}_{2} \mathrm{CO}_{3}-\mathrm{HCO}_{3} \text { equilibrium constant } \end{aligned}$ |
| $\mathrm{C}_{\mathrm{T}}$ | C25 | $\begin{aligned} & =\operatorname{Sum}(\mathrm{C} 22: \mathrm{C} 24) \\ & =\left[\mathrm{CO}_{3}\right]+\left[\mathrm{HCO}_{3}\right]+\left[\mathrm{H}_{2} \mathrm{CO}_{3}\right] \end{aligned}$ |
| Condensate Return |  |  |
| Flow | D8 | $\begin{aligned} & =1054.2 *(1000 / 10.01439) *(1.2 / 264.172) * 24 \\ & =\text { Flow as KLBH converted to } \mathrm{m}^{3} / \text { day } \end{aligned}$ |
| TDS - T. Alk | D10-D21 | $=$ Specific values from original data set |
| $\mathrm{CO}_{3}$ | D22 | $\begin{aligned} & =\left(50000^{*}\left(\mathrm{D} 21 / 50000+\left(10^{\wedge}-\mathrm{D} 20\right)-\left(10^{\wedge}-14 / 10^{\wedge}-\mathrm{D} 20\right)\right) /\left(1+\left(10^{\wedge}-\mathrm{D} 20 /\left(2 * 4.7^{*} 10^{\wedge}-\right.\right.\right.\right. \\ & 11)))) * 0.6 / 60000 \end{aligned}$ |

BLOWDOWN CALCULATION SPREADSHEET (con't)

| Cell(s) | Equations / values and explanation thereof |
| :---: | :---: |
|  | $\begin{aligned} & =50000 \times\left(\mathrm{T} . \mathrm{Alk} / 50000+10-\mathrm{pH}-10^{-14} / 10^{-\mathrm{pH}}\right) /\left(1+\left(2 \mathrm{H}_{2} \mathrm{CO}_{3}-\mathrm{CO}_{3}\right.\right. \\ & \text { equilibrium constant }) \times 0.6 / 60000 \\ & \text {-From Sawyer \& McCarty } 1978 \end{aligned}$ |
| D23 | $\begin{aligned} & =\left(50000 *\left(\mathrm{D} 21 / 50000+\left(10^{\wedge}-\mathrm{D} 20\right)-\left(10^{\wedge}-14 / 10^{\wedge}-\mathrm{D} 20\right)\right) /\left(1+\left(\left(2^{*} 4.7^{*} 10^{\wedge}-11\right) / 10^{\wedge}-\right.\right.\right. \\ & \mathrm{D} 20))) * 1.22 / 61000 \\ & =50000 \times\left(\mathrm{T} . \mathrm{Alk} / 50000+10^{-\mathrm{pH}}-10^{-14} / 10^{-\mathrm{pH}}\right) /\left(1+\left(2 \times \mathrm{H}_{2} \mathrm{CO}_{3}-\mathrm{CO}_{3}\right.\right. \\ & \text { equilibrium constant) } / 10-\mathrm{pH}) \times 1.22 / 61000 \\ & \text {-From Sawyer \& McCarty } 1978 \end{aligned}$ |
| D24 | $\begin{aligned} & =\left(\left(10^{\wedge}-\mathrm{D} 20\right) * \mathrm{D} 23\right) /\left(4.2^{*} 10^{\wedge}-7\right) \\ & =\left(10^{-\mathrm{pH}}\right) \times\left[\mathrm{HCO}_{3}\right] / \mathrm{H}_{2} \mathrm{CO}_{3}-\mathrm{HCO}_{3} \text { equilibrium constant } \end{aligned}$ |
| D25 | $\begin{aligned} & =\text { Sum (D22: D24) } \\ & =\left[\mathrm{CO}_{3}\right]+\left[\mathrm{HCO}_{3}\right]+\left[\mathrm{H}_{2} \mathrm{CO}_{3}\right] \end{aligned}$ |


| E8 | $\begin{aligned} & =\mathrm{D} 8+\mathrm{C} 8 \\ & =\text { Condensate return }+ \text { treated water } \end{aligned}$ |
| :---: | :---: |
| E10-E20 | $\begin{aligned} & =(\mathrm{C} 10: \mathrm{C} 20 * \mathrm{C} 8+\mathrm{D} 10: \mathrm{D} 20 * \mathrm{D} 8) / \mathrm{E} 8 \\ & =[] \mathrm{BFW}=\text { (mass in condensate }+ \text { mass in treated water }) / \text { boiler feedwater return } \end{aligned}$ |
| E22 | $=\left(\left(4.8^{*} 10^{\wedge}-11\right)^{* E 23}\right) /\left(10^{\wedge}-\mathrm{E} 20\right)$ <br> $=\mathrm{CO}_{3}-\mathrm{HCO}_{3}$ equilibrium constant $\mathrm{x}\left[\mathrm{HCO}_{3}\right] /(10-\mathrm{pH})$ |
| E23 | $\begin{aligned} & \left.=\mathrm{E} 25^{*}\left(1 /\left(\left(\left(10^{\wedge}-\mathrm{E} 20\right)\right) /\left(4.2^{*} 10^{\wedge}-7\right)\right)+1+\left(\left(4.8^{*} 10^{\wedge}-11\right) /\left(10^{\wedge}-\mathrm{E} 20\right)\right)\right)\right) \\ & =\left[\mathrm{C}_{\mathrm{T}}\right] \times 1 /\left(10^{-\mathrm{pH}} / \mathrm{H}_{2} \mathrm{CO}_{3} \text { equilibrium constant }+1+\mathrm{CO}_{3}-\mathrm{HCO}_{3}\right. \text { equilibrium } \\ & \text { constant } / 0-\mathrm{pH}) \\ & \text {-From Benefield et al. } 1982 \end{aligned}$ |
| E25 | $\begin{aligned} & =(\mathrm{C} 25 * \mathrm{C} 8+\mathrm{D} 25 * \mathrm{D} 8) / \mathrm{E} 8 \\ & - \text { Same format as cells E10 to E20; [ ] BFW }=\text { (mass in condensate }+ \text { mass in treated } \\ & \text { water) } / \text { boiler feedwater flow } \end{aligned}$ |

Title
$\mathrm{HCO}_{3}$
$\mathrm{H}_{2} \mathrm{CO}_{3}$
$\mathrm{C}_{\mathrm{T}}$
$\mathbf{B}_{\text {Boiler Feedwater (BFW) }}^{\text {Flow }}$
TDS - pH
$\mathrm{CO}_{3}$
$\mathrm{HCO}_{3}$
$\mathrm{C}_{\mathrm{T}}$
BLOWDOWN CALCULATION SPREADSHEET (con't)

| Cell(s) |
| :---: |
| C 36 |
| $\mathrm{C} 38-\mathrm{C} 52$ |

 E36-E48
E49
 E50 equilibrium constant) / $10-\mathrm{pH}) \times 1.22 / 61000$
-From Sawyer \& McCarty 1978

$$
=\text { Specific values from original data set }
$$


$=50000 \times\left(\mathrm{T} . \mathrm{Ak} / 50000+10^{-\mathrm{pH}}-10^{-14} / 10-\mathrm{PH}\right) /\left(1+10^{-\mathrm{PH} / 2} \times \mathrm{H}_{2} \mathrm{CO}_{3}-\mathrm{CO}_{3}\right.$
$=(10-\mathrm{pH}) \times\left[\mathrm{HCO}_{3}\right] / \mathrm{H}_{2} \mathrm{CO}_{3}-\mathrm{HCO}_{3}$ equilibrium constant
$=\left[\mathrm{CO}_{3}\right]+\left[\mathrm{HCO}_{3}\right]+\left[\mathrm{H}_{2} \mathrm{CO}_{3}\right]$
 N $\frac{\text { Title }}{\text { Boilers - Mass Balance }}$
BFW
Flow
$\mathrm{Ca}-\mathrm{C}_{\mathrm{T}}$



450 psi Steam Flow - T.Alk
$\mathrm{CO}_{3}$
$\mathrm{HCO}_{3}$
$\mathrm{H}_{2} \mathrm{CO}_{3}$
$\mathrm{C}_{\mathrm{T}}$



| BLOWDOWN CALCULATION SPREADSHEET (con't) |  |
| :---: | :---: |
| Cell(s) | Equations / values and explanation thereof |
| M50 | $\begin{aligned} & =50000 \times\left(\mathrm{T} . \mathrm{Alk} / 50000+10-\mathrm{pH}-10^{-14} / 10^{-\mathrm{pH}}\right) /\left(1+\left(2 \times \mathrm{H}_{2} \mathrm{CO}_{3}-\mathrm{CO}_{3}\right.\right. \\ & \text { equilibrium constant }) / 10-\mathrm{pH}) \times 1.22 / 61000 \\ & \text {-From Sawyer \& McCarty } 1978 \end{aligned}$ |
| M51 | $=(10-\mathrm{pH}) \times\left[\mathrm{HCO}_{3}\right] / \mathrm{H}_{2} \mathrm{CO}_{3}-\mathrm{HCO}_{3}$ equilibrium constant |
| M52 | $=\left[\mathrm{CO}_{3}\right]+\left[\mathrm{HCO}_{3}\right]+\left[\mathrm{H}_{2} \mathrm{CO}_{3}\right]$ |
| O36 | $\begin{aligned} & =\text { Sum (L36 : M36) } \\ & =450 \text { psi steam }+ \text { BFW flows } \end{aligned}$ |
| O38-O52 | $\begin{aligned} & =(\text { mass in BFW }+ \text { mass in } 450 \text { psi steam }) / \text { net inflow volume }(\text { e.g. }= \\ & \left.\left.\left(\mathrm{L} 38: \mathrm{L} 46^{*} \mathrm{~L} 36+\mathrm{M} 38: \mathrm{M} 46^{*} \mathrm{M} 36\right) / \mathrm{O} 36\right)\right) \end{aligned}$ |
| P36-P48 | $=$ Specific values from original data set |
| P49 | $\begin{aligned} & =50000 \times\left(\mathrm{T} . \mathrm{Alk} / 50000+10-\mathrm{pH}-10^{-14} / 10^{-}-\mathrm{pH}\right) /\left(1+10-\mathrm{pH} /\left(2 \times \mathrm{H}_{2} \mathrm{CO}_{3}-\mathrm{CO}_{3}\right.\right. \\ & \quad \text { equilibrium constant })) \times 0.6 / 60000 \\ & - \text { From Sawyer \& McCarty } 1978 \end{aligned}$ |
| P50 | $\begin{aligned} & =50000 \times\left(\mathrm{T} . \mathrm{Alk} / 50000+10-\mathrm{pH}-10^{-14} / 10-\mathrm{pH}\right) /\left(1+\left(2 \times \mathrm{H}_{2} \mathrm{CO}_{3}-\mathrm{CO}_{3}\right.\right. \\ & \text { equilibrium constant) } / 10-\mathrm{pH}) \times 1.22 / 61000 \\ & \text {-From Sawyer \& McCarty } 1978 \end{aligned}$ |
| P51 | $=(10-\mathrm{pH}) \times\left[\mathrm{HCO}_{3}\right] / \mathrm{H}_{2} \mathrm{CO}_{3}-\mathrm{HCO}_{3}$ equilibrium constant |
| P52 | $=\left[\mathrm{CO}_{3}\right]+\left[\mathrm{HCO}_{3}\right]+\left[\mathrm{H}_{2} \mathrm{CO}_{3}\right]$ |
| Q36 | $\begin{aligned} & =(1.8+142.7+71.9+0.1) *(1000 / 10.01439) *(1.2 / 264.172) * 24 \\ & =\text { Total volume of } 45 \text { and } 150 \text { psi steam venting and condensate exiting the sulphur } \\ & \text { plant } \end{aligned}$ |
| Q38-Q48 | $=$ Specific values from original data set |

Title
$\mathrm{HCO}_{3}$
$\mathrm{H}_{2} \mathrm{CO}_{3}$
$\mathrm{C}_{\mathrm{T}}$
Net Inflow
Flow
$\mathrm{Ca}-\mathrm{C}_{\mathrm{T}}$
Blowdown
Flow - T.Alk
$\mathrm{CO}_{3}$
$\mathrm{HCO}_{3}$
$\mathrm{H}_{2} \mathrm{CO}_{3}$
$\mathrm{C}_{\mathrm{T}}$
Steam and Condensate
Flow
Ca - T.Alk

BLOWDOWN CALCULATION SPREADSHEET (con't) Cell(s)
PREDICTED BLOWDOWN QUALITY

- this section of the "Blowdown Calc." spreadsheet is used to predict boiler and sulphur plant blowdown characteristics based on BFW quality and any sources and sinks detected above.
Boiler Feedwater (BFW)
Treated Water
Flow

Condensate Return
Flow
$\mathrm{Ca}-\mathrm{C}_{\mathrm{T}}$
$\underset{\text { Flow }}{\text { BFW }}$
$\mathrm{Ca}-\mathrm{pH}$
$\mathrm{CO}_{3}$
$\mathrm{HCO}_{3}$
BLOWDOWN CALCULATION SPREADSHEET (con't)
$\stackrel{3}{0}$


## -Same format as cells E62 through E72 $=$ (mass in treated water + mass in cond

$=\left[\mathrm{C}_{\mathrm{T}}\right] \times 1 /\left(10-\mathrm{pH} / \mathrm{H}_{2} \mathrm{CO}_{3}\right.$ equilibrium constant $+1+\mathrm{CO}_{3}-\mathrm{HCO}_{3}$ equilibrium
constant $/ 10-\mathrm{pH})$
-From Benefield et al. 1982
$=(\mathrm{C} 77 * \mathrm{C} 61+\mathrm{D} 77 * \mathrm{D} 61) /(\mathrm{E} 61)$
-Same format as cells E62 through E72
$=\left[\mathrm{C}_{\mathrm{T}}\right] \times 1 /\left(10-\mathrm{pH} / \mathrm{H}_{2} \mathrm{CO}_{3}\right.$ equilibrium constant $+1+\mathrm{CO}_{3}-\mathrm{HCO}_{3}$ equilibrium
constant $/ 10-\mathrm{pH})$
-From Benefield et al. 1982
$=(\mathrm{C} 77 * \mathrm{C} 61+\mathrm{D} 77 * \mathrm{D} 61) /(\mathrm{E} 61)$
-Same format as cells E62 through E72
$=$ (mass in treated water + mass in condensate return) / BFW volume

## $=\mathrm{F} 83+\mathrm{G} 83$

$=450$ psi steam + predicted blowdown flows
$=$ Values of cells E63 through E77, respectively
-Sources and sinks detected above with a value greater than $15 \%$

$=$ Specific value from the original data set
-Based on assumption that 450 psi steam production remains constant over time
=-IF(E85•E93<D11•D19.E85.E93,D11•D19)
-Specifies that the 450 psi steam generated in the boilers will be of equal or better quality than condensate collected on the FNGP -Based on the assumption that if the BFW becomes cleaner than current levels, the quality of the 450 psi steam will similarly improve $=\mathrm{IF}(\mathrm{E} 94<\mathrm{D} 20,8.7, \mathrm{D} 20)$
-To prevent pipe corrosion and/or steam system contamination
 F83
F85-F93


Title
$\mathrm{CO}_{3}$
$\mathrm{HCO}_{3}$
$\mathrm{C}_{\mathrm{T}}$
Predicted Blowdown
Flow
$\mathrm{Ca}-\mathrm{DOC}$
pH
CO 3
HCO
3
BLOWDOWN CALCULATION SPREADSHEET (con't)
$\frac{\mathrm{Cell(s)}}{\mathrm{M} 85-\mathrm{M} 99}$
N85-N99
O83
O85-O93

$\underset{\text { Flow }}{\text { Steam }} \boldsymbol{\&}$ Condensate
$\mathrm{Ca}-\mathrm{DOC}$

88
P83
P85-P93
$=\mathrm{IF}(\mathrm{L} 94<\mathrm{D} 20,8.7$, D20)
-Specifies that steam pH will be 8.7 or greater
-Using same assumption as cell F94
$=\mathrm{IF}\left(\mathrm{N} 99={ }^{\prime}-\mathrm{l}\right.$, ( L
$(1+\mathrm{N} 99)) / \mathrm{O} 83)$

- Same format as ce
(1+ \% gain or lo
$=(1.8+142.7+71.9+0.1) *(1000 / 10.01439) *(1.2 / 264.172) * 24$
$=$ Sum of condensate, $150 \& 45$ psi steam and vented water leaving from the sulphur
plant
$=\mathrm{IF}(\mathrm{E} 85: \mathrm{E} 93<\mathrm{D} 11: \mathrm{D} 19, \mathrm{E} 85: \mathrm{E} 93$, D11:D19)
-Specifies that the steam and condensate generated in the sulphur plant will be of
equal or better quality than condensate collected from the FNGP
- Based on assumption that if BFW quality improves beyond current standards, so too
will the plant's steam and condensate will the plant's steam and condensate $=8.7$
-Assum
-Assume that pH kept constant at 8.7
$=($ mass in 450 psi steam + mass in BFW $) \times(1+\%$ gain or loss across the sulphur plant) / total incoming water value)
$=\mathrm{M} 83+\mathrm{L} 83$
$=450$ psi steam
$=450$ psi steam + BFW flows
$=\operatorname{IF}($ N85:N93 $="-$ ", (L85:L93
$=$ IF(N85:N93 $="-$ ", (L85:L93*L83+M85:M93*M83)/O83, $((\mathrm{L} 85: \mathrm{L} 93 * \mathrm{~L} 83+$
M85:M93*M83)*(1+N85:N93))/O83)

$=$ Sources and sinks detected above with a value greater than $15 \%$
Equations / values and explanation thereof
$=$ Exact same formulas as used for the 450 psi steam produced from the boilers (i.e.
cells F85-F99)

- $1.8+142.7+71.9+0.1) *(1000 / 10.01439) *(1.2 / 264.172) * 24$
品
Title
$\mathrm{Ca}_{\mathrm{C}} \mathrm{C}_{\mathrm{T}}$

| Sources and Sinks |
| :---: |
| $\mathrm{Ca}-\mathrm{C}_{\mathrm{T}}$ |
| Net Inflow |
| Flow |
| $\mathrm{Ca}-$ DOC |
| pH |
| $\mathrm{C}_{\mathrm{T}}$ |
| Steam \& Condensate |
| Flow |
| $\mathrm{Ca}-\mathrm{DOC}$ |
| pH |

\[

\]

LIME TREATER SPREADSHEET
Lime Treater Output - Using Present FNGP System
-This half of the computer worksheet uses information from the Water Treatment Manual and data collected at the plant to estimate some of the chemical characteristics of the lime treater's product and blowdown waters. Given the open and easy-to-follow layout of this spreadsheet, no further explanation of the equations used herein was felt to be required.

## Predicted Lime Treater Products <br> 

[^11]hardness contained in the ENGs original blowdow
-This portion of the "Lime Blowdown ,
$=(\mathrm{K} 79 * 16.3) /(232.1-(5+106.3))$
$=($ Hardness removed $*$ original blowdown flow at the FNGP $) /$ total mass of
hardness contained in the FNGP's original blowdown flow

.


Influent Water
Hardness

## T. Alk


(K79*16.3) $(2321-(5+1063))$

  |  |
| :---: |
|  |
|  |
|  |
|  |
|  |
|  |
|  |
| 131 |

 $\underset{\text { Flow Treater Blowdown }}{\text { Lime }}$ $\underset{\text { Flow Treater Blowdown }}{\text { Lime }}$

-
SOFTENER SPREADSHEET

- As with "Lime Treater Calc." worksheet, the calculations used to predict concentration of chloride and sodium in both the ion-exchange
blowdown and product water flows, as well as volume of blowdown water leaving the ion-exchangers, are, for the most part, explained
within the worksheet itself. No further explanations are necessary.
NANOFILTER SPREADSHEET
Equations / values and explanation thereof
$=$ variable open to manipulation with minimum and maximum values of 1 and 3 ,
respectively
= variable open to manipulation; always kept at $75 \%$ through this study
= variable open to manipulation $(0-75 \%)$; changed from scenario to scenario
$=$ variable open to manipulation $(0-75 \%)$; changed from scenario to scenario
$=$ variable open to manipulation (on or off); activated mainly when using a direct
composite discharge reuse scenario
$=$ variable open to manipulation (on or off); whenever a 3 stage NF unit was used,
this option was activated
$=$ variable open to manipulation $(0-100) ;$ when brine from the 2 nd NF unit was
recycled back to the 1 st NF filter, the volume of brine going back to the 1 st stage
varied from scenario to scenario
$=$ variable open to manipulation $(0-100) ;$ when using a 3 stage NF system, all of the
brine from the 3rd unit was recycled back to the 1 st NF filter (i.e. F6 = 100)

| Title | $\mathrm{Cell}(\mathrm{s})$ |
| :---: | :---: |
| CONSTANTS |  |
| Number of NF stages | D10 |
| Water recovery ratios |  |
| Unit 1 | D4 |
| Unit 2 | E4 |
| Unit 3 | F4 |
| Brine recycling |  |
| Unit 2 | E5 |
| Unit 3 | F5 |
| \% recycle |  |
| Unit 3 | F6 |
| Blending | F7 \& F8 |
| Salt rejection efficiencies Monovalents | 14 |
| Divalents | 15 |
| Organics | 16 |
| Water temperature | 18 |

NANOFILTER SPREADSHEET (con't)

Equations / values and explanation thereof
$=$ used to adjust inlet calcium concentration to avoid salt precipitation within the NF
membranes
Same equations as used in RO worksheet, cells K6 - N6
$=10^{\wedge}-(17052 /(273+\mathrm{I} 8)+215.21 * \log 10(273+\mathrm{I} 8)-0.12675 *(273+\mathrm{I} 8)-$
$545.56)$

- same equation as cell I6 in the "Reverse osmosis" spreadsheet
$=10^{\wedge}-(2902.39 /(273+\mathrm{I} 8)+0.02379 *(273+\mathrm{I} 8)-6.498)$
- $\operatorname{same}$ equation as cell J6 in the "Reverse osmosis" spreadsheet
$=10^{\wedge}-(4787.3 /(273+\mathrm{I} 8)+7.1321 * \log 10(273+\mathrm{I} 8)+0.010365 *(273+\mathrm{I} 8)-$
$22.801)$

| INFLUENT |  |  |
| :---: | :---: | :---: |
| Raw Influent |  |  |
| - if using a NF softener within a back-end ZED model: |  |  |
| Flow | D14 | $\begin{aligned} & =\text { '[FNGP water system] overview'!\$J\$16 } \\ & =\text { hot lime treater outflow } \end{aligned}$ |
| Ca | D16 | $=$ '[FNGP water system] overview'!\$J\$18-C16 <br> $=[\mathrm{Ca}]$ in hot lime treater product water - the required Ca adjustment (if necessary) |
| Ca adjustment | C16 | $=$ '[FNGP water system] overview'!\$J\$18-40 <br> $=[\mathrm{Ca}]$ in hot lime treater product water - desired $[\mathrm{Ca}]$ required to prevent calcium salt precipitation within the nanofilters |
| $\mathrm{Mg}-\mathrm{C}_{\text {T }}$ | D17-D33 | $=$ '[FNGP water system] overview'!\$J\$19:\$J\$32 <br> $=$ contaminant []'s in the hot lime treater product water |
| - if using a NF softener as part of a direct composite discharge scenario: |  |  |
| Flow | D14 | $=$ '[FNGP water system] overview'!\$K\$16 <br> = reverse osmosis product flow |
| Ca | D16 | $=$ = [FNGP water system] overview'!\$K\$18-C16 <br> $=[\mathrm{Ca}]$ in RO product water outflow - the required Ca adjustment (if needed) |
| Ca adjustment | C16 | $=$ '[FNGP water system] overview'!\$K\$18-40 |

## NANOFILTER SPREADSHEET (con't)

 $=\operatorname{sum}(\mathrm{D} 14: \mathrm{F} 14)$
$=$ raw influent flow + recycled brine from units \#2 and \#3
$=(\mathrm{D} 16: \mathrm{D} 24 *$ D14 + E16: E24 * E14 + F16: F24 * F14)/G14
$=$ (mass in raw water inflow + mass in brine from units \#2 and \#3)/ total influent water volume $=(\mathrm{D} 25 * \mathrm{D} 14+\mathrm{E} 25 * \mathrm{E} 14+\mathrm{F} 25 * \mathrm{~F} 14) / \mathrm{G} 14-\mathrm{C} 24$
$=(\mathrm{pH}$ of raw influent $*$ raw influent flow +pH of brine
$=$ (pH of raw influent * raw influent flow +pH of brine from unit \#2 * unit \#2 brine flow +pH of brine from unit \#3 $*$ unit \#3 brine flow)/ total influent volume -
required pH adjustment
$=(\mathrm{D} 25 * \mathrm{D} 14+\mathrm{E} 25 * \mathrm{E} 14+\mathrm{F} 25 * \mathrm{~F} 14) / \mathrm{G} 14-6.5$
 precipitation within the membranes

Title
$\mathrm{Mg}-\mathrm{C}_{\mathrm{T}}$
Brine Recycle - Unit \#2
Flow
$\mathrm{Ca}-\mathrm{C}_{\mathrm{T}}$
Brine Recycle - Unit \#3
Flow
Ca - $\mathrm{C}_{\mathrm{T}}$
NF Unit \#1
Feed
Flow
Ca - DOC
pH
pH adjustment
NANOFILTER SPREADSHEET (con't)

NANOFILTER SPREADSHEET (con't)
Equations / values and explanation thereof



| Cell(s) |
| :---: |
|  |
| H31 |
| H32 |
| H33 |
| H34 |
| H35 |



## EXPLANATION:

Benefield et. al. (1982) indicates that most carbon dioxide dissolved in water does not hydrolyse to carbonic acid. The equilibrium equations used in this model to determine carbonic acid concentrations (e.x. cell D32) assume, however, that the concentration of gases; they will pass unhindered through the filters. Rather than trying to calculate the proportion of carbon dioxide existing on its own and as carbonic acid, all carbonic acid within the RO feedwaters was assumed to be dissolved carbon dioxide. It therefore freely passes though the membranes. RO concentrate and permeate waters will, as a result, have identical $\mathrm{H}_{2} \mathrm{CO}_{3}$ concentrations.
Carbonate and bicarbonate ions will not pass unhindered through the RO filters, due to the limited permeability of RO membranes to monovalent and divalent ions. While permeate and concentrate streams will have identical concentrations of $\mathrm{H}_{2} \mathrm{CO}_{3}$, they will contain different amounts of carbonate and bicarbonate. Neither system will be in equilibrium. There will either be too much or "too litle" carbonate and bicarbonate. Concentrations within each stream will shift until carbonate and bicarbonate equilibriums are re-established. H2CO3- HCO 3 Equilibrium

$$
\begin{array}{rc} 
& {\left[\mathrm{H}_{2} \mathrm{CO}_{3}\right]} \\
\text { initial conditions } & z \\
\text { at equilibrium } & z-b
\end{array}
$$

Now,
NANOFILTER SPREADSHEET (con't)

NANOFILTER SPREADSHEET (con't)

| Title | Cell(s) | Equations / values and explanation thereof |
| :---: | :---: | :---: |
|  |  | = (mass in feed flow * \% monovalent ion rejection)/product flow |
| DOC | I24 | $\begin{aligned} & =\mathrm{G} 14 *(1-\mathrm{I} 6 / 100) * \mathrm{G} 24 / / 14 \\ & =(\text { mass in feed flow } * \% \text { DOC rejection }) / \text { product flow } \end{aligned}$ |
| pH |  |  |
| - equilibrium | I25 | $=-\log 10\left(\left(10^{\wedge}-\mathrm{I} 26\right)+\mathrm{I} 35+\mathrm{I} 34\right)$ |
| - initial | I26 | $=\mathrm{G} 25$ |
|  |  | - see explanation below |
| $\mathrm{CO}_{3}$ |  |  |
| - equlibrium | I27 | $=\mathrm{I} 28+\mathrm{I} 35$ |
| - initial | I28 | $=\mathrm{G} 14 *(1-\mathrm{I} 5 / 100) * \mathrm{G} 27 / \mathrm{I} 14$ <br> - see explanation below |
| $\mathrm{HCO}_{3}$ |  |  |
| - equilibrium | I29 | $=\mathrm{I} 30-\mathrm{I} 35+\mathrm{I} 34$ |
| - initial | I30 | $=\mathrm{G} 14 *(1-\mathrm{I} 4 / 100) * \mathrm{G} 29 / \mathrm{I} 14$ |
| $\mathrm{H}_{2} \mathrm{CO}_{3}$ |  |  |
| - equilibrium | 131 | = I32-I34 |
| - initial | I32 | $\begin{aligned} & =\mathrm{G} 31 \\ & \text { - see explanation below } \end{aligned}$ |
| $\mathrm{C}_{\text {T }}$ | 133 | $\begin{aligned} & =\mathrm{I} 28+\mathrm{I} 30+\mathrm{I} 32 \\ & =\text { initial }\left[\mathrm{CO}_{3}\right]+\text { initial }\left[\mathrm{HCO}_{3}\right]+\text { initial }\left[\mathrm{H}_{2} \mathrm{CO}_{3}\right] \end{aligned}$ |
| [] shift $\left(\mathrm{H}_{2} \mathrm{CO}_{3}-\mathrm{HCO}_{3}\right)$ | I34 | $\begin{aligned} & =\left(-\left(\mathrm{I} 30+10^{\wedge}-\mathrm{I} 26+\mathrm{O} 5\right)+\left(\left(\mathrm{I} 30+10^{\wedge}-\mathrm{I} 26+\mathrm{O}\right)^{\wedge} 2-4 *\left(-\mathrm{O} 5 * \mathrm{I} 32+\left(\mathrm{I} 30 * 10^{\wedge}\right.\right.\right.\right. \\ & \\ & \left.- \text { - } 266)))^{\wedge} 0.5\right) / 2 \\ & - \text { see explanation below } \end{aligned}$ |
| $\square]$ shift ( $\mathrm{HCO}_{3}-\mathrm{CO}_{3}$ ) | 135 | $\begin{aligned} = & \left(-\left(\mathrm{I} 28+\mathrm{I} 34+10^{\wedge}-\mathrm{I} 26+\mathrm{O} 6\right)+\left(\left(\mathrm{I} 28+\mathrm{I} 34+10^{\wedge}-\mathrm{I} 26+\mathrm{O} 6\right)^{\wedge} 2-4 *(-\mathrm{O} 6 *(\mathrm{I} 30\right.\right. \\ & \left.\left.\left.+\mathrm{I} 34)+\mathrm{I} 28 *\left(10^{\wedge}-\mathrm{I} 26+\mathrm{I} 34\right)\right)\right)^{\wedge} 0.5\right) / 2 \end{aligned}$ |
| EXPLANATION: |  |  |
| - as previously explained with the product stream, the concentration of each carbonate species will shift from its initial level in ord re-establish carbonate equilibriums. Although the equilibrium calculations used for the permeate still hold true for the brine flow, values of certain variables had changed: |  |  |
| y is now $=\%$ monovalent ion rejection * feedwater [HCO3] <br> w is now $=\%$ divalent ion rejection $*$ feedwater [CO3] |  |  |

NANOFILTER SPREADSHEET (con't)

| Title | Cell(s) | Equations / values and explanation thereof |
| :---: | :---: | :---: |
| Salt Precipitation Check $\mathrm{CaSO}_{4}$ | J16 | $\begin{aligned} = & \text { if }\left(\mathrm{I} 16 /(1000 * 40) *(\mathrm{I} 19 /(1000 * 96))>\mathrm{L} 4, \text { " } \mathrm{CaSO}_{4} ", \text { "OK" }\right) \\ = & \text { if }[\mathrm{Ca}] *\left[\mathrm{SO}_{4}\right]>\mathrm{CaSO}_{4} \text { equilibrium constant, then } \mathrm{CaSO}_{4} \text { may be precipitating } \\ & \text { within the NF membranes } \end{aligned}$ |
| $\mathrm{CaCO}_{3}$ | J17 | $\begin{aligned} & =\operatorname{if(}(\mathrm{I} 16 /(1000 * 40) * \mathrm{I} 27>\mathrm{L} 5, \text { "CaCO3", "OK") } \\ & =\text { if }[\mathrm{Ca}] *\left[\mathrm{CO}_{3}\right]>\mathrm{CaCO}_{3} \text { equilibrium constant, then } \mathrm{CaCO}_{3} \text { may be precipitating } \\ & \text { within the NF membranes } \end{aligned}$ |
| $\mathrm{MgCO}_{3}$ | J18 | $\begin{aligned} & =\text { if }(\mathrm{I} 17 /(1000 * 24.3) * \mathrm{I} 27>\mathrm{L} 6, \text { " } \mathrm{MgCO} 3 ", \text { "OK") } \\ & =\text { if }[\mathrm{Mg}] *\left[\mathrm{CO}_{3}\right]>\mathrm{MgCO}_{3} \text { equilibrium constant, then } \mathrm{MgCO}_{3} \text { may be precipitating } \\ & \text { within the } \mathrm{NF} \text { membranes } \end{aligned}$ |
| $\mathrm{Mg}(\mathrm{OH})_{2}$ | J19 | $=\mathrm{if}\left(\mathrm{I} 17 /(1000 * 24.3) *\left(\mathrm{O} 7 /\left(10^{\wedge}-\mathrm{G} 25\right)\right)^{\wedge} 2>\mathrm{L} 7, " \mathrm{Mg}(\mathrm{OH})_{2} ",\right. \text { "OK") }$ <br> $=$ if $[\mathrm{Mg}] *[\mathrm{OH}]^{2}>\mathrm{Mg}(\mathrm{OH})_{2}$ equilibrium constant, then $\mathrm{Mg}(\mathrm{OH})_{2}$ may be precipitating within the NF membranes |
| Unit \#2 . |  |  |
| - if $>1 \mathrm{NF}$ unit, is used in the treatment system:Feed |  |  |
|  |  |  |
| Flow | K14 | $\begin{aligned} & =\text { I14 } \\ & =\text { Unit \#1 brine flow } \end{aligned}$ |
| Ca - DOC | K16-K24 | $\begin{aligned} & =\text { I16: I } 24 \\ & =\text { same contaminant []'s as unit \#1 brine flow } \end{aligned}$ |
| pH | K25 | $\begin{aligned} & =\mathrm{I} 25-\mathrm{J} 25 \\ & =\text { Unit \#1 brine } \mathrm{pH}-\text { required } \mathrm{pH} \text { adjustment } \end{aligned}$ |
| pH adjustment | J25 | $\begin{aligned} & =\mathrm{I} 25-6.0 \\ & =\text { Unit \#1 brine } \mathrm{pH}-6.0 \\ & \text { - ensures a feedwater } \mathrm{pH} \text { of } 6.0 \text { for } \mathrm{NF} \text { unit \#2 } \end{aligned}$ |
| $\mathrm{CO}_{3}$ | K27 | $\begin{aligned} & =\left(\mathrm{O}_{6} * \mathrm{~K} 29\right) /\left(10^{\wedge}-\mathrm{K} 25\right) \\ & =\mathrm{CO}_{3}-\mathrm{HCO}_{3} \text { equilibrium constant } \mathrm{x}\left[\mathrm{HCO}_{3}\right] /(10-\mathrm{pH}) \end{aligned}$ |
| $\mathrm{HCO}_{3}$ | K29 | $\begin{aligned} & =\mathrm{K} 33 * 1 /\left(\left(\left(10^{\wedge}-\mathrm{K} 25\right) / \mathrm{O} 5\right)+1+\left(\mathrm{O} 6 /\left(10^{\wedge}-\mathrm{K} 25\right)\right)\right) \\ & =\left[\mathrm{C}_{\mathrm{T}}\right] \times 1 /\left(10^{-} \mathrm{pH} / \mathrm{H}_{2} \mathrm{CO}_{3} \text { equilibrium constant }+1+\mathrm{CO}_{3}-\mathrm{HCO}_{3}\right. \text { equilibrium } \\ & \text { constant } / 10-\mathrm{pH}) \\ & \text { - from Benefield et al. } 1982 \end{aligned}$ |
| $\mathrm{H}_{2} \mathrm{CO}_{3}$ | K31 | $=\left(10^{\wedge}-\mathrm{K} 25\right) * \mathrm{~K} 29 / \mathrm{O} 5$ |

NANOFILTER SPREADSHEET (con't)

| Title | Cell(s) | Equations / values and explanation thereof |
| :---: | :---: | :---: |
|  |  | $=(10-\mathrm{pH}) \times\left[\mathrm{HCO}_{3}\right] / \mathrm{H}_{2} \mathrm{CO}_{3}-\mathrm{HCO}_{3}$ equilibrium constant |
| $\mathrm{C}_{\mathrm{T}}$ | K33 | $\begin{aligned} & =\mathrm{I} 33 \\ & =\text { Unit \#1 brine }[\mathrm{Ct}] \end{aligned}$ |
| Permeate |  |  |
| Flow - $\mathrm{C}_{\text {T }}$ | L14-L35 | - same basic calculations as cells H14 through H35 |
| Brine |  |  |
| Flow - $\mathrm{C}_{\mathrm{T}}$ | M14-M36 | - same basic calculations as cells I14 through I35 |
| - if only one NF unit is used in the treatment system: |  |  |
| Feed |  |  |
| Flow - $\mathrm{C}_{\text {T }}$ | K14-K35 | $=0$ |
| Permeate |  |  |
| Flow - $\mathrm{C}_{\text {T }}$ | L14-L35 | $=0$ |
| Brine |  |  |
| Flow - $\mathrm{C}_{\text {T }}$ | M14-M35 | $=0$ |
| Salt Precipitation Check $\mathrm{CaSO}_{4}-\mathrm{Mg}(\mathrm{OH})_{2}$ | N16-N19 | - same basic calculations as cells J16 through J19 |
| Product Water |  |  |
| Flow | O14 | $\begin{aligned} & =\text { H14 }+ \text { L14 } \\ & =\text { Unit \#1 permeate flow }+ \text { Unit \#2 permeate flow } \end{aligned}$ |
| $\mathrm{Ca}-\mathrm{pH}$ | O16-O25 | $\begin{aligned} & =(\mathrm{H} 16: \mathrm{H} 25 * \mathrm{H} 14+\mathrm{L} 16: \mathrm{L} 25 * \mathrm{~L} 14) /(\mathrm{H} 14+\mathrm{L} 14) \\ & =(\text { mass in unit \#1 permeate }+ \text { mass in unit \#2 permeate) } /(\text { unit \#1 permeate flow }+ \\ & \text { unit \#2 permeate flow) } \end{aligned}$ |
| $\mathrm{CO}_{3}$ | O27 | $\begin{aligned} & =\left(\mathrm{O}_{6} * \mathrm{O} 29\right) /\left(10^{\wedge}-\mathrm{O} 25\right) \\ & =\mathrm{CO}_{3}-\mathrm{HCO}_{3} \text { equilibrium constant } \mathrm{x}\left[\mathrm{HCO}_{3}\right] /\left(10^{-}-\mathrm{pH}\right) \end{aligned}$ |
| $\mathrm{HCO}_{3}$ | O29 | $\begin{aligned} = & \mathrm{O} 33 * 1 /\left(\left(\left(10^{\wedge}-\mathrm{O} 25\right) / \mathrm{O} 5\right)+1+\left(\mathrm{O} 6 /\left(10^{\wedge}-\mathrm{O} 25\right)\right)\right) \\ = & {\left[\mathrm{C}_{\mathrm{T}}\right] \times 1 /\left(10^{-} \mathrm{pH} / \mathrm{H}_{2} \mathrm{CO}_{3} \text { equilibrium constant }+1+\mathrm{CO}_{3}-\mathrm{HCO}_{3}\right. \text { equilibrium }} \\ & \text { constant } \left./ 10^{-} \mathrm{pH}\right) \end{aligned}$ |

NANOFILTER SPREADSHEET (con't)
Equations / values and explanation thereof
$=($ mass in unit \#1 permeate + mass in unit \#2 permeate $) /($ unit \#1 permeate flow + unit \#2 permeate flow)


Cells)
031
033


$=$ if (F7 $=$ "no", $\mathrm{Q} 14, \mathrm{Q} 14 *(\mathrm{~F} / 100))$
$=$ inlet water flow $* \%$ blending (if activated)
$=$ same contaminent [] 's as the inlet water

- same basic calculations as cells H14 through H35
- same basic calculations as cells I14 through I35
- if using only one or two NF units in the treatment system:
Unit \#3 inlet water
Flow- $\mathrm{C}_{\mathrm{T}}$
Q14-Q33
$\begin{array}{lll}\text { Unit \#3 }\end{array}$
$\begin{aligned} & \text { Feed } \\ & \text { Flow }-\mathrm{C}_{\mathrm{T}}\end{aligned}$
NANOFILTER SPREADSHEET (con't)
Equations / values and explanation thereof
Equations / values and explanation thereof
$=0$
$=0$
- same basic calculations as cells J16 through J19
$=$ values in cells $\mathrm{O} 14-\mathrm{O} 33$
 $\begin{aligned} &\text { constant } / 10-\mathrm{pH}) \\ & \text { - from Benefield et al. } 1982 \\ &=\left(10^{\wedge}-\mathrm{G} 53\right) * \mathrm{G} 55 / \mathrm{O} 5 \\ &=(10-\mathrm{pH}) \times\left[\mathrm{HCO}_{3}\right] / \mathrm{H}_{2} \mathrm{CO}_{3}-\mathrm{HCO}_{3} \text { equilibrium constant }\end{aligned}$
$=$ values in cells $\mathrm{T} 14-\mathrm{T} 33$


990

- if not using blending option:
Flow $-\mathrm{C}_{\mathrm{T}}$
NANOFILTER SPREADSHEET (con't)
Equations / values and explanation thereof
$=$ values in cells F42 - F52
$=$ value in cell F53 +pH adjustment
$=10-$ value in cell F53
- ensures that NF product water will have a pH $=10.0$
$=$ values in cells F54 - F57
$=$ values in cells G42 - G52
$=$ value in cell G53 + pH adjustment
$=10-$ value in cell G53
- ensures that NF product water will have a pH $=10.0$
$=$ values in cells G54 - G57

[^12]
RO and bCS SPREADSHEET
Equations / values and explanation thereof
$=96 \%$
$=$ average value from the reviewed literature
$=98 \%$
$=$ average value from the reviewed literature
$=90 \%$
$=$ average value from the reviewed literature
$=99.99 \%$

- assumed value derived to give very high quality product water as stipulated by
Bowlin \& Ludlum 1992 and Penkratz \& Johanson 1992
- variable open to manipulation with minimum and maximum values of 1 and 3,
respectively

evaluation
- assumed value
evaluation
- assumed value
$\mathrm{pK}=2902.39 / \mathrm{T}+0.02379(\mathrm{~T})-6.498$, where $\mathrm{T}=$ temperature in degree Kelvin
- this equation (from Benefield et. al. 1982 ) accounts for the effects of water
temperature on the $\mathrm{HCO} 3 / \mathrm{CO} 3$ equilibrium constant.
$\mathrm{pK}=17.052 / \mathrm{T}+215.2(\log \mathrm{~T})-0.12675(\mathrm{~T})-545.56$, where $\mathrm{T}=$ temperature in
degree Kelvin
- this equation (from Benefield et. al. 1982) accounts for the effects of water
$\mathrm{pK}=2902.39 / \mathrm{T}+0.02379(\mathrm{~T})-6.498$, where $\mathrm{T}=$ temperature in degree Kelvin
- this equation (from Benefield et. al. 1982 ) accounts for the effects of water
temperature on the $\mathrm{HCO} 3 / \mathrm{CO} 3$ equilibrium constant.
$\mathrm{pK}=17.052 / \mathrm{T}+215.2(\log \mathrm{~T})-0.12675(\mathrm{~T})-545.56$, where $\mathrm{T}=$ temperature in
degree Kelvin
- this equation (from Benefield et. al. 1982) accounts for the effects of water
$\mathrm{pK}=2902.39 / \mathrm{T}+0.02379(\mathrm{~T})-6.498$, where $\mathrm{T}=$ temperature in degree Kelvin
- this equation (from Benefield et. al. 1982 ) accounts for the effects of water
temperature on the $\mathrm{HCO} 3 / \mathrm{CO} 3$ equilibrium constant.
$\mathrm{pK}=17.052 / \mathrm{T}+215.2(\log \mathrm{~T})-0.12675(\mathrm{~T})-545.56$, where $\mathrm{T}=$ temperature in
degree Kelvin
- this equation (from Benefield et. al. 1982) accounts for the effects of water
$\mathrm{pK}=2902.39 / \mathrm{T}+0.02379(\mathrm{~T})-6.498$, where $\mathrm{T}=$ temperature in degree Kelvin
- this equation (from Benefield et. al. 1982 ) accounts for the effects of water
temperature on the $\mathrm{HCO} 3 / \mathrm{CO} 3$ equilibrium constant.
$\mathrm{pK}=17.052 / \mathrm{T}+215.2(\log \mathrm{~T})-0.12675(\mathrm{~T})-545.56$, where $\mathrm{T}=$ temperature in
degree Kelvin
- this equation (from Benefield et. al. 1982) accounts for the effects of water
$\mathrm{pK}=2902.39 / \mathrm{T}+0.02379(\mathrm{~T})-6.498$, where $\mathrm{T}=$ temperature in degree Kelvin
- this equation (from Benefield et. al. 1982 ) accounts for the effects of water
temperature on the $\mathrm{HCO} 3 / \mathrm{CO} 3$ equilibrium constant.
$\mathrm{pK}=17.052 / \mathrm{T}+215.2(\log \mathrm{~T})-0.12675(\mathrm{~T})-545.56$, where $\mathrm{T}=$ temperature in
degree Kelvin
- this equation (from Benefield et. al. 1982) accounts for the effects of water temperature on the $\mathrm{H}_{2} \mathrm{CO}_{3} / \mathrm{HCO}_{3}$ equilibrium constant. $\mathrm{K}=2.4 \mathrm{e}-5-$ from Kotz \& Purcell 1987
$\mathrm{pK}=0.01183 \mathrm{t}+8.03$, where $\mathrm{t}=$ temperature in degree Celsius
本
$\frac{\text { Title }}{\text { CONSTANTS }}$
Salt rejection efficiencies
RO
- monovalents
- divalents
$\quad$ - organics
BCS
Water temperature
오
$\begin{array}{ll}\text { A } & \text { 今 } \\ \text { ஆ } & \infty \\ \vdots & \vdots\end{array}$

$\stackrel{0}{4}$
Number of RO stages
90
G6
D7-F7
 RO assembly
BCS units
Carbonate constants
$\mathrm{HCO}_{3} / \mathrm{CO}_{3}$ constant
$\mathrm{H}_{2} \mathrm{CO}_{3} / \mathrm{HCO}_{3}$ constant
Solubility constants
$\mathrm{CaSO}_{4}$
$\mathrm{CaCO}_{3}$
RO AND BCS SPREADSHEET (con't)
Equations / values and explanation thereof
- this equation (from Benefield et. al. 1982) accounts for the effects of water
temperature on the CaCO3 equilibrium constant.
$\mathrm{K}=4.0 \mathrm{e}-5-\mathrm{Kotz} \&$ Purcell 1987
$\mathrm{pK}=0.0175 \mathrm{t}+9.97$, where $\mathrm{t}=$ temperature in degree Celsius
- this equation (from Benefield et. al. 1982) accounts for the effects of water
temperature on the $\mathrm{Mg}(\mathrm{OH}) 2$ equilibrium constant.
$\mathrm{pK}=4787.3 / \mathrm{T}+7.1321(\mathrm{log} \mathrm{T})+0.010365(\mathrm{~T})-22.801$, where $\mathrm{T}=$ temperature in
degree Kelvin
- this equation (from Benefield et. al. 1982) accounts for the effects of water
temperature on the water equilibrium constant.
- maximum Ca concentration that can enter the RO system without triggering salt
precipitation during filtration
$=$ IF(G13=0, 0, '[FNGP water system $]$ overview'! $\$ \mathrm{O} \$ 46-\mathrm{D} 9)$
$=$ if no RO technology is involved in the recycling system, then there is no Ca
pretreatment needed. When RO technology is used, the mass of Ca requiring
removal $=$ incoming concentration $([\mathrm{I})$ - desired Ca levels ([ ]d).
$=\mathrm{IF}(\mathrm{G} 13=0,0$, (D26*D15+E26*E15)/F15-5.8)
- feedwater pH was always adjusted to 5.8 , as recommended by 51,7 and 5

\footnotetext{


| Title | Cell(s) |
| :---: | :---: |
| $\mathrm{MgCO}_{3}$ | M6 |
| $\mathrm{Mg}(\mathrm{OH})_{2}$ | N6 |
| Kw | O6 |
| Desired Ca concentration | D9 |
| INCOMING WATER |  |
| Required Ca adjustment | C17 |
| pH adjustment | C25 |
| - if using a back-end design: |  |
| Composite discharge |  |
| Flow | D15 |
| Calcium | D17 |
| Other parameters | D18-D26 |

RO AND BCS SPREADSHEET (con't)


- if using a direct composite discharge reuse scenario


## Carbonic acid

## Total carbonates

 Lime product> Title
Carbonate
Bicarbonate
D32
D34
= flow calculated in cell J17 in the Overview spreadsheet
$=\mathrm{if}(\mathrm{C} 17>0$, '[FNGP water system] overview'!\$J\$23-C17,

- indicates that if required calcium adjustment $>0$, then calcium concentration in composite discharge stream $=$ current $[\mathrm{Ca}]$ - required adjustment. Otherwise, there is no change in [Ca] from that shown in cell J23 in the Overview spreadsheet $=$ values calculated in cells J 24 to J 32 in the Overview spreadsheet $=\left(\right.$ carbonate equilibrium constant $\left.\mathrm{x}\left[\mathrm{HCO}_{3}\right]\right) /(10-\mathrm{pH})$
$=[$ total carbonates $] \times(1 /((10-\mathrm{pH} /$ bicarbonate equilibrium constant $)+1+($ carbonate
- formula from Benefield et. al. 1982
$=\left(10-\mathrm{pH}_{x}\left[\mathrm{HCO}_{3}\right]\right) /$ bicarbonate equilibrium constant
$=$ value calculated in cell J37 in Overview spreadsheet
= if(and('[FNGP water system]overview'!\$D\$19="Composite", '[FNGP water system]overview'!\$O\$15="RO"), '[FNGP water
system]overview'!\$N $21: \$ \mathrm{~N} \$ 37,0$ )
D18-D26

దం
D32
D34
E15-E34
Softener Blowdown
All parameters Mg through pH
Calcium
Bicarbonate
Carbonic acid
Total carbonates

| Title | Cell(s) |
| :---: | :---: |
| Carbonate | D28 |
| Bicarbonate | D30 |
| Carbonic acid | D32 |
| Total carbonates | D34 |
| - if using a direct composite discharge reuse s |  |
| Flow | D15 |
| Calcium | D17 |
| Mg through pH | D18-D26 |
| Carbonate | D28 |
| Bicarbonate | D30 |
| Carbonic acid | D32 |
| Total carbonates | D34 |
| Softener Blowdown |  |
| All parameters | E15-E34 |

RO AND BCS SPREADSHEET (con't)
$\left.\begin{array}{l}\text { Title } \\ \hline \text { - indicates that if cells D19 and O15 in the Overview worksheet }=\text { "composite' and "RO", respectively, then cells E15 to E34 will have } \\ \text { the same value as cells N21 through N37 in the Overview worksheet. Otherwise, } \\ \text { these cells will have values equal to zero. } \\ \text { - this equation allows one to discharge softener wastes directly to the RO system, } \\ \text { instead of to the lime pond. This option was never used during this study. }\end{array}\right]$
RO AND BCS SPREADSHEET (con't)



$$
\begin{aligned}
& =\text { if(G13 }(\mathrm{G}=0,0, \mathrm{G} 33-\mathrm{G} 35) \\
& \text { F } \mathrm{F} 22 \\
& - \text { see explanation below } \\
& =\text { if(G13 }=0,0, \mathrm{G} 33+\mathrm{G} 31+\mathrm{G} 29) \\
& =0 \text { if no RO filtration is used, or initial }\left[\mathrm{CO}_{3}\right]+\text { initial }\left[\mathrm{HCO}_{3}\right]+\text { initial }\left[\mathrm{H}_{2} \mathrm{CO} 3\right] \text { if } \\
& \text { RO filters are included in the ZED treatment train } \\
& =\text { if(G13 }=0,0,\left(-\left(\mathrm{G} 31+10^{-\mathrm{G} 27}+\mathrm{J} 6\right)+\left(\left(\mathrm{G} 31+10^{-\mathrm{G} 27}+\mathrm{J} 6\right)^{2}-4(-\mathrm{J} 6 \times \mathrm{G} 33+\right.\right. \\
& \left.\left.\mathrm{G} 31 \times 10^{-\mathrm{G} 27} 0.5\right) / 2\right) \\
& - \text { see explanation below }
\end{aligned}
$$


RO AND BCS SPREADSHEET (con't)
Equations / values and explanation thereof
$=\mathrm{if}\left(\mathrm{G} 13=0,0,\left(-\left(\mathrm{G} 29+\mathrm{G} 35+10^{-\mathrm{G} 27}+\mathrm{I} 6\right)+\left(\left(\mathrm{G} 29+\mathrm{G} 35+10^{-\mathrm{G} 27}+\mathrm{I} 6\right)^{2}-4(-\right.\right.\right.$
$\left.\left.\left.\left.\mathrm{I} 6 \times(\mathrm{G} 31+\mathrm{G} 35)+\mathrm{G} 29\left(10^{-\mathrm{G} 27}+\mathrm{G} 35\right)\right)\right)^{0.5}\right) / 2\right)$
-
.

## EXPLANATION:

Benefield et. al. (1982) indicates that most carbon dioxide dissolved in water does not hydrolyse to carbonic acid. The equilibrium equations used in this model to determine carbonic acid concentrations (e.g. cell D32) assume, however, that the concentration of carbonic acid is = to the sum of dissolved carbon dioxide and actual carbonic acid molecules. RO membranes cannot eliminate dissolved gases; they will pass unhindered through the filters. Rather than trying to calculate the proportion of carbon dioxide existing on its own and as carbonic acid, all carbonic acid within the RO feedwaters was assumed to be dissolved carbon dioxide. It therefore freely passes' though the membranes. RO concentrate and permeate waters will, as a result, have identical $\mathrm{H}_{2} \mathrm{CO}_{3}$ concentrations.
Carbonate and bicarbonate ions will not pass unhindered through the RO filters, due to the limited permeability of RO membranes to monovalent and divalent ions. While permeate and concentrate streams will have identical concentrations of $\mathrm{H}_{2} \mathrm{CO}_{3}$, they will contain different amounts of carbonate and bicarbonate. Neither system will be in equilibrium. There will either be "too much" or "too little". carbonate and bicarbonate. Concentrations within each stream will shift until carbonate and bicarbonate equilibriums are re-established. $\mathrm{H} 2 \mathrm{CO} 3-\mathrm{HCO} 3$ Equilibrium

RO AND BCS SPREADSHEET (con't)

RO AND BCS SPREADSHEET (con't)
Equations / values and explanation thereof



## Total carbonate



EXPLANATION:
-as previously explained with the product stream, the concentration of each carbonate species will shift from its initial level in order to
re-establish carbonate equilibriums. Although the equilibrium calculations used for the permeate still hold true for the brine flow, the
values of certain variables had changed:

$$
\begin{aligned}y \text { is now }=\% \text { monovalent ion rejection } * \text { feedwater [ } \mathrm{HCO} 3] \\ \text { w is now }=\% \text { divalent ion rejection * feedwater [CO3] }\end{aligned}
$$

EXPLANATION:
-as previously explained with the product stream, the concentration of each carbonate species will shift from its initial level in order to
re-establish carbonate equilibriums. Although the equilibrium calculations used for the permeate still hold true for the brine flow, the
values of certain variables had changed:

$$
\begin{aligned}y \text { is now }=\% \text { monovalent ion rejection * feedwater [ } \mathrm{HCO} 3] \\ \text { w is now }=\% \text { divalent ion rejection } * \text { feedwater [CO3] }\end{aligned}
$$

EXPLANATION:
-as previously explained with the product stream, the concentration of each carbonate species will shift from its initial level in order to
re-establish carbonate equilibriums. Although the equilibrium calculations used for the permeate still hold true for the brine flow, the
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$$
\begin{aligned}y \text { is now }=\% \text { monovalent ion rejection * feedwater [ } \mathrm{HCO} 3] \\ \text { w is now }=\% \text { divalent ion rejection } * \text { feedwater [CO3] }\end{aligned}
$$

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-as previously explained with the product stream, the concentration of each carbonate species will shift from its initial level in order to
re-establish carbonate equilibriums. Although the equilibrium calculations used for the permeate still hold true for the brine flow, the
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\begin{aligned}y \text { is now }=\% \text { monovalent ion rejection * feedwater [ } \mathrm{HCO} 3] \\ \text { w is now }=\% \text { divalent ion rejection } * \text { feedwater [CO3] }\end{aligned}
$$

EXPLANATION:
-as previously explained with the product stream, the concentration of each carbonate species will shift from its initial level in order to
re-establish carbonate equilibriums. Although the equilibrium calculations used for the permeate still hold true for the brine flow, the
values of certain variables had changed:

$$
\begin{aligned}y \text { is now }=\% \text { monovalent ion rejection * feedwater [ } \mathrm{HCO} 3] \\ \text { w is now }=\% \text { divalent ion rejection } * \text { feedwater [CO3] }\end{aligned}
$$

Conc.shift ( $\mathrm{HCO}_{3}-\mathrm{CO}_{3}$ )

Bicarbonate
[ש!!!u! -
um!!q!!!nbe -
Carbonic acid
Lentuq!!!

- initial (
RO AND BCS SPREADSHEET (con't)

$\frac{\text { Title }}{\text { Salt precipitation check }}$
$\mathrm{CaSO}_{4}$
$\mathrm{CaCO}_{3}$
$\mathrm{MgCO}_{3}$
$\mathrm{Mg}(\mathrm{OH})_{2}$


[^13]
-Same basic calculations as cells J17 through J20 - variable open to manipulation
= L15
$=$ Unit \#2 brine flow
= L17 : L25
= Contaminant []'s as Unit \#3 brine flow
$=$ L26-M26
$=$ Unit \#3 brine $\mathrm{pH}-\mathrm{pH}$ adjustment

$=\mathrm{L} 26-5.8$ | Title | Cell(s) |
| :--- | :---: |
| Permeate | K15-K36 |
| Concentrate | L15-L36 |
| -if only using a 1-stage RO system: |  |
| Feed | J15-J36 |
| Product water | K15-K36 |
| Concentrate | L15-L36 |
|  |  |
| $\begin{array}{l}\text { Salt Precipitation Check } \\ \text { CaSO }\end{array}$ |  | RO UNIT \#3

Water recovery ratio $\quad \mathrm{O} 13$

RO AND BCS SPREADSHEET (con't)
Equations / values and explanation thereof
$=$ Unit \#3 brine $\mathrm{pH}-5.8$
-Ensures feedwater $\mathrm{pH}=5.8$
$=\left(\mathrm{I}^{*} \mathrm{~N} 30\right) /\left(10^{\wedge}-\mathrm{N} 26\right)$
$=\mathrm{CO}_{3}-\mathrm{HCO}_{3}$ equilibrium constant $\times\left[\mathrm{HCO}_{3}\right] /(10-\mathrm{pH})$
$=\mathrm{N} 34 *\left(1 /\left(\left(\left(10^{\wedge}-\mathrm{N} 26\right) / \mathrm{J} 6\right)+1+\left(\mathrm{I} 6 / 10^{\wedge}-\mathrm{N} 26\right)\right)\right)$
$=\left[\mathrm{C}_{\mathrm{T}}\right] \times 1 /\left(10^{-\mathrm{pH}} / \mathrm{H}_{2} \mathrm{CO}_{3}\right.$ equilibrium constant $+1+\mathrm{CO}_{3}-\mathrm{HCO}_{3}$ equilibrium
constant $\left./ 10^{-}-\mathrm{pH}\right)$
-From Benefield et al. 1982
$=\left(\left(10^{\wedge}-\mathrm{N} 26\right) * \mathrm{~N}_{3}\right) / \mathrm{J}^{2}$
$=\left(10^{-}-\mathrm{pH}\right) \times\left[\mathrm{HCO}_{3}\right] / \mathrm{H}_{2} \mathrm{CO}_{3}-\mathrm{HCO}_{3}$ equilibrium constant
$=\mathrm{L} 34$
$=$

## $=$ Same basic equations as cells G15 through G36 <br> 

$=0$
$=0$
$=0$
-Same basic calculations as cells J17 through J20

- as stated earlier, feedwater pH always lowered to 5.8

| Title | Cell(s) |
| :--- | :---: |
| $\mathrm{CO}_{3}$ | N 28 |
| $\mathrm{HCO}_{3}$ | N 30 |
|  |  |
| $\mathrm{H}_{2} \mathrm{CO}_{3}$ | N 32 |
| $\mathrm{C}_{\mathrm{T}}$ | N 34 |
| Permeate | $\mathrm{O} 15-\mathrm{O} 36$ |
| Concentrate | $\mathrm{P} 15-\mathrm{P} 36$ |
| -if only using 1 or 2 RO units: |  |
| Feed |  |
| Product water - N36 |  |
| Concentrate | $\mathrm{O} 15-\mathrm{O} 36$ |
| Salt Precipitation Check |  |
| CaSO |  |
| - Mg (OH) |  |
| pH adjustment | $\mathrm{Q} 17-\mathrm{Q} 20$ |

RO AND BCS SPREADSHEET (con't)


R28-R32
T15-T34
U15-U34
Overall Brine
BRINE CONCENTRATOR
Feedwater D43-D62

F43-F62
H43-H62

$\frac{\text { Cell(s) }}{}$
R15
R17-R26 + R34
U15-U34
Water recovery ratio
Product water
$\frac{\text { SPRAY DRYER }}{\text { Feedwater }}$
Concentrate
Feedwater
RO AND BCS SPREADSHEET (con't)

| Title | Cell(s) |
| :--- | :---: |
| Waste | J43 |
| Flow | Chemical parameters |
| Total waste mass | J62 |

## $\frac{\text { WATER RECLAMATION SYSTEM }}{\text { Product water } \mathrm{pH}} \frac{\text { L54 }}{}$ <br> Product water pH L54

tSW
~
$\mathrm{N} 45-\mathrm{N} 53+\mathrm{N} 62$
N 54
N56-N60
 BCS assembly was not part of the trea
flow = RO permeate flows, cell R15

- similar equations as for product water
= product water $\mathrm{pH}+\mathrm{pH}$ adjustment
- calculated in similar format as feedwat recycling system, then $=$ cell U15 is added to the solid wast to is added to the solid waste output
- identical format to cell P43

| Product water <br> Flow | N 43 |
| :--- | :---: |
|  |  |
| $\mathrm{Ca}-\mathrm{DOC}+\mathrm{CT}$ | $\mathrm{N} 45-\mathrm{N} 53+\mathrm{N} 62$ |
| pH | N 54 |
| $\mathrm{CO} 3-\mathrm{H}_{2} \mathrm{CO} 3$ | $\mathrm{~N} 56-\mathrm{N} 60$ |
| Waste | P 43 |
| Flow | P 45 |
| Calcium | $\mathrm{P} 46-\mathrm{P} 60$ |

$\mathrm{Ca}-\mathrm{DOC}+\mathrm{C}_{\mathrm{T}}$
$\stackrel{\mathrm{pH}}{\mathrm{CH}_{3}}$
Waste
RO AND BCS SPREADSHEET (con't)
Equations / values and explanation thereof
$=$ total carbonate in RO permeate (cell U34) if there is no BCS assembly used in the
recycling system. If a BCS assembly is used, then this cell $=$ total mass released
from the spray dryer (sum of cells P45-P60)
事


## APPENDIX F

## SIMULATOR SUMMARY SHEETS

A summary sheet was made every time a different scenario was tested with the computer simulator. As a result, more than 100 summary sheets were generated over the course of this study. While all of them are included on the computer disk accompanying this document, only a few were converted into hard copies and included in this appendix to illustrate the type of information that was inputted into the simulator.

固தckrend model - fosiege RO, RCS, IX, no recyele
Test Conditions


| Water flows | F.E.T. inlet | Lime blow. Soften blow. | T.water | R.O. inlet | R.O. waste | B/C feed |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  | 578.3 | 13.9 | 19.0 | 545.3 | 281.6 | 140.8 | 140.8 |


| Water quality |  | Raw water | Water from RO \& B/C | Treated water | Current <br> T. water | Final wast Solid |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| Flow | m3/day | 407.4 | 267.5 | 545.3 | - | - |  |
| Chem. |  |  |  |  |  |  |  |
| Ca | $\mathrm{mg} / \mathrm{L}$ | 81.3 | 1.3 | 1.4 | 1.4 | $3.2 \mathrm{E}+01$ | kg/day |
| Mg | $\mathrm{mg} / \mathrm{L}$ | 18.3 | 0.3 | 0.5 | 0.5 | $4.2 \mathrm{E}+00$ | kg/day |
| Fe | $\mathrm{mg} / \mathrm{L}$ | 0.39 | 0.0 | 0.6 | 0.6 | 4.2E-01 | $\mathrm{kg} /$ day |
| SO4 | $\mathrm{mg} / \mathrm{L}$ | 86.8 | 7.4 | 61.6 | 81.9 | $9.6 \mathrm{E}+01$ | kg/day |
| SiO 2 | $\mathrm{mg} / \mathrm{L}$ | 5.2 | 0.1 | 0.9 | 0.9 | $1.2 \mathrm{E}+00$ | kg/day |
| PO4-P | $\mathrm{mg} / \mathrm{L}$ | 0.06 | 0.01 | 0.07 | 0.07 | 9.1E-02 | kg/day |
| Cl | $\mathrm{mg} / \mathrm{L}$ | 2.9 | 19.2 | 2.1 | 2.4 | $1.2 \mathrm{E}+02$ | kg/day |
| Na | $\mathrm{mg} / \mathrm{L}$ | 12.5 | 20.3 | 49.5 | 50.5 | $1.3 \mathrm{E}+02$ | kg/day |
| DOC | $\mathrm{mg} / \mathrm{L}$ | 9.3 | 8.4 | 6.0 | 6.1 | $2.0 \mathrm{E}+01$ | kg/day |
| pH |  | 7.9 | 4.7 | 10.0 | 10.1 |  |  |
| CO3 | $\mathrm{mol} / \mathrm{L}$ | 1.4E-05 | 2.1E-06 | 1.4E-04 | 1.2E-04 | 7.6E-04 | $\mathrm{kg} / \mathrm{day}$ |
| HCO3 | $\mathrm{mol} / \mathrm{L}$ | 3.6E-03 | 1.4E-03 | 3.0E-04 | 2.0E-04 | $7.2 \mathrm{E}+00$ | $\mathrm{kg} / \mathrm{day}$ |
| H2CO3 | $\mathrm{mol} / \mathrm{L}$ | 1.0E-04 | 1.0E-04 | 7.1E-08 | 3.7E-08 | $1.3 \mathrm{E}+00$ | kg/day |
| Ctot | $\mathrm{mol} / \mathrm{L}$ | 3.7E-03 | 1.5E-03 | 4.4E-04 | 3.1E-04 |  |  |
| Required pH adjustments: |  |  | 2.8 | for recyc. $\mathrm{H} 20 \mathrm{pH}=7.5$ |  | 414.8 | kg/day |

回®ckrend model = justage RO, BCS, IK, 『ecycle
Test Conditions


## Results

| Solubility Check | Nanofiltration |  |  | Reverse Osmosis |  |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  | unit 1 | unit 2 | unit 3 | unit 1 | unit 2 | unit 3 |
| CaSO4 | N/A | N/A | N/A | ok | N/A | N/A |
| CaCO3 | N/A | N/A | N/A | ok | N/A | N/A |
| MgCO 3 | N/A | N/A | N/A | ok | N/A | N/A |
| $\mathrm{Mg}(\mathrm{OH}) 2$ | N/A | N/A | N/A | ok | N/A | N/A |


| Water flows | F.E.T. inlet | Lime blow. Soften blow. | T.water | R.O. inlet | R.O. waste | B/C feed |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  | 536.6 | 7.2 | 3.8 | 525.5 | 239.9 | 119.9 | 119.9 |


| Water quality |  | Raw water | Recycled Water | Treated water | Current <br> T. water |  | Final wast Solid |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| Flow | m3/day | 137.8 | 227.9 | 525.5 | - |  | - |  |
| Chem. |  |  |  |  |  |  |  |  |
| Ca | $\mathrm{mg} / \mathrm{L}$ | 81.3 | 1.3 | 1.4 | 1.4 |  | $1.4 \mathrm{E}+01$ | kg/day |
| Mg | $\mathrm{mg} / \mathrm{L}$ | 18.3 | 0.2 | 0.5 | 0.5 |  | $2.6 \mathrm{E}+00$ | kg/day |
| Fe | $\mathrm{mg} / \mathrm{L}$ | 0.39 | 0.0 | 0.3 | 0.6 |  | 3.9E-01 | kg/day |
| SO4 | $\mathrm{mg} / \mathrm{L}$ | 86.8 | 5.8 | 25.3 | 81.9 |  | $6.4 \mathrm{E}+01$ | kg/day |
| SiO 2 | mg/L | 5.2 | 0.1 | 0.9 | 0.9 |  | 8.8E-01 | kg/day |
| PO4-P. | $\mathrm{mg} / \mathrm{L}$ | 0.06 | 0.01 | 0.04 | 0.07 |  | 9.0E-02 | kg/day |
| Cl | $\mathrm{mg} / \mathrm{L}$ | 2.9 | 7.9 | 4.2 | 2.4 |  | $4.3 \mathrm{E}+01$ | kg/day |
| Na | $\mathrm{mg} / \mathrm{L}$ | 12.5 | 14.7 | 16.6 | 50.5 |  | $8.0 \mathrm{E}+01$ | kg/day |
| DOC | $\mathrm{mg} / \mathrm{L}$ | 9.3 | 9.7 | 6.0 | 6.1 |  | $2.0 \mathrm{E}+01$ | kg/day |
| pH |  | 7.9 | 7.5 | 10.0 | 10.1 |  |  |  |
| CO 3 | $\mathrm{mol} / \mathrm{L}$ | 1.4E-05 | 2.2E-06 | $7.0 \mathrm{E}-05$ | 1.2E-04 |  | 7.8E-04 | kg/day |
| HCO | $\mathrm{mol} / \mathrm{L}$ | 3.6E-03 | $1.5 \mathrm{E}-03$ | $1.5 \mathrm{E}-04$ | 2.0E-04 |  | $6.5 \mathrm{E}+00$ | kg/day |
| H 2 CO | $\mathrm{mol} / \mathrm{L}$ | 1.0E-04 | 1.1E-04 | $3.5 \mathrm{E}-08$ | 3.7E-08 |  | $1.2 \mathrm{E}+00$ | kg/day |
| Ctot | $\mathrm{mol} / \mathrm{L}$ | 3.7E-03 | 1.6E-03 | 2.2E-04 | 3.1E-04 |  |  |  |
| Required pH adjustments: |  |  | 2.8 | for recyc. $\mathrm{H} 20 \mathrm{pH}=7.5$ Total |  |  | 233.7 | kg/day |

Recksend model - !astege RO, BCS, NF, n® 『ecycle
Test Conditions

| Raw water origin: | No name creek |  | NF | R.O. | $B / C$ |
| :---: | :---: | :---: | :---: | :---: | :---: |
| Volume 8\# vented: | $0.0 \mathrm{~m} / \mathrm{day}$ | \# stages | 3 | 1 | 1 |
| Softener type: | NF | Water temp (C) | 40 | 35 | 25 |
| Water recycling: |  | Removal |  |  |  |
| -Brine disposal | Evaporated | monovalent | 67\% | 96\% | 99.99\% |
| -Water recycled? | no | divalent | 94\% | 98\% | 99.99\% |
|  |  | TOC | 98\% | 90\% | 99.99\% |
|  |  |  | Water recovery: |  | 90\% |


|  | Nanofiltration |  |  | Reverse Osmosis |  |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  | unit 1 | unit 2 | unit 3 | unit 1 | unit 2 | unit 3 |
| Water recovery | 75\% | 50\% | 75\% | 50\% | N/A | N/A |
| Brine recycle | - | no | yes |  |  |  |
| \% recycle |  | 20\% | 100\% |  |  |  |
| Blending | Vekm |  | no |  |  |  |
| \% of product from \#1+2 into \#3 50\% |  |  |  |  |  |  |
| Feedwater pH | 6.5 | 6.0 | 6.2 | 5.8 | N/A | N/A |
| adjustment | 2.7 | 0.8 |  | 1.8 | N/A | N/A |
| Feedwater Ca | 35.0 |  |  | 65 | Yevemue | *** |
| adjustment | 0.0 |  |  | 50 |  |  |

## Results

| Solubility Check | Nanofiltration |  |  |  | Reverse Osmosis |  |  |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| CaSO4 | unit 1 | unit 2 | unit 3 | unit 1 | unit 2 | unit 3 |  |  |
|  | ok | ok | ok | ok | $\mathrm{N} / \mathrm{A}$ | $\mathrm{N} / \mathrm{A}$ |  |  |
| CaCO 3 | ok | ok | ok | ok | $\mathrm{N} / \mathrm{A}$ | $\mathrm{N} / \mathrm{A}$ |  |  |
| MgCO 3 | ok | ok | ok | ok | $\mathrm{N} / \mathrm{A}$ | $\mathrm{N} / \mathrm{A}$ |  |  |
| $\mathrm{Mg}(\mathrm{OH}) 2$ | ok | ok | ok | ok | $\mathrm{N} / \mathrm{A}$ | $\mathrm{N} / \mathrm{A}$ |  |  |


| Water flows | F.E.T. inlet | Lime blow. Soften blow. | T.water | R.O. inlet | R.O. waste | B/C feed |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  | 576.1 | 13.9 | 89.9 | 472.2 | 279.3 | 139.7 | 139.7 |


| Water qual | lity | Raw <br> water | $\mathrm{RO}+\mathrm{B} / \mathrm{C}$ <br> Water | Treated water | Current <br> T. water |  | Final waste Solid |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| Flow | m3/day | 405.1 | 265.4 | 472.2 | - |  | . - |  |
| Chem. |  |  |  |  |  |  |  |  |
| Ca | $\mathrm{mg} / \mathrm{L}$ | 81.3 | 1.38 | 0.28 | 1.4 |  | $3.2 \mathrm{E}+01$ | kg/day |
| Mg | $\mathrm{mg} / \mathrm{L}$ | 18.3 | 0.33 | 0.01 | 0.5 |  | $4.2 \mathrm{E}+00$ | kg/day |
| Fe | $\mathrm{mg} / \mathrm{L}$ | 0.39 | 0.03 | 0.00 | 0.6 |  | 4.3E-01 | kg/day |
| SO4 | $\mathrm{mg} / \mathrm{L}$ | 86.8 | 7.67 | 0.50 | 81.9 |  | $9.9 \mathrm{E}+01$ | kg/day |
| SiO 2 | mg/L | 5.2 | 0.09 | 0.01 | 0.9 |  | $1.2 \mathrm{E}+00$ | kg/day |
| PO4-P | $\mathrm{mg} / \mathrm{L}$ | 0.06 | 0.01 | 0.00 | 0.07 |  | 9.0E-02 | kg/day |
| Cl | $\mathrm{mg} / \mathrm{L}$ | 2.9 | 3.54 | 0.71 | 2.4 |  | $2.2 \mathrm{E}+01$ | kg/day |
| Na | $\mathrm{mg} / \mathrm{L}$ | 12.5 | 10.81 | 3.15 | 50.5 |  | $6.9 \mathrm{E}+01$ | kg/day |
| DOC | $\mathrm{mg} / \mathrm{L}$ | 9.3 | 8.30 | 0.01 | 6.1 |  | $2.0 \mathrm{E}+01$ | kg/day |
| pH |  | 7.9 | 4.7 | 10.0 | 10.1 |  |  |  |
| CO 3 | $\mathrm{mol} / \mathrm{L}$ | 1.4E-05 | 2.1E-06 | 8.1E-05 | 1.2E-04 |  | 7.6E-04 | kg/day |
| HCO3 | $\mathrm{mol} / \mathrm{L}$ | 3.6E-03 | $1.4 \mathrm{E}-03$ | 1.7E-04 | 2.0E-04 |  | $7.2 \mathrm{E}+00$ | kg/day |
| H2CO3 | $\mathrm{mol} / \mathrm{L}$ | 1.0E-04 | $1.0 \mathrm{E}-04$ | 4.0E-08 | 3.7E-08 |  | $1.3 \mathrm{E}+00$ | kg/day |
| Ctot | $\mathrm{mol} / \mathrm{L}$ | 3.7E-03 | $1.5 \mathrm{E}-03$ | 2.5E-04 | 3.1E-04 |  |  |  |
| Required pH adjustments: |  |  | $2.8$ | for recyc. $\mathrm{H} 20 \mathrm{pH}=7.5$ <br> for T.W. from $N F=10.0$ |  | Total | 256.3 | kg/day |

R®ckreగd model - jostege RO, RCS, NF; recycle
Test Conditions


Results

| Solubility Check | Nanofiltration |  |  |  | Reverse Osmosis |  |  |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  | unit 1 | unit 2 | unit 3 | unit 1 | unit 2 | unit 3 |  |  |
| CaSO4 | ok | ok | ok | ok | $\mathrm{N} / \mathrm{A}$ | $\mathrm{N} / \mathrm{A}$ |  |  |
| CaCO 3 | ok | ok | ok | ok | $\mathrm{N} / \mathrm{A}$ | $\mathrm{N} / \mathrm{A}$ |  |  |
| MgCO 3 | ok | ok | ok | ok | $\mathrm{N} / \mathrm{A}$ | $\mathrm{N} / \mathrm{A}$ |  |  |
| $\mathrm{Mg}(\mathrm{OH}) 2$ | ok | ok | ok | ok | $\mathrm{N} / \mathrm{A}$ | $\mathrm{N} / \mathrm{A}$ |  |  |


| Water flows | F.E.T. inlet | Lime blow. Soften blow. | T.water | R.O. inlet | R.O. waste B/C feed |  |  |
| ---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  | 521.2 | 7.4 | 41.9 | 471.8 | 224.5 | 112.2 | 112.2 |


| Water q | ality | Raw water | Recycled Water | Treated water | Current <br> T. water |  | Final waste Solid |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| Flow | m3/day | 137.0 | 213.3 | 471.8 | - |  | - |  |
| Chem. |  |  |  |  |  |  |  |  |
| Ca | $\mathrm{mg} / \mathrm{L}$ | 81.3 | 1.43 | 0.04 | 1.4 |  | $1.4 \mathrm{E}+01$ | kg/day |
| Mg | $\mathrm{mg} / \mathrm{L}$ | 18.3 | 0.26 | 0.01 | 0.5 |  | $2.7 \mathrm{E}+00$ | kg/day |
| Fe | $\mathrm{mg} / \mathrm{L}$ | 0.39 | 0.04 | 0.00 | 0.6 |  | $4.0 \mathrm{E}-01$ | $\mathrm{kg} /$ day |
| SO4 | $\mathrm{mg} / \mathrm{L}$ | 86.8 | 6.34 | 0.19 | 81.9 |  | $6.6 \mathrm{E}+01$ | kg/day |
| SiO 2 | $\mathrm{mg} / \mathrm{L}$ | 5.2 | 0.09 | 0.01 | 0.9 |  | 8.9E-01 | kg/day |
| PO4-P | $\mathrm{mg} / \mathrm{L}$ | 0.06 | 0.01 | 0.00 | 0.07 |  | 8.9E-02 | kg/day |
| Cl | $\mathrm{mg} / \mathrm{L}$ | 2.9 | 4.43 | 0.82 | 2.4 |  | $2.3 \mathrm{E}+01$ | kg/day |
| Na | $\mathrm{mg} / \mathrm{L}$ | 12.5 | 13.36 | 2.88 | 50.5 |  | $6.8 \mathrm{E}+01$ | kg/day |
| DOC | $\mathrm{mg} / \mathrm{L}$ | 9.3 | 10.23 | 0.01 | 6.1 |  | $2.0 \mathrm{E}+01$ | kg/day |
| pH |  | 7.9 | 7.5 | 10.0 | 10.1 |  |  |  |
| CO 3 | $\mathrm{mol} / \mathrm{L}$ | 1.4E-05 | 2.4E-06 | 6.2E-05 | 1.2E-04 |  | 6.7E-04 | kg/day |
| HCO3 | $\mathrm{mol} / \mathrm{L}$ | 3.6E-03 | 1.6E-03 | 1.3E-04 | 2.0E-04 |  | $6.6 \mathrm{E}+00$ | kg/day |
| H2CO3 | $\mathrm{mol} / \mathrm{L}$ | 1.0E-04 | 1.2E-04 | 3.1E-08 | 3.7E-08 |  | $1.2 \mathrm{E}+00$ | kg/day |
| Ctot | $\mathrm{mol} / \mathrm{L}$ | 3.7E-03 | 1.7E-03 | 1.9E-04 | 3.1E-04 |  |  |  |
| Required pH adjustments: |  |  | $2.8$ | for recyc. $\mathrm{H} 20 \mathrm{pH}=7.5$ <br> for $\mathrm{T} . \mathrm{W}$. from $\mathrm{NF}=10.0$ |  | Total | 202.6 kg/day |  |
|  |  |  | 4.2 |  |  |  |  |  |

Beckoend nodel - \{ostbo( RO, DW, OK, no recycle

## Test Conditions

| Raw water origin: Volume 8\# vented: | No name creek |  |  | NF | R.O. | $B / C$ |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  | 0.0 | m3/day | \# stages | N/A | 1 | N/A |
| Softener type: <br> Water recycling: | Ion-X | Water temp (C) |  | N/A | 35 | N/A |
|  |  | Removal |  |  |  |  |
| -Brine disposal <br> -Water recycled? | Deep well | monovalent |  | N/A | 96\% | N/A |
|  | no |  | divalent | N/A | 98\% | N/A |
|  |  |  | TOC | N/A | 90\% | N/A |
|  |  |  |  | Water recovery: |  | N/A |
|  | Nanofiltration |  |  | Reverse Osmosis |  |  |
|  | unit 1 | unit 2 | Unit 3 | unit 1 | unit 2 | unit 3 |
| Water recovery Brine recycle | N/A | N/A | N/A | 50\% | N/A | N/A |
|  | N/A | N/A | N/A |  |  |  |
| \% recycle | - - | $N / A$ | N/A |  | K |  |
| Blending | \%\%yyyyyy |  | N/A |  |  |  |
| \% of prod | duct from \#1+ | 2 into \#3 | N/A |  |  |  |
| Feedwater pH adjustment | N/A | N/A | N/A | 5.8 | N/A | N/A |
|  | N/A | N/A | - | 1.8 | N/A | N/A |
| Feedwater Ca | N/A |  |  | 65 | -10 Kumume |  |
| adjustment | N/A | \%osikimit |  | 48 |  |  |

## Results

| Solubility Check | Nanofiltration |  |  |  | Reverse Osmosis |  |  |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| CaSO4 | unit 1 | unit 2 | unit 3 | unit 1 | unit 2 | unit 3 |  |  |
| CaCO3 | N/A | N/A | N/A | ok | N/A | N/A |  |  |
| MgCO3 | N/A | N/A | N/A | ok | N/A | N/A |  |  |
| $M g(O H) 2$ | N/A | N/A | N/A | ok | N/A | N/A |  |  |
| M | ok | N/A | N/A |  |  |  |  |  |


| Water flows | F.E.T. inlet | Lime blow. Soften blow. | T.water | R.O. inlet | R.O. waste $\mathrm{B} / \mathrm{C}$ feed |  |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| $\mathrm{m} / \mathrm{day}$ | 578.3 | 13.9 | 19.0 | 545.3 | 281.6 | 140.8 | $\mathrm{~N} / \mathrm{A}$ |


| Water qua | lity | Raw water | Recycled Water | Treated water | Current <br> T. water | Liquid | Final waste Solid |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| Flow | m3/day | 407.4 | 0.0 | 545.3 | - | 140.8 |  |  |
| Chem. |  |  | \% of wastewater not recovered |  |  | 50.0\% |  |  |
| Ca | $\mathrm{mg} / \mathrm{L}$ | 81.3 | 0 | 1.4 | 1.4 | $2.2 \mathrm{E}+02$ | $3.2 E+01$ | kg/day |
| Mg | $\mathrm{mg} / \mathrm{L}$ | 18.3 | 0.0 | 0.5 | 0.5 | $3.0 \mathrm{E}+01$ | $4.2 E+00$ | $\mathrm{kg} / \mathrm{day}$ |
| Fe | $\mathrm{mg} / \mathrm{L}$ | 0.39 | 0.0 | 0.6 | 0.6 | $3.0 E+00$ | 4.2E-01 | kg/day |
| SO4 | $\mathrm{mg} / \mathrm{L}$ | 86.8 | 0 | 61.6 | 81.9 | $6.8 \mathrm{E}+02$ | $9.6 \mathrm{E}+01$ | kg/day |
| SiO 2 | $\mathrm{mg} / \mathrm{L}$ | 5.2 | 0.0 | 0.9 | 0.9 | $8.3 \mathrm{E}+00$ | $1.2 \mathrm{E}+00$ | $\mathrm{kg} /$ day |
| PO4-P | $\mathrm{mg} / \mathrm{L}$ | 0.06 | 0.00 | 0.07 | 0.07 | 6.5E-01 | 9.1E-02 | kg/day |
| Cl | $\mathrm{mg} / \mathrm{L}$ | 2.9 | 0 | 2.1 | 2.4 | $8.7 \mathrm{E}+02$ | $1.2 \mathrm{E}+02$ | kg/day |
| Na | $\mathrm{mg} / \mathrm{L}$ | 12.5 | 0 | 49.5 | 50.5 | $9.2 \mathrm{E}+02$ | $1.3 \mathrm{E}+02$ | kg/day |
| DOC | $\mathrm{mg} / \mathrm{L}$ | 9.3 | 0.0 | 6.0 | 6.1 | $1.4 \mathrm{E}+02$ | $2.0 \mathrm{E}+01$ | kg/day |
| pH |  | 7.9 | 0.0 | 10.0 | 10.1 | 6.1 |  |  |
| CO3 | $\mathrm{mol} / \mathrm{L}$ | 1.4E-05 | 0.0E+00 | 1.4E-04 | 1.2E-04 | 5.5E-08 | 4.6E-04 | kg/day |
| HCO3 | $\mathrm{mol} / \mathrm{L}$ | 3.6E-03 | $0.0 \mathrm{E}+00$ | 3.0E-04 | 2.0E-04 | 8.3E-04 | $7.2 \mathrm{E}+00$ | kg/day |
| H2CO3 | $\mathrm{mol} / \mathrm{L}$ | 1.0E-04 | $0.0 \mathrm{E}+00$ | 7.1E-08 | 3.7E-08 | 1.5E-03 | $1.3 \mathrm{E}+01$ | kg/day |
| Ctot | $\mathrm{mol} / \mathrm{L}$ | 3.7E-03 | $0.0 \mathrm{E}+00$ | 4.4E-04 | 3.1E-04 | 2.3E-03 |  |  |
|  | quired pH | djustments: | 2.7 | for recyc. | $\mathrm{pH}=7.5$ | Total | 426.3 | kg/day |



## Test Conditions

| Raw water origin: | No name creek |  |  | NF | R.O. | B/C |
| :--- | :---: | :---: | :---: | :---: | :---: | :---: |
| Volume 8\# vented: | 0.0 | $\mathrm{~m} 3 /$ day | \# stages | $\mathrm{N} / \mathrm{A}$ | 1 | $\mathrm{~N} / \mathrm{A}$ |
| Softener type: | lon-X |  | Water temp (C) | $\mathrm{N} / \mathrm{A}$ | 35 | $\mathrm{~N} / \mathrm{A}$ |
| Water recycling: |  | Removal |  |  |  |  |
| -Brine disposal | Deep well | monovalent | $\mathrm{N} / \mathrm{A}$ | $96 \%$ | $\mathrm{~N} / \mathrm{A}$ |  |
| -Water recycled? | yes | divalent | $\mathrm{N} / \mathrm{A}$ | $98 \%$ | $\mathrm{~N} / \mathrm{A}$ |  |
|  |  |  | TOC | $\mathrm{N} / \mathrm{A}$ | $90 \%$ | $\mathrm{~N} / \mathrm{A}$ |
|  |  |  |  | Water recovery: |  | $\mathrm{N} / \mathrm{A}$ |


|  | Nanofiltration |  |  | Reverse Osmosis |  |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  | unit 1 | unit 2 | unit 3 | unit 1 | unit 2 | unit 3 |
| Water recoveryBrine recycle | N/A | N/A | N/A | 50\% | N/A | N/A |
|  | N/A | N/A | N/A |  |  |  |
| \% recycle | - | N/A | N/A |  |  |  |
| Blending | 10, \%3 | M \% \% \& \& | N/A |  |  |  |
| \% of prod | from \# | into \#3 | N/A |  |  |  |
|  | N/A | N/A | N/A | 5.8 | N/A | N/A |
| adjustment | N/A | N/A | - | 1.8 | N/A | N/A |
| Feedwater Ca | N/A |  |  | 75 | \#\#, |  |
| adjustment | N/A |  |  | 11 |  |  |

Results

| Solubility | Check | Nanofiltration |  |  | Reverse Osmosis |  |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  |  | unit 1 | unit 2 | unit 3 | unit 1 | unit 2 | unit 3 |
|  | CaSO4 | N/A | N/A | N/A | ok | N/A | N/A |
|  | CaCO 3 | N/A | N/A | N/A | ok | N/A | N/A |
|  | MgCO 3 | N/A | N/A | N/A | ok | N/A | N/A |
|  | $\mathrm{Mg}(\mathrm{OH}) 2$ | N/A | N/A | N/A | ok | N/A | N/A |


| Water flows |  |  |  |  |  |  |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  | F.E.T. inlet | Lime blow. Soften blow. | T.water | R.O. inlet | R.O. waste $\mathrm{B} / \mathrm{C}$ feed |  |  |
|  | 559.2 | 10.1 | 11.1 | 538.0 | 262.5 | 131.3 | $\mathrm{~N} / \mathrm{A}$ |


| Water qu | ity | Raw water | Recycled Water | Treated water | Current <br> T. water | Liquid | Final waste Solid |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| Flow | m3/day | 257.1 | 131.3 | 538.0 | - | 131.3 | - |  |
| Chem. |  |  | $\%$ of wastewater not recovered |  |  | 50.0\% |  |  |
| Ca | mg/L | 81.3 | 3 | 1.4 | 1.4 | $1.7 \mathrm{E}+02$ | $2.2 \mathrm{E}+01$ | kg/day |
| Mg | mg/L | 18.3 | 0.5 | 0.5 | 0.5 | $2.6 \mathrm{E}+01$ | $3.4 \mathrm{E}+00$ | kg/day |
| Fe | $\mathrm{mg} / \mathrm{L}$ | 0.39 | 0.1 | 0.4 | 0.6 | $3.1 \mathrm{E}+00$ | 4.0E-01 | kg/day |
| SO4 | $\mathrm{mg} / \mathrm{L}$ | 86.8 | 12 | 43.3 | 81.9 | $6.1 \mathrm{E}+02$ | $8.0 \mathrm{E}+01$ | kg/day |
| SiO 2 | mg/L | 5.2 | 0.2 | 0.9 | 0.9 | $7.7 \mathrm{E}+00$ | $1.0 \mathrm{E}+00$ | kg/day |
| PO4-P | $\mathrm{mg} / \mathrm{L}$ | 0.06 | 0.01 | 0.05 | 0.07 | 6.9E-01 | 9.1E-02 | kg/day |
| Cl | $\mathrm{mg} / \mathrm{L}$ | 2.9 | 26 | 7.5 | 2.4 | $6.3 \mathrm{E}+02$ | $8.3 \mathrm{E}+01$ | kg/day |
| Na | $\mathrm{mg} / \mathrm{L}$ | 12.5 | 34 | 37.2 | 50.5 | $8.1 \mathrm{E}+02$ | $1.1 \mathrm{E}+02$ | kg/day |
| DOC | $\mathrm{mg} / \mathrm{L}$ | 9.3 | 17.4 | 7.4 | 6.1 | $1.6 \mathrm{E}+02$ | 2.1E+01 | kg/day |
| pH |  | 7.9 | 7.5 | 10.0 | 10.1 | 6.1 |  |  |
| CO 3 | $\mathrm{mol} / \mathrm{L}$ | 1.4E-05 | 2.2E-06 | 1.4E-04 | 1.2E-04 | 5.6E-08 | 4.4E-04 | kg/day |
| HCO3 | $\mathrm{mol} / \mathrm{L}$ | 3.6E-03 | $1.4 \mathrm{E}-03$ | 3.0E-04 | 2.0E-04 | 8.5E-04 | $6.8 \mathrm{E}+00$ | kg/day |
| H 2 CO 3 | $\mathrm{mol} / \mathrm{L}$ | $1.0 \mathrm{E}-04$ | 1.1E-04 | 7.1E-08 | 3.7E-08 | 1.5E-03 | $1.2 \mathrm{E}+01$ | kg/day |
| Ctot | $\mathrm{mol} / \mathrm{L}$ | 3.7E-03 | 1.5E-03 | 4.4E-04 | 3.1E-04 | $2.3 \mathrm{E}-03$ |  |  |
|  | uired pH | djustments | 2.7 | or recyc. | $\mathrm{pH}=7.5$ | Total | 335.2 | /da |

Backeend model - fosiage RO, DW, NP, no pecycle
Test Conditions

| Raw water origin: | No name creek |  |  | NF | R.O. | $B / C$ |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| Volume 8\# vented: | 0.0 | m3/day | \# stages | 3 | 1 | N/A |
| Softener type: | NF |  | Water temp (C) | 40 | 35 | N/A |
| Water recycling: | Removal |  |  |  |  |  |
| -Brine disposal | Deep well |  | monovalent | 67\% | 96\% | N/A |
| -Water recycled? | no |  | divalent | 94\% | 98\% | N/A |
|  |  |  | TOC | 98\% | 90\% | N/A |
|  |  |  |  |  | ter recovery: | N/A |
|  | Nanofiltration |  |  | Reverse Osmosis |  |  |
|  | unit 1 | unit 2 | Unit 3 | unit 1 | unit 2 | unit 3 |
| Water recovery | 75\% | 50\% | 75\% | 50\% | N/A | N/A |
| Brine recycle |  | no | yes |  |  |  |
| \% recycle | - | 20\% | 100\% |  |  |  |
|  |  |  |  |  |  |  |
| \% of product from \#1+2 into \#3 50\% |  |  |  |  |  |  |
| Feedwater pH | 6.5 | 6.0 | 6.2 | 5.8 | N/A | N/A |
| adjustment | 2.7 | 0.8 | - | 1.8 | N/A | N/A |
| Feedwater Ca | 35.0 | 4, |  | 65 |  |  |
| adjustment | 0.0 |  |  | 50 |  |  |

Results

| Solubility Check | Nanofiltration <br>  <br>  <br> unit 1 <br> unit 2 |  |  |  | unit 3 |  |  |  | unit 1 | unit 2 | unit 3 |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  | ok | ok | ok | ok | $\mathrm{N} / \mathrm{A}$ | $\mathrm{N} / \mathrm{A}$ |  |  |  |  |  |
| CaCO 3 | ok | ok | ok | ok | $\mathrm{N} / \mathrm{A}$ | $\mathrm{N} / \mathrm{A}$ |  |  |  |  |  |
| MgCO 3 | ok | ok | ok | ok | $\mathrm{N} / \mathrm{A}$ | $\mathrm{N} / \mathrm{A}$ |  |  |  |  |  |
| $\mathrm{Mg}(\mathrm{OH}) 2$ | ok | ok | ok | ok | $\mathrm{N} / \mathrm{A}$ | $\mathrm{N} / \mathrm{A}$ |  |  |  |  |  |


| Water flows | F.E.T. inlet | Lime blow. Soften blow. | T.water | R.O. inlet | R.O. waste | B/C feed |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  | 576.1 | 13.9 | 89.9 | 472.2 | 279.3 | 139.7 | N/A |


| Water qual | ity | Raw <br> water | Recycled Water | Treated water | Current <br> T. water | Liquid | Final wast Solid |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| Flow | m3/day | 405.1 | 0.0 | 472.2 | - | 139.7 | - |  |
| Chem. |  |  | \% of wastewater not recovered |  |  | 50.0\% |  |  |
| Ca | mg/L | 81.3 | 0.00 | 0.3 | 1.4 | $2.3 \mathrm{E}+02$ | $3.2 \mathrm{E}+01$ | kg/day |
| Mg | $\mathrm{mg} / \mathrm{L}$ | 18.3 | 0.00 | 0.0 | 0.5 | $3.0 \mathrm{E}+01$ | $4.2 \mathrm{E}+00$ | kg/day |
| Fe | $\mathrm{mg} / \mathrm{L}$ | 0.39 | 0.00 | 0.0 | 0.6 | $3.1 \mathrm{E}+00$ | 4.3E-01 | kg/day |
| SO4 | $\mathrm{mg} / \mathrm{L}$ | 86.8 | 0.00 | 0.5 | 81.9 | 7.1E+02 | $9.9 \mathrm{E}+01$ | kg/day |
| SiO 2 | $\mathrm{mg} / \mathrm{L}$ | 5.2 | 0.00 | 0.0 | 0.9 | $8.3 \mathrm{E}+00$ | $1.2 \mathrm{E}+00$ | kg/day |
| PO4-P | $\mathrm{mg} / \mathrm{L}$ | 0.06 | 0.00 | 0.00 | 0.07 | 6.4E-01 | 9.0E-02 | kg/day |
| Cl | $\mathrm{mg} / \mathrm{L}$ | 2.9 | 0.00 | 0.7 | 2.4 | $1.6 \mathrm{E}+02$ | $2.2 \mathrm{E}+01$ | kg/day |
| Na | $\mathrm{mg} / \mathrm{L}$ | 12.5 | 0.00 | 3.2 | 50.5 | $4.9 \mathrm{E}+02$ | $6.9 \mathrm{E}+01$ | kg/day |
| DOC | $\mathrm{mg} / \mathrm{L}$ | 9.3 | 0.00 | 0.0 | 6.1 | $1.4 \mathrm{E}+02$ | $2.0 \mathrm{E}+01$ | kg/day |
| pH |  | 7.9 | 0.0 | 10.0 | 10.1 | 6.1 |  |  |
| CO 3 | $\mathrm{mol} / \mathrm{L}$ | 1.4E-05 | $0.0 \mathrm{E}+00$ | 8.1E-05 | 1.2E-04 | 5.5E-08 | 4.6E-04 | kg/day |
| HCO3 | $\mathrm{mol} / \mathrm{L}$ | 3.6E-03 | $0.0 \mathrm{E}+00$ | 1.7E-04 | 2.0E-04 | 8.4E-04 | $7.2 \mathrm{E}+00$ | kg/day |
| H2CO3 | $\mathrm{mol} / \mathrm{L}$ | 1.0E-04 | $0.0 \mathrm{E}+00$ | 4.0E-08 | 3.7E-08 | 1.5E-03 | $1.3 E+01$ | kg/day |
| Ctot | $\mathrm{mol} / \mathrm{L}$ | 3.7E-03 | $0.0 \mathrm{E}+00$ | 2.5E-04 | 3.1E-04 | 2.3E-03 |  |  |
|  | ired pH | djustments: | $\begin{aligned} & 2.7 \\ & 4.2 \end{aligned}$ | for recyc for T.W. from | $\begin{gathered} \mathrm{pH}=7.5 \\ \mathrm{NF}=10.0 \end{gathered}$ | Total | 267.8 | kg/day |

Backoen@ moodel - fostage RO, DW, NP, Pecycle
Test Conditions

| Raw water origin: | No name creek |  |  | NF | R.O. | $B / C$ |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| Volume 8\# vented: | 0.0 | m3/day | \# stages | 3 | 1 | N/A |
| Softener type: | NF |  | Water temp (C) | 40 | 35 | N/A |
| Water recycling: | Removal |  |  |  |  |  |
| -Brine disposal | Deep well |  | monovalent | 67\% | 96\% | N/A |
| -Water recycled? | yes |  | divalent | 94\% | 98\% | N/A |
|  |  |  | TOC | 98\% | 90\% | N/A |
|  |  |  |  |  | er recovery: | N/A |
|  | Nanofiltration |  |  | Reverse Osmosis |  |  |
|  | unit 1 | unit 2 | unit 3 | unit 1 | unit 2 | unit 3 |
| Water recoveryBrine recycle | 75\% | 65\% | 75\% | 50\% | N/A | N/A |
|  | - | no | yes |  |  |  |
| $\%$ recycle | *) | 20\% | 100\% |  |  |  |
| \% of produ | duct from \#1+ |  | 50\% |  |  |  |
| Feedwater pH adjustment | 6.5 | 6.0 | 6.2 | 5.81.8 | N/A | $\mathrm{N} / \mathrm{A}$ |
|  | 2.7 | 0.8 |  |  | N/A | $N / A$ |
| Feedwater Ca | 19.8 | U1/umuke |  | 70 | Mekumumektur |  |
| adjustment | 0.0 |  |  | 21 |  |  |

Results

| Solubility Check | Nanofiltration |  |  |  | Reverse Osmosis |  |  |  |
| ---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  | unit 1 | unit 2 | unit 3 | unit 1 | unit 2 | unit 3 |  |  |
| CaSO4 | ok | ok | ok | ok | $\mathrm{N} / \mathrm{A}$ | $\mathrm{N} / \mathrm{A}$ |  |  |
| CaCO 3 | ok | ok | ok | ok | $\mathrm{N} / \mathrm{A}$ | $\mathrm{N} / \mathrm{A}$ |  |  |
| MgCO 3 | ok | ok | ok | ok | $\mathrm{N} / \mathrm{A}$ | $\mathrm{N} / \mathrm{A}$ |  |  |
| $\mathrm{Mg}(\mathrm{OH}) 2$ | ok | ok | ok | ok | $\mathrm{N} / \mathrm{A}$ | $\mathrm{N} / \mathrm{A}$ |  |  |


| Water flows | F.E.T. inlet | Lime blow. Soften blow. | T.water | R.O. inlet | R.O. waste | B/C feed |  |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  | m3/day | 542.8 | 10.1 | 60.4 | 472.2 | 246.1 | 123.0 | $\mathrm{~N} / \mathrm{A}$ |


| Water qua | lity | Raw <br> water | Recycled Water | Treated water | Current <br> T. water | Liquid | Final waste Solid |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
| Flow | m3/day | 248.8 | 123.0 | 472.2 | - | 123.0 |  |  |
| Chem. |  |  | \% of wastewater not recovered |  |  | 50.0\% |  |  |
| Ca | $\mathrm{mg} / \mathrm{L}$ | 81.3 | 2.80 | 0.15 | 1.4 | $1.8 \mathrm{E}+02$ | $2.2 \mathrm{E}+01$ | kg/day |
| Mg | $\mathrm{mg} / \mathrm{L}$ | 18.3 | 0.55 | 0.01 | 0.5 | $2.7 \mathrm{E}+01$ | $3.3 \mathrm{E}+00$ | kg/day |
| Fe | $\mathrm{mg} / \mathrm{L}$ | 0.39 | 0.07 | 0.00 | 0.6 | $3.4 \mathrm{E}+00$ | 4.1E-01 | kg/day |
| SO4 | $\mathrm{mg} / \mathrm{L}$ | 86.8 | 13.47 | 0.33 | 81.9 | $6.6 \mathrm{E}+02$ | $8.1 \mathrm{E}+01$ | kg/day |
| SiO 2 | $\mathrm{mg} / \mathrm{L}$ | 5.2 | 0.17 | 0.01 | 0.9 | $8.2 \mathrm{E}+00$ | $1.0 \mathrm{E}+00$ | kg/day |
| PO4-P | $\mathrm{mg} / \mathrm{L}$ | 0.06 | 0.01 | 0.00 | 0.07 | 7.3E-01 | 8.9E-02 | kg/day |
| Cl | $\mathrm{mg} / \mathrm{L}$ | 2.9 | 7.77 | 1.02 | 2.4 | $1.9 \mathrm{E}+02$ | $2.3 \mathrm{E}+01$ | kg/day |
| Na | $\mathrm{mg} / \mathrm{L}$ | 12.5 | 23.57 | 3.74 | 50.5 | $5.7 \mathrm{E}+02$ | $7.0 \mathrm{E}+01$ | kg/day |
| DOC | $\mathrm{mg} / \mathrm{L}$ | 9.3 | 18.33 | 0.01 | 6.1 | $1.6 \mathrm{E}+02$ | $2.0 \mathrm{E}+01$ | kg/day |
| pH |  | 7.9 | 7.5 | 10.0 | 10.1 | 6.1 |  |  |
| CO 3 | $\mathrm{mol} / \mathrm{L}$ | 1.4E-05 | 2.3E-06 | 7.9E-05 | 1.2E-04 | 5.9E-08 | 4.4E-04 | kg/day |
| HCO3 | $\mathrm{mol} / \mathrm{L}$ | 3.6E-03 | $1.5 \mathrm{E}-03$ | 1.6E-04 | 2.0E-04 | 9.0E-04 | $6.8 \mathrm{E}+00$ | kg/day |
| H 2 CO 3 | $\mathrm{mol} / \mathrm{L}$ | 1.0E-04 | 1.1E-04 | 3.9E-08 | 3.7E-08 | 1.6E-03 | $1.2 \mathrm{E}+01$ | $\mathrm{kg} / \mathrm{day}$ |
| Ctot | $\mathrm{mol} / \mathrm{L}$ | 3.7E-03 | 1.6E-03 | 2.4E-04 | 3.1E-04 | $2.5 \mathrm{E}-03$ |  |  |
| Required pH adjustments: |  |  | $\begin{aligned} & 2.7 \\ & 4.2 \end{aligned}$ | for recyc. $\mathrm{H} 2 \mathrm{OH}=7.5$ <br> for T.W. from $N F=10.0$ |  | Total | 239.8 | kg/day |

## APPENDIX G

COST ESTIMATES

## COST ESTIMATING FACTORS

## Reverse osmosis systems

According to Osmonics, Inc. (located in Minnetonka, Minnesota), RO systems generally cost US $\$ 2000$ per gpm of produced permeate. Given their price estimate for a $360 \mathrm{~m}^{3} /$ day RO filter system (listed below), it was assumed that a dual-media filter in any RO assembly would account for 14.9 \% of the total cost. The remaining $85.1 \%$ would be involved in purchasing the RO membranes and associated support equipment.

| Equipment | Price (US\$) |
| :--- | :---: |
| Dual-media filter | 17500 |
| Chemical injection system | 4000 |
| RO skid | 80000 |
| Cleaning system | 9250 |
| Membranes | 7500 |
| Total | 118250 |

## Nanofiltration costs

Nanofiltration system generally cost $85 \%$ of the RO assemblies (Osmonics, Inc., personal com.). When no pretreatment equipment is required for the NF filters, as is the case when they are used in place of the ion-exchangers, they were assumed to cost $72.3 \%$ of RO systems.

## BCS assembly

BCS unit prices for three different flow rates were provided by Resources Conservation Company (located in Bellevue, Washington):

| Flow rate $\left(\mathrm{m}^{3} /\right.$ day $)$ | Cost (US\$) |
| :---: | :---: |
| 200 | 1650000 |
| 360 | 1900000 |
| 500 | 2250000 |

From a regression analysis on the three figures, BCS costs were estimated using the following equation:

$$
\text { price }=1990 * \text { feedwater flow }+1230235
$$

## Ion-exchangers

Nalco Canada Inc. representative indicated that an ion-exchange unit would cost between $\$ 50$ 000 and $\$ 100000$ Canadian dollars, regardless of the particular ZED configuration. A price of $\$ 50000$ US dollars was used in the cost estimates.

## COST ESTIMATES

## Closed-Loop Designs

## 1 -stage RO systems

| Equipment | BE, BCS, IX ${ }^{\text {a }}$ |  | BE, BCS, NFa |  | CD, BCS, IX ${ }^{\text {a }}$ |  | CD, BCS, NFa |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  | $\begin{aligned} & \text { Flow }{ }^{\text {F }} \\ & \left(\mathrm{m}^{3} / \text { day }\right) \end{aligned}$ | Cost (US\$) | $\begin{aligned} & \text { Flow }{ }^{\text {Flo }} \\ & \left(\mathrm{m}^{3} / \text { day }\right) \end{aligned}$ | $\begin{gathered} \text { Cost } \\ \text { (US\$) } \end{gathered}$ | $\begin{aligned} & \text { Flow }{ }^{\text {F }} \\ & \left(\mathrm{m}^{3} / \text { day }\right) \end{aligned}$ | $\begin{gathered} \text { Cost } \\ \text { (US\$) } \end{gathered}$ | $\begin{aligned} & \text { Flow }^{\text {b }} \\ & \left(\mathrm{m}^{3} / \text { day }\right) \end{aligned}$ | Cost <br> (US\$) |
| RO filters | 120 | 37453 | 112 | 35049 | 269 | 83992 | 257 | 80245 |
| RO pretreat: |  |  |  |  |  |  |  |  |
| Ca removal | - | 50000 | - | 50000 | - | - | - | - |
| Sand filter | - | 6558 | - | 6137 | - | - | - | - |
| BCS | 120 | 1.47 e 6 | 112 | 1.45 e 6 | 269 | 1.77 e 6 | 257 | 1.74 e 6 |
| NF filters | - | - | 472 | 125156 | - | - | 478 | 126906 |
| Total |  | 1.57 e 6 |  | 1.67 e 6 |  | 1.85 e 6 |  | 1.95 e 6 |
| with initial flows |  | 1.61 e 6 |  | 1.73 e 6 |  | 1.85 e 6 |  | 1.95 e 6 |

${ }^{\text {a }} \mathrm{BE}=$ back-end, $\mathrm{CD}=$ composite discharge, $\mathrm{IX}=$ ion-exchange, $\mathrm{NF}=$ nanofilters, $\mathrm{BCS}=$ brine concentrator and spray dryer
b flow $=$ produced permeate for RO and NF, and feedwater for the BCS unit

## 2-stage RO models

| Equipment | BE, BCS, IX $^{\text {a }}$ |  | BE, BCS, NFa |  | CD, BCS, $\mathrm{IX}^{\text {a }}$ |  | CD, BCS, NFa |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  | $\begin{aligned} & \text { Flow }{ }^{\text {b }} \\ & \left(\mathrm{m}^{3} / \text { day }\right) \end{aligned}$ | $\begin{gathered} \text { Cost } \\ \text { (US\$) } \end{gathered}$ | $\begin{gathered} \text { Flow }^{\text {b }} \\ \left(\mathrm{m}^{3} / \text { day }\right) \end{gathered}$ | $\begin{aligned} & \text { Cost } \\ & \text { (US\$) } \end{aligned}$ | $\begin{aligned} & \text { Flowb } \\ & \left(\mathrm{m}^{3} / \mathrm{day}\right) \end{aligned}$ | $\begin{aligned} & \text { Cost } \\ & \text { (US\$) } \end{aligned}$ | $\begin{gathered} \text { Flow } \\ \left(\mathrm{m}^{3} / \mathrm{day}\right) \end{gathered}$ | Cost <br> (US\$) |
| RO filters | 146 | 45711 | 135 | 42171 | 397 | 123880 | 382 | 119196 |
| RO pretreat: |  |  |  |  |  |  |  |  |
| Ca removal | - | 50000 | - | 50000 | - | - | - | - |
| Sand filter | - | 8004 | - | 7384 | - | - | - | - |
| BCS | 97.6 | 1.42 e 6 | 90.0 | 1.41 e 6 | 132 | 1.49 e 6 | 127 | 1.48 e 6 |
| NF filters | - | - | 472 | 125262 | - | - | 486 | 128976 |
| Total |  | 1.53 e 6 |  | 1.63e6 |  | 1.62 e 6 |  | 1.73 e 6 |
| with initial flows |  | 1.57 e 6 |  | 1.69 e 6 |  | 1.77 e 6 |  | 1.87 e 6 |

${ }^{\mathrm{a}} \mathrm{BE}=$ back-end, $\mathrm{CD}=$ composite discharge, $\mathrm{IX}=$ ion-exchange, $\mathrm{NF}=$ nanofilters, $\mathrm{BCS}=$ brine concentrator and spray dryer
b flow = produced permeate for RO and NF , and feedwater for the BCS unit

## BCS only configurations

| Equipment | BE, BCS, $\mathrm{IX}^{\text {a }}$ |  | BE, BCS, NFa |  |
| :---: | :---: | :---: | :---: | :---: |
|  | $\begin{aligned} & \text { Flowb } \\ & \left(\mathrm{m}^{3} / \text { day }\right) \end{aligned}$ | Cost <br> (US\$) | $\begin{gathered} \text { Flowb } \\ \left(\mathrm{m}^{3} / \mathrm{day}\right) \end{gathered}$ | Cost (US\$) |
| BCS | 219 | 1.67 e 6 | 223 | 1.67 e 6 |
| NF filters | - | - | 471 | 125050 |
| Total |  | 1.67e6 |  | 1.80 e 6 |
| with initial flows |  | 1.79 e 6 |  | 1.91 e 6 |

a $\mathrm{BE}=$ back-end, $\mathrm{IX}=$ ion-exchange, $\mathrm{NF}=$ nanofilters, $\mathrm{BCS}=$ brine concentrator and spray dryer
${ }^{\mathrm{b}}$ flow $=$ produced permeate for NF, and feedwater for the BCS unit

## Deep well configurations

## 1-stage RO options

| Equipment | BE, IX $^{\text {a }}$ |  | BE, NFa |  | CD, $\mathrm{IX}^{\text {a }}$ |  | CD, NFa |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  | $\begin{gathered} \text { Flow }^{\text {b }} \\ \left(\mathrm{m}^{3} / \text { day }\right) \end{gathered}$ | $\begin{gathered} \text { Cost } \\ \text { (US\$) } \end{gathered}$ | $\begin{gathered} \text { Flow }^{\mathrm{b}} \\ \left(\mathrm{~m}^{3} / \mathrm{day}\right) \end{gathered}$ | $\begin{gathered} \text { Cost } \\ \text { (US\$) } \end{gathered}$ | $\begin{gathered} \text { Flowb } \\ \text { (m³/day) } \end{gathered}$ | $\begin{gathered} \text { Cost } \\ \text { (US\$) } \end{gathered}$ | $\begin{aligned} & \text { Flow } \\ & \left(\mathrm{m}^{3} / \text { day }\right) \end{aligned}$ | Cost <br> (US\$) |
| RO filters | 131 | 40981 | 123 | 38421 | 514 | 160474 | 490 | 152871 |
| RO pretreat: |  |  |  |  |  |  |  |  |
| Ca removal | - | 50000 | - | 50000 | - | - | - | - |
| Sand filter | - | 7175 | - | 6727 | - | - | - | - |
| NF filters | - | - | 472 | 125262 | - | - | 480 | 127.331 |
| Tot |  | 98156 |  | 220410 |  | 160474 |  | '280 202 |
| with initial | flows - | 101660 |  | 226500 |  | 161582 |  | 281257 |

a $\mathrm{BE}=$ back-end, $\mathrm{CD}=$ composite discharge, $\mathrm{IX}=$ ion-exchange, $\mathrm{NF}=$ nanofilters
b flow $=\mathrm{m}^{3} /$ day of produced permeate

## 2-stage RO designs

| Equipment | BE, $\mathrm{IX}^{\text {a }}$ |  | BE, NFa |  | CD, $\mathrm{IX}^{\text {a }}$ |  | CD, NFa |  |
| :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: | :---: |
|  | $\begin{gathered} \text { Flow } \\ \left(\mathrm{m}^{3} / \text { day }\right) \end{gathered}$ | Cost <br> (US\$) | $\begin{aligned} & \text { Flow } \\ & \left(\mathrm{m}^{3} / \text { day }\right) \end{aligned}$ | $\begin{gathered} \text { Cost } \\ \text { (US\$) } \end{gathered}$ | $\begin{aligned} & \text { Flow }{ }^{\text {b }} \\ & \left(\mathrm{m}^{3} / \text { day }\right) \end{aligned}$ | $\begin{gathered} \text { Cost } \\ \text { (US\$) } \end{gathered}$ | $\begin{gathered} \text { Flow }^{\mathrm{b}} \\ \left(\mathrm{~m}^{3} / \text { day }\right) \end{gathered}$ | $\begin{gathered} \text { Cost } \\ \text { (US\$) } \end{gathered}$ |
| RO filters | 158 | 49346 | 148 | 46105 | 518 | 161817 | 496 | 155002 |
| RO pretreat: |  |  |  |  |  |  |  |  |
| Ca removal | - | 50000 | - | 50000 | - | - | - | - |
| Sand filter | - | 8640 | - | 8072 | - | - | - | - |
| NF filters | - | - | 472 | 125262 | - | - | 487 | 129082 |
| Total 107986 |  |  | 229439 |  | 161817 |  | 284084 |  |
| with initial flows 111992 |  |  | 236770 |  | 164287 |  | 286099 |  |

a $\mathrm{BE}=$ back-end, $\mathrm{CD}=$ composite discharge, $\mathrm{IX}=$ ion-exchange, $\mathrm{NF}=$ nanofilters


[^0]:    ${ }^{1}$ The time lapse between the collection of water flow and water chemistry data was due to the desire to fully analyze the flow data prior to water sampling. Developing a complete image of the plant's water distribution network before initiating the sampling program ensured that all of the pertinent water flows were sampled.

[^1]:    ${ }^{\text {a }}$ Refer to Appendix D for calculations

[^2]:    ${ }^{\text {a }}$ See Table 5.3

[^3]:    a Refer to Appendix D for calculations
    b ? = unknown source or sink; insig. difference $=$ insignificant difference; Ion-x regeneration $=$ sodium chloride used to regenerate the ion-exchangers

[^4]:    a Refer to Appendix D for calculations
    ${ }^{\mathrm{b}}$ precipitated $=$ missing mass settled out of solution within the lime ponds; ?= unknown source or sink; insig. difference $=$ insignificant difference

[^5]:    ${ }^{\text {a }}$ see Appendix G for cost calculations

[^6]:    , ${ }^{\mathrm{a}} \mathrm{Ca}=$ calcium, $\mathrm{Mg}=$ magnesium, $\mathrm{SO}_{4}=$ sulphate, $\mathrm{T} . \mathrm{Alk} .=$ total alkalinity, $\mathrm{Cl}=$ cloride, $\mathrm{DOC}=$ dissolved organic carbon

[^7]:    ${ }^{1}$ Simulator output indicated that salt precipitation in an RO unit was inevitable without inlet softening, and the presence of algae and other suspended solids in the FNR discharge will necessitate a RO pre-filter.

[^8]:    ${ }^{\text {a }}$ R.W. = raw water; Calcium pretreat. = calcium pretreatment requirements; RO f.water $=$ feedwater flow to reverse osmosis filters; $\mathrm{B} / \mathrm{C}$ f.water $=$ feedwater flow to brine concentrator
    ${ }^{\mathrm{b}} \mathrm{T} / \mathrm{O}=$ thermo-oxidizer; $\mathrm{E} / \mathrm{P}=$ effluent plant

[^9]:    $\begin{array}{lll}\text {-If using direct composite discharge reuse scenario: } \\ \begin{array}{l}\text { Flow }\end{array} & \\ & \text { D16 } & =039 \\ \mathrm{Ca}-\mathrm{C}_{\mathbf{T}} & \text { D18-D32 } & \text {-Same } \\ & & =041 \\ & & - \text { These }\end{array}$
    
    $\underset{\sim}{\sim}$
    F16 $=\mathrm{H} 5$ $\substack{\text { Inlet water } \\ \text { Flow } \\ \mathrm{Ca}-\mathrm{pH} \\ \mathrm{CO}_{3} \\ \mathrm{HCO}_{3}}$ $=\mathrm{C} 16+\mathrm{D} 16$
    $=$ Raw water + recycled water
    $=(\mathrm{C} 18: \mathrm{C} 26 * \mathrm{C} 16+\mathrm{D} 18: \mathrm{D} 26 * \mathrm{D} 16) / \mathrm{E} 16$ $=\mathrm{C} 16+\mathrm{D} 16$
    $=$ Raw water + recycled water
    $=(\mathrm{C} 18: \mathrm{C} 26 * \mathrm{C} 16+\mathrm{D} 18: \mathrm{D} 26 * \mathrm{D} 16) / \mathrm{E} 16$ - total contamination mass in the raw and recycled waters/the combined water
    volume; formula has more general form of $\left([]_{1} \mathrm{Q}_{1}+[]_{2} \mathrm{Q}_{2}\right) /\left(\mathrm{Q}_{1}+\mathrm{Q}_{2}\right)$ where []$=$ concentration; $\mathrm{Q}=$ flow, $1=$ raw water, and $2=$ recycled water $=\left(\left(4.8^{*} 10^{\wedge}-11\right)^{* E 30}\right) /\left(10^{\wedge}-\mathrm{E} 27\right)$
    $=\mathrm{CO}_{3}-\mathrm{HCO}_{3}$ equilibrium constant $\mathrm{x}\left[\mathrm{HCO}_{3}\right] /(10-\mathrm{pH})$ $=\left(\left(4.8^{*} 10^{\wedge}-11\right)^{*} \mathrm{E} 30\right) /\left(10^{\wedge}-\mathrm{E} 27\right)$
    $=\mathrm{CO}_{3}-\mathrm{HCO}_{3}$ equilibrium constant $\mathrm{x}\left[\mathrm{HCO}_{3}\right] /\left(10^{-\mathrm{pH}}\right)$
    $=\mathrm{CO}_{3}-\mathrm{HCO}_{3}$ equilibrium constant $\times\left[\mathrm{HCO}_{3}\right] /\left(10^{-\mathrm{pH}}\right)$
    $=\mathrm{E} 32^{*}\left(1 /\left(\left(\left(10^{\wedge}-\mathrm{E} 27\right) /\left(4.2^{*} 10^{\wedge}-7\right)\right)+1+\left(\left(4.8^{*} 10^{\wedge}-11\right) /\left(10^{\wedge}-\mathrm{E} 27\right)\right)\right)\right)$
    $=\left[\mathrm{C}_{\mathrm{T}}\right] \times 1 /\left(10-\mathrm{pH} / \mathrm{H}_{2} \mathrm{CO}_{3}\right.$ equilibrium constant $+1+\mathrm{CO}_{3}-\mathrm{HCO}_{3}$ equilibrium
    $=\mathrm{E} 32^{*}\left(1 /\left(\left(\left(10^{\wedge}-\mathrm{E} 27\right) /\left(4.2^{*} 10^{\wedge}-7\right)\right)+1+\left(\left(4.8^{*} 10^{\wedge}-11\right) /\left(10^{\wedge}-\mathrm{E} 27\right)\right)\right)\right)$
    $=\left[\mathrm{C}_{\mathrm{T}}\right] \times 1 /\left(10^{-\mathrm{pH}} / \mathrm{H}_{2} \mathrm{CO}_{3}\right.$ equilibrium constant $+1+\mathrm{CO}_{3}-\mathrm{HCO}_{3}$ equilibrium constant $/ 10-\mathrm{pH}$ )

    - from Benefield et al. 1982
    $=(\mathrm{C} 32 * \mathrm{C} 16+\mathrm{D} 32 * \mathrm{D} 16) / \mathrm{E} 16$ / the combined water value
    -Same value as Cell 039
    $=041-055$
    - These cells have the sam
    - These cells have the same values as cells 041 through 055 , respectively $=\left(\left(4.8^{*} 10^{\wedge}-11\right)^{*} \mathrm{E} 30\right) /\left(10^{\wedge}-\mathrm{E} 27\right)$
    $=\mathrm{CO}_{3}-\mathrm{HCO}_{3}$ equilibrium constant $\mathrm{x}\left[\mathrm{HCO}_{3}\right] /\left(10^{-\mathrm{pH}}\right)$
    / the combined water

[^10]:    = "N/A" (e.g. = IF(D14 = "Composite", 'Reverse Osmosis'!\$N\$45:\$N\$53, "N/A"))
    $=$ Specific values observed in mass balances constructed with the original data set
    model:
    $=\mathrm{J} 16$ $=$ P18 Current treated water levels (i.e. $1.4 \mathrm{mg} / \mathrm{L}$ )
    Based on assumption that ion-exchanger product waters always have $[\mathrm{Ca}]=1.4$
    $\mathrm{mg} / \mathrm{L}$
    mgL

    - Stipulates that ion-exchanger product waters will always have $\mathrm{a}[\mathrm{Mg}]<$ or $=0.5$ $\mathrm{mg} / \mathrm{L}$
    - Based on assumption that ion-exchangers cannot lower [Mg]'s below $0.5 \mathrm{mg} / \mathrm{L}$,
     when feedwater [ Mg ] is already $0.5 \mathrm{mg} / \mathrm{L}$
    $=\mathrm{N} 20-\mathrm{N} 23 \& \mathrm{~N} 26$
    - [ ]'s in blowdown flow
    $=[\mathrm{Cl}]$ of ion-exchanger blowdown
    
    M23 \& M26
    $\sum_{\sum}^{N} \sum_{\sum}^{N}$


    ## M19 <br> M16 <br> M18

    If Using a Back-End ZED
    Flow $-\mathrm{C}_{\mathrm{T}}$
    Kcenario:
    K16-32 Sources $\&$ Sinks
    $\mathrm{Ca}-\mathrm{C}_{\mathrm{T}}$

    Softener Product
    Flow
    a
    Mg
    $\mathrm{Fe}-\mathrm{PO}_{4} \& \mathrm{DOC}$
    U $\quad \underset{Z}{\sim}$

[^11]:    -This portion of the "Lime Blowdown Calc." worksheet combines information from the Water Treatment Manual and the "Overview" spreadsheet to predict the quality and quantity of blowdown and product waters exiting the hot lime treaters. Written notes within the
    worksheet itself were felt to be sufficient to guide one though most of the involved equations. The values displayed in certain cells may, however, still need some further explanation. spreadsheet to predict the quality and quan

[^12]:    $=$ values in cells I14 through I33 if only one NF stage used, or M14 through M33 if two NF units used
    $=$ values in cells U14 through U33
    

[^13]:    RO UNIT\#2
    if more than 1 RO unit is used in the ZED treatment train:
    
    = Same contaminant []'s as Unit \#1 brine flow
    $=\mathrm{H} 26$ - 126
    $=\left[\mathrm{C}_{\mathrm{T}}\right] \times 1 /\left(10-\mathrm{pH} / \mathrm{H}_{2} \mathrm{CO}_{3}\right.$ equilibrium constant $+1+\mathrm{CO}_{3}-\mathrm{HCO}_{3}$ equilibrium constant $/ 10-\mathrm{pH}$ ) -From Benefield et al. 1982
    $=\left(\left(10^{\wedge}-\mathrm{J} 26\right) * \mathrm{~J} 30\right) / \mathrm{J} 6$
    $=(10-\mathrm{pH}) \times\left[\mathrm{HCO}_{3}\right] / \mathrm{H}_{2} \mathrm{CO}_{3}-\mathrm{HCO}_{3}$ equilibrium constant
    $=\mathrm{H} 34$

