Red Dog Mine
Closure and Reclamation Plan

SD E4: Assessment of Water Treatment Methods
Applicable for Closure

Assessment of Water Treatment Methods Applicable for Closure Red Dog Mine, Alaska

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Executive Summary

This report was prepared as part of the development of a closure and reclamation plan for the Red Dog Mine, and presents the results of Task W4 - Assess Water Treatment Methods. Specifically, the report presents an evaluation of water treatment methods that could conceivably be used during and after closure of the mine.

The major wastewater sources after closure of the mine are expected to be the main waste stockpile, the pits, and any waste rock backfilled into the pits. Planning for closure of those areas is still underway. Under the assumption that all runoff from these sources will require treatment, it is estimated that the total (effluent) flow would be 1500 million gallons per year.

The acidity loadings from all sources at the site were initially assumed to be the same as estimated in the Suspension Study Report. That report uses year 2003 loadings as a basis. Even with the assumption that the sources remain at current levels, acidity concentrations in the tailings pond are predicted to rise significantly after closure. The reason is the cessation of water reclaim to the mill and the associated treatment in WTP#1. The current tailings pond average TDS, sulfate and zinc concentrations of 3800, 2500 and 300 mg/L, respectively, are estimated to reach post-closure steady state levels of 5100, 3500 and 750 mg/L unless source controls or treatment of mine water and waste stockpile seepage was undertaken before discharge into the tailings pond.

Six categories of wastewater treatment methods were identified as potentially applicable: chemical treatment methods, membrane processes, ion exchange, evaporation/distillation, biological sulfate removal, and freeze separation. Within these categories, a total of 18 methods were identified and briefly reviewed. The results indicate that there are no reasonable alternatives to the use of lime for effluent treatment at Red Dog. The High Density Sludge (HDS) lime treatment process is the preferred treatment process. It produces high quality effluents more cost-effectively and reliably than other treatment processes.

Four applications of the HDS process were considered. The preferred variant is the separate treatment of high strength wastewaters at source. Treatment of mine sump water and waste stockpile drainage prior to mixing with the tailings pond would reduce TDS levels by about 30% over current levels.

The future cost for effluent treatment is very sensitive to changes in the assumed wastewater flow and acidity. The net present value (NPV) of perpetual treatment at the assumed annual flow of 1500 million gallons and acidity loading of 12,000 tonnes is estimated to be $98 million (using a discount rate of 4%). If source controls are applied (e.g. covers over the waste stockpiles) and a 50% reduction in the acidity loading is achieved, the NPV treatment cost would reduce by about $23 million. A 75% reduction would decrease the NPV by approximately $34 million. Conversely, a doubling of the acidity loading would increase the NPV treatment cost by approximately

$45 million. These results indicate the potential value of the source control methods that are being evaluated in other closure planning tasks.

Long-term management of sludges produced during water treatment will be an important consideration. Assuming a high density sludge is produced at 20% solids, treatment at the current acidity loading would produce about 104,000 yd^3/year of sludge. Options for managing the sludge are discussed in a separate report.

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1   Introduction

Teck Cominco Alaska Incorporated (TCAK) is in the process of developing a closure and reclamation plan for the Red Dog Mine. As part of that process, and as part of continuing optimization of the site, TCAK is investigating wastewater treatment systems that could replace or compliment the existing lime treatment systems.

The Red Dog Mine annually discharges about 1.5 billion gallons of water via Outfall 001 to the Middle Fork of Red Dog Creek. This water is seasonally treated through WTP# 2 (HDS lime treatment plant) at a controlled rate to ensure the regulatory limits are not exceeded in the downstream receiving waters. The key constituents of concern in the wastewater at Red Dog arise from acid generation of the sulfidic ore and waste and include acidity, metals and TDS. Over time, TDS levels in the treated discharge have risen and methods to control TDS levels have and continue to be investigated (e.g. Andrews, 1997).

This study focuses on wastewater treatment methods that could be applied to control TDS levels in discharges from the site after closure.   The purpose of the study is to:

1) Provide an overview of water treatment methods that are available and assess whether there are practical alternatives that should be considered for effluent treatment at the time of mine closure.

2) Provide screening level cost estimates that can be applied in the evaluation of closure options and costs.

2   Key Information

2.1   Wastewater Characteristics

There are three primary sources of wastewater that are currently intercepted and managed at the site. They are drainage from the main waste stockpile and its surrounding areas, mine water from the main pit, and water used to transport concentrator tailings to the tailings pond. These streams are combined in the tailings pond. Reclaim water from the pond is either, treated and sent to the concentrator as process water, or treated and discharged to Red Dog Creek. The first two sources will continue after closure of the mine. The current characteristics of the major wastewater sources and reclaim water are shown in Table 2.1.

Table 2.1: Typical Wastewater Characteristics

Feed Type	pH	Zn (mg/L)	Pb (mg/L)	Cd (mg/L)	Fe (mg/L)	Ca (mg/L)	Mg (mg/L)	SO$_4$ (mg/L)	TDS (mg/L)
Reclaim	4.8	298	2	4	9	581	94	2460	3825
Mine Sump	3.7	1678	3	44	574	286	151	5530	8225
Waste Stockpile Drainage	2.9	3343	1	31	319	308	686	12470	15425

The waste stockpile drainage is estimated to be the source of about 40 to 50% of the acidity loading in the system. The other primary source of acidity is the mine water pumped from the pit which is estimated to contribute about 30 to 40% of the acidity loading.

The wastewaters are treated in two treatment plants, WTP #1 and #2. WTP #2 treats reclaim water from the tailings pond during the open water season and discharges it to Outfall 001. Lime is added to precipitate base metals and sodium sulfide is added to reduce cadmium levels. The effluent is filtered through sand filters to remove suspended solids before release. WTP #1 operates year-round, treating tailings reclaim and/or mine sump water to supply process water to the concentrator. Both plants produce a low density sludge that is discharged to the tailings basin.

At mine closure, WTP#2 would likely be retained and operations at WTP#1 would be suspended. If a new HDS plant is not constructed to treat mine water and waste stockpile drainage, then the tailings pond water would become much more acidic and contain higher levels of TDS and sulfate. WTP #1 currently adds about 2/3 of the total amount of lime used in treating reclaim water at the site. Without this lime addition, the acidity and sulfate levels in the tailings pond will climb.   With no changes to mine water and waste stockpile drainage, future tailings pond quality is expected to be as shown in Table 2.2. This water would be the future feed for WTP#2. The higher sulfate and zinc levels would lead to more gypsum and zinc hydroxide precipitation, and sludge production at the plant would rise dramatically.

Table 2.2: Current and Post-Closure Tailings Pond Water Quality

Feed Type	pH	Zn (mg/L)	Pb (mg/L)	Cd (mg/L)	Fe (mg/L)	Ca (mg/L)	Mg (mg/L)	SO$_4$ (mg/L)	TDS (mg/L)
2004 Reclaim	4.8	298	2	4	9	581	94	2460	3825
After Mine Closure[*]	2-3	750	2	20	250	360	190	3530	5120

* Assumes no application of source treatment or control measures

2.2   TDS Control Opportunities

Experience with WTP#2 indicates that the TDS reduction when treating reclaim water is modest, typically about 5%. However, there are opportunities to treat source water, which will result in reduced TDS concentrations in the tailings pond and, subsequently, in discharge water. These include:

- Separate treatment of mine sump water - TDS reduction is typically expected to be about 35% as a result of increased precipitation of gypsum from the higher sulfate levels in the feed.

- Separate treatment of waste stockpile drainage - TDS reduction of greater than 65% is projected from waste waters where the influent sulfate level exceeds 10,000 mg/L. A single test on lime treatment of MWD-18 reduced TDS levels from 36,000 to 2,800 mg/L (92% reduction).

Source control opportunities are also being evaluated as a part of closure planning. These include physical and chemical acid rock drainage control methods that have the potential to greatly reduce TDS concentrations and overall treatment costs.

2.3   Overall Site Acidity Production

Acidity is a measure of the amount of lime, or other alkalinity source, reported as units of CaCO$_3$ equivalent required to neutralize a solution to a pH of about 8.3. Current lime use at Red Dog is approximately 10,500 tonnes per year (as CaO).

In future, acidity loadings are likely to rise as the waste stockpile and mine footprints are expanded. However, this increase will be offset at closure by the cessation of ore processing. Based on the current load balance, the ore processing circuit contributes on average about 1,200 tonnes per year of net acidity.

3   Potential Treatment Options

3.1   Overview

3.1.1   General

A number of treatment options could be considered as stand alone treatment systems or as add-ons to the existing plants.   Options considered for application at Red Dog were assembled from information in the following reports:

1)   Mine Closure Workshop 2004. The closure option workshop suggested several options be considered including wetlands, RO, IX, evaporation, chemical precipitation (lime and barium salt), freezing, and waste segregation/treatment. The issue of sludge disposal was also raised with options to consider including concentration methods, co-disposal with tailings or waste rock, pit disposal, freeze concentration and metals recovery.

2)   "Treatment of Sulphate in Mine Effluents". Report to the International Network for Acid Prevention (INAP) (LORAX, 2003).   This report is a recent review on sulfate removal from mine effluents. Given that sulfate is the primary contributor to TDS levels at Red Dog, this review is very relevant.

3)   "Effluent Treatment and Water Management for TDS Control". G. Andrews, 1997.

4)   Internal SENES reports on acidity and TDS control.

3.1.2   Potential Treatment Needs

There are three potential applications that have been considered.

- A treatment plant to replace the entire existing system

- A treatment plant to manage only selected process streams (for example waste stockpile drainage and/or mine sump water)

- A treatment system add-on to treat or polish all or a portion (for cost estimates assumed to be 30%) of the effluent,.

3.1.3   Design Requirements

3.1.3.1 Design Flows

Initial estimates of flow rates for the streams that could require treatment at mine closure are summarized in Table 3.1. The design flows for treatment thus range from 1.5 billion US gal/year for the entire site run-off to 0.21 billion US gal/year for waste stockpile drainage only. Mine water with or without waste stockpile drainage was considered for separate treatment. For the purposes of

assessing add-on treatment processes, it was assumed that a 25% reduction in TDS would be required. To achieve this reduction, it was assumed that about 30% of the treated effluent would need to undergo additional treatment for TDS removal.

Table 3.1: Treatment Plant Design Flows

Source	Annual Flow 10^6 US gal
1) Entire discharge	1500
2) Mine Water sump	410
3) Waste Stockpile Drainage	210
4) Mine Water + Waste Stockpile Drainage	620
4) Effluent Polishing for TDS reduction	450

3.1.3.2 Source Concentrations

The key constituent concentrations in the wastewater streams are shown in Table 3.2.

Table 3.2: Estimated Constituent Concentrations

Source	Annual Flow (10^6 US gal)	TDS (mg/L)	SO_4 (mg/L)	Zinc (mg/L)
Reclaim 2004	1,500	3,825	2,460	300
Reclaim at Closure	1,500	5,120	3,530	750
Mine Water Sump	410	8,230	5,240	1,680
Waste Stockpile Drainage	210	15,500	10,000	3,350
Mine Water+ Waste Stockpile Drainage	620	9,700	5,900	2,600
Influent to Polishing Systems	450	3,300	2,460	<1

3.2   Review of TDS Control Methods

The potentially applicable treatment processes can be placed into one of six primary categories:
1) Chemical treatment, 2) Membrane processes, 3) Ion exchange systems, 4) Evaporation/distillation systems, 5) Biological sulfate removal, and 6) Freeze separation. The following section reviews the potential for each category of processes to be used to treat the waste water flows and characteristics described in Tables 3.1and 3.2.

3.2.1 Chemical Treatment Systems

A number of chemical reagents have been proposed for TDS control. These include lime/limestone, barium compounds and aluminum salts.

3.2.1.1 Lime and Limestone

Lime and/or limestone have been used to reduce TDS levels through pH adjustment and the precipitation of gypsum and metal hydroxides/carbonates. Methods to achieve this include:

a) Conventional lime/limestone treatment producing a low density sludge

b) High Density Sludge (HDS) lime treatment systems which tend to produce lower TDS levels through improved gypsum precipitation and also resulting in a much higher sludge density

c) Lime softening where lime is added to precipitate gypsum and magnesium hydroxide followed by calcite precipitation with the addition of CO_2 gas.

Lime treatment is the current state-of-the-art for metals and TDS control for management of acidic mining effluents.   Specific applications would include the ongoing use of WTP#2 and the construction of a new treatment plant to treat the primary sources of acidity (the mine water and waste stockpile drainage).

Costs for HDS treatment at WTP#2

The treatment of effluent at WTP#2 at closure has an estimated annual cost of about $2.61/1000 gal of effluent discharged. This is an all found cost which includes all labor, maintenance, power, consumables to operate the plant and a capital equipment replacement allowance (see section 4 for details).

Potential Costs for HDS Treatment of Mine Water and Waste Stockpile Drainage

Source control often proves to be environmentally sound and treatment of the major acidity sources at Red Dog (i.e. the mine water and waste stockpile drainage) is a reasonable option for consideration. The combined flow from these streams is 0.62 billion gallons per year with an estimated average TDS level of 9700 mg/L, sulphate level of 5900 mg/L and zinc level of 2600 mg/L.

The HDS plant to treat this wastewater would have a design similar to WTP#2 and would operate seasonally similar to WTP #2. The mine water and waste stockpile drainage could be stored in the pit over the winter months. Much of the plant could be salvaged from WTP#1 and the newly proposed WTP#3 which is under design to treat the high strength acid waters from the mine and waste dump. Therefore future capital costs are expected to be modest while operating costs for the site will increase as two plants would need to operated and maintained.

A first order cost estimate for the new HDS treatment plant in combination with WTP#2 (for mine water and waste stockpile drainage treatment) results in a capital cost of about $7 million and an incremental operating cost of about $0.40/1000 gal treated. This would add about $600,000/year to the operating costs for water treatment and would result in a sulfate reduction at closure from about 2300 mg/L to about 1700 mg/L. Given the marginal costs and substantial TDS reduction, treating the concentrated acid streams at source is a reasonable option for TDS management.

3.2.1.2 Barium Salts

The addition of barium hydroxide is an effective method for reducing TDS levels. Barium hydroxide can be used in place of lime to precipitate barium sulfate and other metal hydroxides. As a stand alone system, barium precipitation would not be cost effective. However as an add-on system in combination with lime treatment, barium precipitation could be used to reduce TDS levels to acceptable limits.

A second process uses barium sulfide to precipitate barite (LORAX 2003). The advantage of this process is the potential to recover barium for reuse and recover sulfur as a by-product. This process includes:

* Thermal reduction of barium sulfate to form barium sulfide

* Addition of barium sulfide to wastewater to precipitate barium sulfate

* Stripping and oxidation of hydrogen sulfide

* Metals precipitation with lime, followed by sludge removal.

It has also been noted that one of the limestone types available near the Red Dog Mine (Kogruk Limestone) appear to contain significant levels of barium. The use of barium carbonates in water treatment does not appear to have been tested elsewhere. There is a possibility that the limestone could be roasted to form a BaO/CaO which could be used in lieu of the existing CaO; however this is unlikely to prove to be an economic proposition. The possibility of using this limestone as a barium-rich neutralizing agent might be considered for further research.

Potential Costs for Barium Precipitation

Should barium be considered for use at Red Dog, the application would likely be as an add-on to the existing WTP#2 HDS plant for effluent polishing for sulfate reduction. The requirements for this facility would include:

* Building to house new facilities

* Chemical storage, makeup and feeding facilities for barium hydroxide

* Precipitation reactor similar to the lime precipitation reactors in the HDS circuit

* Thickener to settle and recover barium sulfate sludge

* Sludge return/disposal pumps

A first order estimate of costs suggests that to treat 30% of the discharge flow from WTP#2 would require a capital cost of about \$5 million and an operating cost of about \$26/1000 gal treated. This would add about \$10 million/year to the operating costs for water treatment and would result in a modest sulfate reduction at closure from about 2300 mg/L to about 1800 mg/L.

3.2.1.3 Aluminum Salt

CESR (Walhalla Process)

The Cost Effective Sulfate Removal (CESR) process involves the addition of calcium aluminate following lime neutralization. Dissolved salts (i.e. calcium sulfate) are removed by this process as all of the reagents added to the wastewater stream precipitate as a solid product. This procedure is also known as the Walhalla process and was originally marketed by Walhalla Technologies Inc. Removal is achieved by the precipitation of an insoluble sulfate mineral known as ettringite ($[Ca_3Al(OH)_6 \cdot 12H_2O]_2 (SO_4)_3 \cdot 2H_2O$). The precipitation reaction can be described by:

$$CaAl_2O_4(s) + 2Ca(OH)_2(aq) + 3CaSO_4(aq) + 30H_2O \rightarrow$$
$$[Ca_3Al(OH)_6 \cdot 12H_2O]_2 \cdot (SO_4)_3 2H_2O(s)$$
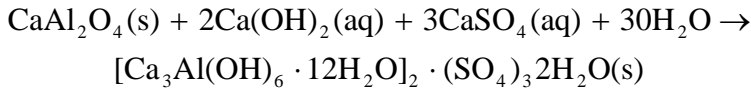
Mintek has a similar process and now market the process as SAVMIN (Sivilski, 2001). This process is an add-on to conventional lime treatment and is described below.

- Conventional lime precipitation/sludge removal to pH ~12

- Gypsum precipitation/sludge removal (de-saturation) with seeding crystals

- Ettringite precipitation/sludge removal using aluminum hydroxide

- Aluminum hydroxide recovery by pH adjustment with sulfuric acid

- Final pH control with CO_2 addition and calcite precipitation/sludge removal

Neither of the Walhala or Savmin processes has to our knowledge been applied at a large scale for treatment of mine effluents. However, such processes could be considered subject to laboratory and pilot plant trials. Based upon our understanding of these processes, the Walhalla process uses lime treatment as the first stage in treatment and as such could be applied as an add-on process for sulphate removal in the effluent from WTP#2.

Potential Costs for Aluminum Salt Precipitation

The Walhalla Process, if applied to treat 30 % of the effluent from the WTP#2, would require the following new facilities.

- Building to house new facilities

- Chemical storage, makeup and feeding facilities for calcium aluminate

- Precipitation reactor similar to the lime precipitation reactors in the HDS circuit

- Thickener to settle and recover ettringite sludge

- CO_2 addition for pH adjustment

- Thickener for aluminum hydroxide/ettringite

- Sludge return/disposal pumps

A first order estimate of cost for Walhalla (or Savmin) treatment plant that would treat 30% of the discharge flow from WTP#2 would be a capital cost of about $8 million and an operating cost of about $3.75/1000 gal treated. This would add about $1.5 million/year to the operating costs for water treatment and would result in a modest sulfate reduction at closure from about 2300 mg/L to about 1800 mg/L.

3.2.2 Membrane Processes

The following section has been adapted from SENES 1999.

A number of separation processes utilizing membranes have been developed and are typically used for the removal of dissolved salts from water. All membrane processes separate the feed stream into a permeate (the desalinated water that passes through the membrane) and a concentrate (the stream in which the salts are concentrated). Membrane processes can be distinguished based on the driving force for removal and the size of molecules rejected by the membrane. Reverse osmosis (RO) removes materials less than 0.001 μm (e.g. removal of monovalent salts); nano-filtration (NF) is used to separate larger molecules such as sugars and divalent salts; ultra-filtration (UF) can retain proteins and colloids in the 0.001 to 0.1 μm range; and micro-filtration (MF) is used for sterilization and removal of suspended material ranging in size from 0.1 to 10 μm.

Membrane systems are susceptible to fouling by particulates, gas bubbles and other fouling contaminants and therefore the feed must be pretreated. Fine particulates and colloidal silica are usually removed by micro-filtration. Scaling agents such as iron salts, silica, calcium sulfate and calcium carbonate can also cause problems as saturation conditions may be met at the membrane surface resulting in precipitation of solids on the membrane. Calcium removal by ion exchange or soda softening and the addition of anti-scaling agents are used to alleviate some of these scaling problems. Pre-treatment systems can be expensive and may account for as much as half of the desalination cost depending on the extent of pre-treatment required.

Another drawback to using membranes for wastewater treatment is the production of highly concentrated brine that must be treated prior to final disposal. The brine is often evaporated using a distillation and crystallization process or evaporation ponds (only practical in arid locations). Safe disposal of the dry salts can also be problematic and requires specially constructed disposal facilities. In some cases, brine may also be chemically treated.

Reverse Osmosis

Reverse osmosis (RO) utilizes a semi-permeable membrane to separate solutes from water. Osmosis is defined as the spontaneous movement of water through a semi-permeable membrane from a dilute solution to a more concentrated one. The movement of water through the membrane causes an increase in pressure in the concentrated solution. When the system attains equilibrium, the pressure difference between the dilute and concentrated solutions is called the osmotic pressure. A pressure difference across the membrane in excess of the osmotic pressure of the concentrated solution drives water through the membrane from the concentrated solution to the dilute solution.

The principles of reverse osmosis are utilized for the removal of dissolved salts from water and wastewater using flow-through systems. Typical operating pressures range from 1.4 to 5 MPa for brackish water desalination and 5 to 10 MPa for seawater desalination. Feed flow rate, concentration and type of dissolved species, operating pressure, temperature and pH are the important operating variables for RO. Water recovery in RO systems is generally 80 - 90% and salt rejection is approximately 80%. A number of stages may be required to achieve the desired effluent characteristics.

Membrane processes are not often used for the treatment of high strength liquid effluents such as those associated with mining due to high operating costs, pre-treatment requirements, gypsum scaling and the need for downstream treatment of concentrate brines. Operating costs are highly dependent on the characteristics of the water being treated and are primarily due to membrane cleaning and replacement costs and energy requirements to generate high pressures.   For all these reasons, RO is not a reasonable treatment option for Red Dog.

Electrodialysis

The process of electrodialysis reversal (EDR) uses alternating anion and cation selective membranes to concentrate wastewaters. A direct current is applied across a stack of ion selective membranes and causes ions to move toward the corresponding electrode. Anions will move toward the anode and can pass easily through the anion permeable membranes but cannot move through cation selective membranes and become trapped in the channel between membranes. Similarly, cations moving in the opposite direction toward the cathode accumulate in the same channel. The anode and cathode, as well as the brine and product channels, are periodically reversed, facilitating cleaning and reducing the potential for scaling of the membranes. This "self-cleaning" ability reduces the pre-treatment requirements and in some cases the only pre-treatment requirement is filtration for removal of fine particles. Gypsum fouling can be a problem in concentrated streams and lime softening may be required as part of the pre-treatment process.

EDR has some distinct advantages over RO, despite higher energy consumption. Pre-treatment of EDR feed can be reduced as the membranes are less susceptible to fouling and can operate over a broader range of temperature and pH. Construction and maintenance costs are reduced due to considerably lower operating pressures, eliminating the need for high pressure vessels. RO membranes undergo gradual compaction and need to be replaced more frequently, whereas EDR membranes generally do not. The high energy demands and lack of any proven use of EDR on a large scale make this process a poor candidate for consideration at Red Dog.

Testing of Membrane Processes

A process modification that has made RO more attractive for the treatment of acid mine waters is the incorporation of gypsum seed crystals in the feed stream. A pilot-scale operation was described by Juby (1992) that utilized seed crystals in a tubular RO configuration. Two pilot plants were operated (0.18 m^3/h and 1.8 m^3/h, 0.8 gpm and 7.9 gpm)) with promising results. Salt rejection of 80 to 90% was achieved with 85 to 95% water recovery. The drawbacks of this system included high energy

consumption, high gypsum recirculation rate and poor mass balance control of gypsum seed and brine. Further refinements to this system lead to the development of the Slurry Precipitation and Recycle Reverse Osmosis (SPARRO) process.

The SPARRO process was patented in 1988 and the operation of a pilot plant was described in Pulles et al. (1992). A pilot plant was constructed to produce 0.85 L/s (13.5 gpm) of potable water. Pre-treatment initially entailed neutralization to pH 10 using sodium hydroxide, oxidation with potassium permanganate, polymer flocculant addition, clarification, dual medium filtration and final pH adjustment with sulfuric acid to pH 5 to 6. After more than 4000 hours of operation, modifications were made to the pre-treatment circuit. Membrane degradation during the initial operating period was attributed to bacterial attack on the cellulose acetate membranes and subsequently chlorination was added as a pre-treatment step. Sodium hydroxide neutralization, potassium permanganate addition and flocculation were also discontinued at this time due to the high cost of these additives.

The SPARRO pilot plant was operated for a period of approximately 6000 hours with an average water recovery of 95%. Membrane flux rates decreased during the operating period by about 45%, possibly due to compaction or fouling of the membranes.

A pilot-scale test of EDR for the treatment of mine service water was described in Pulles et al. 1992. The service water contained high levels of dissolved sodium and chloride and was non-scaling with respect to gypsum.

A pilot plant was constructed to treat 5.8 m^3/h (25.5 gpm) and operated for about 6000 hours. Membrane foulants were removed by pre-treatment with potassium permanganate to oxidize iron and manganese, polymer flocculant addition, primary filtration using a dual medium filter and secondary filtration with a 10 µm cartridge filter. An anti-scaling agent was also added to prevent barium sulfate precipitation on the EDR membranes. The system was capable of achieving an average salt rejection of 80% and water recovery as high as 84%.

A comparison of the capital and operating costs for EDR, TRO and SPARRO for the desalination of mine waters with elevated gypsum levels was performed by Pulles et al. 1992. Costs were estimated for a plant capable of producing 46.3 L/s (730 gpm) of desalinated water at an operating water recovery of 90%. The cost figures presented for EDR and TRO include provisions for soda ash softening of the raw feed water to reduce the scaling potential of the water prior to desalination. The capital and operating costs per m^3 of product water are presented in Table 3.3 for the three alternatives. The cost of brine disposal was not included in the estimates but would be expected to be significant in all cases.

Table 3.3: Comparison Of Capital And Operating Costs Of RO Processes

(From SENES 1999-Adapted from Pulles et al. 1992)

Process	Capital Cost ($)	Operating Cost ($/m^3)	Operating Cost ($/1000 gal)
EDR	4,014,000	0.66	2.50
TRO	4,760,000	0.62	2.35
SPARRO	5,022,000	0.46	1.74

The SPARRO process, although not proven, could possibly be considered for use at Red Dog as an add-on process after lime treatment at WTP#2. However substantial laboratory and pilot scale investigation would be needed before this process could be considered as a practical option.

Potential Costs for SPARRO PROCESS

A seeded reverse osmosis plant treating an assumed 30% of the effluent stream would treat water after filtration in the sand filters. The RO plant would produce a low TDS effluent for discharge and a gypsum rich brine which would be discharged back to the tailings pond. The brine discharge to the basin would result in an increase in TDS/sulfate levels in the tailings pond. The WTP#2 plant would likely reduce sulfate levels back to about 2300 mg/L thus providing indirect chemical treatment for the brine.

The SPARRO plant would include:

- Building to house new facilities

- Seeded RO treatment module

- Cyclone to recover gypsum seed crystals from the brine

- Reactor to enhance gypsum seed crystal growth

- Recycle Pumps

- Brine disposal pumps

A first order estimate of costs for treating 30% of the discharge flow from WTP#2 would be a capital cost of $8-10 million and an operating cost of about $4/1000 gal treated. This would add about $1.5 to 2 million/year to the operating costs for water treatment and would result in a sulfate reduction at closure from about 2300 mg/L to about 1800 mg/L.

3.2.3 Ion Exchange

The following section has been adapted from SENES 1999.

Ion exchange is the reversible process of transferring undesirable ions (i.e. Cu^{2+}, Ni^{2+}, Ca^{2+}) from an aqueous solution to a solid support with the concomitant transfer of innocuous ions (i.e. H^+) from the solid to the solution. This process is widely used in a variety of industries for the purification of water (e.g. residential water softening) or the recovery of a valued component from solution (e.g. concentration of gold cyanide complexes).

Ion exchange resins consist of an insoluble support that has active functional groups attached to it. The functional groups are ionogenic (i.e. they are capable of dissociation) and are electrostatically balanced with mobile ions of the opposite charge. Undesirable ions in solution are substituted for these mobile ions during the ion exchange reaction. Cation exchangers are used to replace metals, and other cations in solution with H^+ or Na^+; anion exchangers substitute sulfate, phosphate, etc. with OH^- or Cl^-. Loaded resins are regenerated with acid or caustic, depending on the type of resin, to produce concentrated brine containing the ions removed from solution.

A number of natural materials exhibit ion exchange capabilities such as zeolites (a class of aluminosilicate minerals), coal, metallic oxides and some types of algae and bacteria. Hundreds of synthetic ion exchange resins have been developed and they tend to be more efficient than natural ion exchangers, have higher capacities and can be designed to be highly selective. The properties of a particular resin are largely defined by the functional groups attached to it. Ion exchange resins are commonly distinguished based on the nature of the functional groups. Strong acid cation exchange resins are derived from strong acids such as H_2SO_4, whereas weak acid resins are derived from weak acids such as H_2CO_3. Weak base anion resins contain functional groups composed of amines. Quaternary ammonium functional groups are used to generate strong base exchange resins. The equations below demonstrate the ion exchange or loading reactions (where the constituent is removed from solution to the resin) for a generic example of each type of resin; regeneration is performed simply by reversing the reaction (stripping of the constituent into a brine). The organic network of the synthetic resin is represented as R in the equations; $R=_3N$ represents a quaternary ammonium group where R' is an organic functional group such as CH_3.

1. Strong acid cation exchange, regenerate with strong acid such as HCl or H_2SO_4

$$2R - SO_3H + Ca^{2+} \rightarrow (R - SO_3)_2\,Ca + 2H^+$$
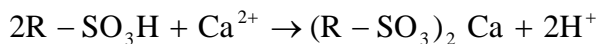
2. Weak acid cation exchange, regenerate with strong acid such as HCl or H_2SO_4

$$2R - COOH + Ca^{2+} \rightarrow (R - COO)_2\,Ca + 2H^+$$
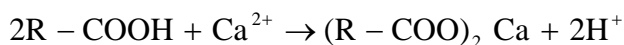
3. Strong base anion exchange, regenerate with strong base such as NaOH

$$2R - R_3NOH + SO_4^{2-} \rightarrow (R - R_3N)_2\,SO_4 + 2OH^-$$

4. Weak base anion exchange, regenerate with strong base such as NaOH
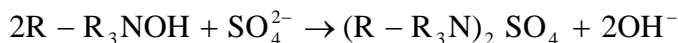
$$2R - NH_3OH + SO_4^{2-} \rightarrow (R - NH_3)_2\,SO_4 + 2OH^-$$
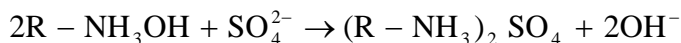
Regeneration of the loaded resin is accomplished by passing a concentrated solution of the counter-ion through the resin bed to reverse the equilibrium reactions shown above. This process produces a concentrated stream containing the contaminants. The concentrated stream often requires treatment prior to disposal.

In theory, ion exchange can be applied for the removal of any dissolved contaminants related to mining. Exchange resins have been developed for the removal of heavy metals, calcium and magnesium, radionuclides, sulfate and ammonium. Although ion exchange process can remove contaminants effectively, their application is generally limited by high capital and operating costs, process complexity requiring highly skilled operators and the need for additional treatment of spent regenerant.

Ion Exchange Applications

In general, ion exchange is not economically feasible for treating large volumes of concentrated wastewater due to the high cost of regeneration agents and the need to treat a concentrated regeneration brine. Ion exchange may have applications as a secondary treatment step to selectively remove specific contaminants such as heavy metals, ammonia, radionuclides or calcium sulfate. There are several systems that have been developed for this purpose.

The modified Desal process described by Dinardo (1991) utilizes a weak-base anion exchange resin to remove sulfate from solution in exchange for bicarbonate. Aeration, lime neutralization, settling and filtration are used to remove iron and other metals from the effluent stream following ion exchange. This process produces potable water but is limited to the treatment of waste streams containing less than 2,200 mg/L sulfate. The precipitation of ferric hydroxide in the anion exchange resin bed may be a problem in some cases. A demonstration plant was constructed in the early 1970s in Pennsylvania using the modified Desal process. The capital cost of the plant was $2.3 million and the operating costs in 1975 were $0.14/m^3 ($0.53 per 1000 gal) for 2,300 m^3/day (420 gpm) capacity.

The Sirotherm process was developed by CSIRO, a research organization in Australia. Sirotherm utilizes a polymeric resin having both anion and cation exchange capabilities. The system uses hot water rather than chemical regenerants to strip contaminants from the loaded resin. Pilot-scale testing has shown that the capital costs of the process are competitive with membrane processes. Operating costs are higher unless a source of cheap, low-grade heat is available. Sirotherm is not expected to be competitive for treating waste streams containing more than 2,000 ppm total dissolved solids (ANSTO 1994).

GYP-CIX is another ion exchange process that has shown promise for the removal of calcium and sulfate from wastewater. This technology is particularly useful for the treatment of effluent from a lime neutralization circuit. Lime neutralization of acid mine drainage produces a gypsum saturated solution that has a great propensity to cause scaling. The GYP-CIX ion exchange system alleviates scaling problems by removing calcium and sulfate from the treated water. The resins are regenerated using sulfuric acid and lime to produce a gypsum slurry waste stream. Resin fouling is minimized by seeding the regeneration stage with gypsum crystals on which further precipitation takes place.

The operation of a GYP-CIX pilot plant was described by Robertson and Röhrs (1995). The plant had a capacity of 24 m^3/day (4.4 gpm). The concentration of various constituents in the lime-neutralized feed to the GYP-CIX system and treated effluent water quality are shown in Table 3.4.

The first stage of the GYP-CIX process is the removal of metals and other cations. The neutralized acid mine drainage is gravity fed to a series of six fluidized bed reactors, the cation exchange resin flows in a counter current direction. Resin is regenerated using a 5% sulfuric acid solution in a batch reactor containing gypsum crystals. The water is degassed to remove carbon dioxide and sent to the anion exchange system. The feed is again gravity fed to a series of six fluidized bed reactors and contacted with the anion exchange resin in a counter current manner. The loaded anion exchange resin is regenerated using a 2% lime slurry that has been seeded with gypsum crystals.

The product water from the GYP-CIX process described by Robertson and Röhrs (1995) contained significantly less sulfate and calcium than the feed and was neutral in pH water recovery was more than 90%. Radium and uranium levels were also observed to decrease. Loss of resin was approximately 5 and 10% per annum for the anion and cation resin beads respectively. Operating costs for a 4000 m^3/day (730 gpm) plant were estimated to be about $0.37/$m^3$, with regeneration chemicals making up approximately 65% of the operating cost. Depending on site-specific factors, the sale of gypsum may off-set the cost of treatment.

Capital costs for a full scale GYP-CIX plant to treat 1,000 m^3/day (180 gpm) were estimated to be between Rand 2.5 and 3.5 million (CHEMEFFCO, undated). Operating costs for this plant were estimated to be Rand 2.50/m^3. (At the current exchange rate 1$US is approximately 6 Rand.)

Table 3.4: Feed and Effluent Water Quality After GYP-CIX Process

(Adapted from Robertson and Röhrs 1995)

Parameter	Units	Feed Concentration	Effluent Concentration
Total Dissolved Solids (TDS)	mg/L	3,000	500
pH		6-8	6-8
Calcium	mg/L	500	<50
Magnesium	mg/L	100	<20
Sodium	mg/L	400	<100
Sulfate	mg/L	1200	<200
Chloride	mg/L	250	<50
Nitrate	mg/L	60	<10
Radium	pCi/L	30	<1
Uranium	µg/L	1000	<20

Potential Costs for GYP-CIX

The GYP-CIX process is the only ion exchange process that would be considered as a treatment process to be used in lieu of HDS (WTP#2). The only advantage of this plant is the production of a low TDS effluent and as such would only be considered if dramatic reductions to TDS were required. However, before any consideration of the process is given, substantial laboratory and pilot scale investigations would need to be completed to assure that the plant can meet heavy metals effluent criteria as well as function to reduce TDS levels over the long term.

The GYP-CIX circuit is somewhat complex in that it includes two conventional lime precipitation circuits plus a cation and anion exchange loading/regeneration stages. The cation exchange system is regenerated with sulfuric acid and the anion exchange system is regenerated with lime. For planning purposes, the operating period is assumed to remain at 4 months with the same seasonal discharge as occurs with WTP#2. The gypsum underflow from regeneration stages would be disposed in the tailings pond and/or one of the closed pits. It is likely that the sulfide addition and sand filters would be retained to assure low total metals levels in the effluent discharge.

The GYP-CIX facility would include:

- Building to house new facilities

- Chemical storage, makeup and feeding facilities for acid. The existing lime facilities would be used.

- Cation exchange system

- Cation exchange regeneration system with gypsum thickener

- Anion exchange system

- Anion exchange regeneration system with gypsum thickener

- Sludge pumps

A first order estimate of cost for the GYP-CIX treatment plant that would replace WTP#2 would be a capital cost of about $15-20 million and an operating cost of about $3.30/1000 gal discharged. This would add more than about $1.0 million/year to the operating costs for water treatment but has the potential to result in a sulfate reduction at closure from about 2300 mg/L to <500 mg/L.

3.2.4 Evaporation/Distillation

The following section has been adapted from SENES 1999.

Evaporation is the removal of a volatile solvent from a solution. In the case of mine water treatment, the solvent is water. Evaporation processes for the treatment of mine effluents can be divided into two categories, solar and mechanical evaporation. Solar evaporation involves simple disposal of the effluent to a shallow pond where it is allowed to evaporate. Obviously, solar evaporation of water can only be achieved in locations where evaporation exceeds precipitation. Mechanical or thermal evaporation refers to the use of various types of heat exchange equipment to achieve the heat transfer necessary for evaporation. Mechanical evaporation of an effluent stream to dryness is a very costly treatment option due to high energy consumption. For this reason, evaporators are usually only used for treating concentrated streams (e.g. following reverse osmosis) or if water recovery is imperative.

Evaporation processes produce a crystalline solid product and water vapor. In some cases, it may be possible to market some of the solid products such as gypsum, sodium sulfate and sodium chloride and offset the cost of treatment. Disposal of the soluble salt product may present significant

problems and is often very expensive. Mechanical evaporation allows for the condensation of the vapor phase, which can then be recycled or discharged.

Evaporation Application

The only application where mechanical evaporation is known to have been applied at a mine was as an add-on process after a reverse osmosis plant. A full-scale RO/Evaporation plant was constructed to treat saline mine water from the Debiensko and Budryk mines in Poland (Ericsson and Hallmans 1996). A total mine water volume of 12,400 m^3/day (2270 gpm) with a total dissolved solids (TDS) concentration of 16,000 mg/L is treated by RO followed by thermal evaporation and crystallization. Pre-treatment of the feed includes algaecide dosing, flocculation and sedimentation, dual medium filtration and a two-stage microfilter (50 µm and 5 µm). The pH of the pretreated feed is adjusted using sulfuric acid and an anti-scaling agent is added prior to RO. The RO stage operates at 6 to 7 MPa and consists of several spiral wound RO units that can be operated independently. The permeate from the RO stage is decarbonated and used for drinking water. The brine is concentrated in a seeded, falling-film evaporator equipped with a mechanical vapor recompressor. The evaporator is seeded to allow the precipitation of calcium sulfate. The concentrated NaCl brine produced by the evaporator is sent to a thermal crystallizer. The crystallizer produces a solid NaCl product, distillate and a purge stream containing non-sodium chloride impurities. The purge stream is further treated to produce calcium sulfate, sodium chloride, carnallite and magnesium chloride. All solid chemicals recovered are sold to offset the cost of treatment.

The total investment cost of the entire plant was quoted as $60 million (Ericsson and Hallmans 1996). Operating costs were not given but energy consumption was approximately 57.4 kWh/m^3 (0.22 kWh/gal) of feed. Income is generated by the sale of salt, drinking water and other recovered chemicals, which off-sets the operating costs.

The high power demand and extremely high capital and operating costs make evaporative processes unsuitable for long term use at Red Dog. For comparison, at the energy consumption rate of 0.22 kWh/gal quoted above, the power cost at Red Dog would be roughly $22 per 1000 gallons, or 50 times the power cost for HDS treatment.

3.2.5 Biological Sulfate Reduction

3.2.5.1 Bioreactor Processes

Biological sulfate reduction has been applied for the removal of metals from contaminated waters. Systems that use biological reduction of sulfate to produce sulfide have many of the same positive and negative aspects as chemical sulfide precipitation, i.e. high removal of metals, possible recovery of saleable metal sulfide concentrates, H_2S gas emissions, sludge disposal issues. Bacterial sulfate reducing processes also have the added advantage of removing sulfate.

The Biosulfide process, patented by NTBC Research, separates the chemical precipitation step from the biological generation of H_2S. Pilot-scale testing at the Britannia mine in British Columbia was

conducted using a reactor system capable of treating 50 L/min (13 gpm), (Rowley et al. 1997). Copper levels were reduced from 12 mg/L to less than 0.001 mg/L, zinc concentrations of 16 mg/L were reduced to 0.37 mg/L. This system was capable of producing saleable grade copper concentrate; however the zinc sulfide concentrate was not of sufficient quality due to copper contamination. The capital costs of a full-scale treatment plant with a 17,000 m^3/day (3100 gpm) capacity was estimated to be \$2.2 million, the revenue generated by copper and zinc concentrate production would off-set the operating costs and this plant in theory would operate at a profit of \$130,000/year and would not produce any solid waste requiring disposal.

The first full-scale plant based on sulfate reducing bacteria (SRB) technology started operation in June 1992, in Budel-Dorplein, Netherlands, at the Budelco zinc refinery site (Scheeren et al. 1992 and Barnes et al. 1992). About 300 m^3/hour (1300 gpm) of groundwater is extracted from a combination of 12 shallow and deep wells and is currently being treated for the simultaneous removal of sulfate and metals by SRB, using ethanol as a source of carbon. The groundwater is contaminated with about 250 mg/L of zinc, 1,200 mg/L of sulfate and minor quantities of other metals. The final quality of the mine water treated at this facility has effluent concentrations of less than 0.1 mg/L for zinc, less than 0.2 mg/L for sulfate and less than 0.05 mg/L for cadmium, copper, iron and lead. In addition, the effluent pH varies between 6 and 7.

The Budelco process subjects the bioreactor to the entire acid mine drainage stream. This type of treatment system can reduce sulfate levels significantly but is more susceptible to fluctuations in the feed strength and flow. As well, the need for very large quantities of available carbon sources (e.g. ethanol) would make this treatment approach prohibitive at Red Dog.

A system using biological oxidation of iron is currently being used for the treatment of AMD from a closed sulfur/pyrite mine in Japan. The iron oxidizing bacteria, T. ferrooxidans, oxidizes Fe^{2+} to Fe^{3+} in four parallel reactors. The bacteria grow on the surface of ferric sulfate crystals that precipitate out of solution in the reactor. The bacteria-laden sludge is separated from the oxidized stream and recycled to the bioreactor. Neutralization of the effluent is achieved using ground limestone. Total iron levels are reduced from 300 to 400 mg/L to 2 to 3 mg/L and arsenic levels are reduced from about 1.8 mg/L to 0.01 mg/L.

For biological treatment to be considered at Red Dog, the plant would need to operate continuously as biological systems are not amenable to rapid start-up. Cold weather operation would require heating of the water which in itself would make this process uneconomic. Limited testing has been completed on with Red Dog waste waters with limited success.   As such, biological treatment would not be considered as a reasonable treatment option for long term treatment of waste waters from the Red Dog site.

3.2.5.2 Constructed Wetlands

There are two generic categories of constructed wetland systems, aerobic and anaerobic. Combinations of the two may also be considered. However, wetlands are not likely to be appropriate

technologies for TDS control at Red Dog due to the very high flows, cold temperatures and the requirement for large land areas for wetland development.

3.2.6 Other Physical Processes

3.2.6.1 Freeze Thaw Technologies

There are a number of technologies that take advantage of freeze concentration of contaminants (i.e. when water freezes, the salts remain in the brine and the ice is low in salt content). Although novel concepts have been conceived for using snow making and or freezing technologies for management of acid and contaminated streams, the process has not proven to be practical, particularly for very large scale applications. In order to apply the technology, areas need to be developed where solutions can be frozen and the brine recovered for separate management. This requires large areas to be created to contain the contaminated solutions for freezing and a treatment plant to manage the resulting brine. Even after freezing, there will be a need to polish the effluent as residual contamination in the ice is a certainty.

Another limitation is the rate of melting. When snow making was attempted at Red Dog, the resulting pile of snow required more than one summer to melt. The project was dropped from further consideration.

3.3 Comparison of the Options

Table 3.5 provides a brief summary of the pros and cons of treatment options addressed above with an assessment of the applicability to Red Dog Mine. The following appear to be the most attractive options for treating wastewater after closure of Red Dog:

1) HDS Lime treatment at WTP#2 as a stand alone process for control of all acidity, metals and TDS.

2) HDS Lime treatment of mine water and/or waste stockpile drainage as a separate stream.

The proven performance of the HDS process makes it clearly the most attractive option for treatment of the discharge water. Other options, in particular, the GYP-CIX ion exchange process, appear to offer the potential to reduce TDS levels but these processes are substantially unproven and appear to be complex and expensive. If TDS reduction beyond that possible with the stand-alone HDS plant is required, treating a separate stream of mine water and waste stockpile drainage in a new HDS plant is the preferred option for Red Dog.

Table 3.5: Summary Comparison of Treatment Options

Process	Pros	Cons	Applicable to Closure at Red Dog?
HDS Lime Treatment	Proven existing plant. Cost effective for metals/acidity control. Current state of the art for effluent treatment	Minor TDS reduction unless treating concentrated streams.	Yes. System is in place and working well. TDS removal and sludge density expected to increase at closure. Use of HDS to treat source waters could significantly further reduce TDS.
Barium Hydroxide Precipitation	Can produce very low sulfate levels. Simple add on process.	High cost.	No. Much less cost efficient than HDS options.
Barium Sulfide Precipitation	Can produce very low sulfate levels	SO_2 emissions. Unproven in Full scale. High cost for power to thermally reduce barium sulfate to sulfide. Process is complex.	No. Costly, unproven, and too complex for remote site.
Walhalla Process (CESR)	Can produce very low sulfate levels.	Unproven in full scale. Process is complex. High cost.	Possible add-on polishing system, but not as cost effective as HDS options.
SAVMIN Process	Can produce very low sulfate levels.	Unproven in Full scale. Process is complex. High cost.	Possible add-on polishing system, but not as cost effective as HDS options.
RO	Proven process. Can produce very low sulfate levels.	Membrane fouling Brine disposal. High power demand. High cost.	No. Costly, complex and not suitable for strong TDS waters or remote site.
Tubular RO	Can produce very low sulfate levels.	Unproven in full scale. Brine disposal. High power demand. High cost.	No. Costly, complex and not suitable for remote site.
EDR	Can produce very low sulfate levels.	Unproven in full scale. Membrane fouling. Brine disposal. High power demand. High cost.	No. Costly, complex and not suitable for strong TDS water or remote site.

Process	Pros	Cons	Applicable to Closure at Red Dog?
SPARRO	Can produce very low sulfate levels.	Unproven in full scale. Process is complex. Brine disposal. High cost.	No. Costly, complex and not suitable for remote site.
IX	Proven process. Can produce very low sulfate levels.	Gypsum scaling. Brine disposal. High cost.	No. Costly, complex and not suitable for strong TDS water or remote site.
GYP-CIX	Can produce very low sulfate levels.	Unproven on large scale Increased sludge production. Process is complex. Higher cost.	Possible, but lack of treatability data, track record at similar facilities makes it unattractive in comparison to HDS options.
Evaporation	Proven process. Cost effective in dry climates.	Disposal of salt. Requires warm dry climate.	No. Too power intensive and complex for remote mine application
Evaporation/ Crystallization	Produce high quality water.	Very high power demand. Disposal of salt. Highly complex Very high cost.	No. Too power intensive and complex for remote mine application
Biological Sulphate Reduction - Bioreactors	Proven process.	Nutrient source required. Not readily amenable to seasonal discharge. Potential high cost.	No. Not practical for seasonal start-up and costly to heat wastewater. Not practical to import ethanol
Biological Sulphate Reduction - Anaerobic Wetlands	Proven process. Suitable for low flow and low concentrations	Concentrations too high. Not readily amenable to seasonal discharge. Potential high cost.	No. Unproven and not practical.
Snow Making	Suitable for cold climate	Summer too short to melt snow, power intensive Brine disposal	No. Unproven.
Ice Making	Suitable for cold climate	Unproven for this application Brine disposal Impractical for large volumes	No. Unproven.

4 Long Term Treatment Options and Costs

4.1 Introduction

The above review of treatment processes has concluded that lime treatment using the HDS process remains the state-of-the-art for treatment of acidic wastewaters produced at the Red Dog site. The only outstanding issue is how best to treat effluents over the long term. At the present time, all wastewaters are directed to the tailings pond and then water reclaimed from the pond is treated in WTP#1 and WTP#2. Once the mine is closed, WTP#1 would not be required and it has been proposed that only WTP#2 would be operated.

The review also suggested that there may be advantages to separate treatment of high strength wastewaters at source. By treating the waste streams at source before addition to the tailings pond, there is the potential to precipitate more gypsum and reduce the TDS loadings in the final discharge. This option offers a cost effective method for reducing TDS levels below those achieved by the current arrangement.

In this section, combinations of the preferred options are developed and cost estimates prepared. The combinations selected for consideration are:

- Option 1 - Continued operation of WTP#2 to treat the total flow from the site,
- Option 2 - Continued operation of WTP#2 plus separate treatment of mine water using HDS,
- Option 3 - Continued operation of WTP#2 plus separate treatment of waste stockpile drainage using HDS,
- Option 4 - Continued operation of WTP#2 plus separate treatment of mine water plus waste stockpile drainage using HDS.

4.2 Wastewater Characteristics

The wastewater flows and constituent levels corresponding to the above options are shown in Table 4.1. These inputs were previously discussed in Section 3.1.3. For all options WTP#2 would need to treat 1500 million gal of water per year. The characteristics of the influent to WTP #2 would change if source treatment was applied.

Table 4.1: Wastewater Flows and Characteristics

Source	Annual Flow (10^6 US gal)	TDS (mg/L)	SO$_4$ (mg/L)	Zinc (mg/L)
WTP#2 Treating Tailings Pond Reclaim without source treatment (Option 1)	1,500	5,120	3,530	750
Mine Water Sump Characteristics (Option 2)	410	8,230	5,240	1,680
Waste Stockpile Drainage Characteristics (Option 3)	210	15,500	10,000	3,350
Combined Mine Water + Waste Stockpile Drainage Characteristics (Option 4)	620	9,700	5,900	2,600

4.3   Treatment Costs

4.3.1 Capital Costs

The capital costs for a new treatment system will depend in large part on the size of the facility. WTP #2 is a key component in all options and has a replacement value of about $12 million. The estimated costs for the new HDS plants needed under Options 2, 3 and 4 are as shown in Table 4.2.

Table 4.2: Capital Costs for HDS Treatment Options

	Average Design Flow Billion gal/year	Construction/Replacement Cost Million $
Option 1	1.5	12
Option 2	0.41	5.5
Option 3	0.21	4.0
Option 4	0.62	7.1

4.3.2 Operating Costs

4.3.2.1 Option 1-HDS Treatment at WTP#2

The cost for treating effluent at Red Dog was summarized in detail in the "Suspension Study" prepared by Teck Cominco Alaska Incorporated in May 2003.   A summary of the costs for the operation of WTP#2 is provided in Table 4.3. Sludge disposal costs are included in the pumping costs and assume all sludge is pumped to the tailings ponds.

As shown on Table 4.3, the annual cost to treat the effluent at WTP #2 is approximately $2.6 million. At closure many of the operating costs will be the same. If we assume that the acidity loadings are not substantially changed, then the expected cost would be as shown in Table 4.4. Key assumptions are:

1) There will be an increase in acidity treated at WTP#2 as WTP#1 has been decommissioned. This increases the lime demand for WTP#2.

2) An additional cost is included for capital equipment replacement allowance as the plant will be required for the long term.

3) The reduction in acidity loading as a result of not processing ore with acidity is assumed to be offset by increased acidity loading from the mine and waste stockpile.

The effect of these assumptions is to increase the annual operating costs for WTP #2 by $1.4 million to $3.9 million/year.

Table 4.3: Option 1 Operating Cost WTP#2 Only (Current Conditions)

| Cost Item	Unit	Usage	Unit Rate	WTP#2
Operating Labor				$360,000
Maintenance				$394,000
Power (Note 1)	kWh	7,740,000	$0.104[1]	$805,000
Consumables				
Lime	tonnes	3906	$174	$679,644
Flocculant	tonnes	39.1	$3,008	$117,613
Sodium Sulfide	tonnes	279	$722	$201,438
Total				$2,557,695
Water Treated	gal	1.95 billion		
Effluent discharged	gal	1.5 billion		
Cost/1000 gal treated	$/1000 gal			$1.31
Cost/1000 gal discharged	$/1000 gal			$1.71

Note 1: Reflects average projected cost of power

Table 4.4: Option 1 Operating Cost WTP#2 Only (Closure Conditions)

Cost Item	Unit	Usage	Unit Rate	WTP#2
Operating Labor				$360,000
Maintenance				$394,000
Power	kWh	7,740,000	$0.104	$805,000
Consumables				
Lime	tonnes	10,695	$174	$1,860,930
Flocculant	tonnes	39.1	$3,008	$117,613
Sodium Sulfide	tonnes	279	$722	$201,438
Capital Replacement Allowance	1.5% capital			$180,000
Total				$3,918,981
Water Treated	gal	1.75 billion		
Effluent discharged	gal	1.5 billion		
Cost/1000 gal treated	$/1000 gal			$2.24
Cost/1000 gal discharged	$/1000 gal			$2.61

The operating cost could further increase or decrease in future depending upon the acidity loading from the site. The sensitivity of the treatment costs to changing acidity loadings is shown in Figure 4.1. If the acidity at the site was reduced by 75% as a result of source controls (e.g. covers), the annual treatment cost would reduce by $1.4 million. Conversely, if the acidity level increases by 50% with the expanded mine and waste dumps, the annual treatment cost would increase by about $930,000.

Figure 4.1: Annual Operating Costs versus Acidity
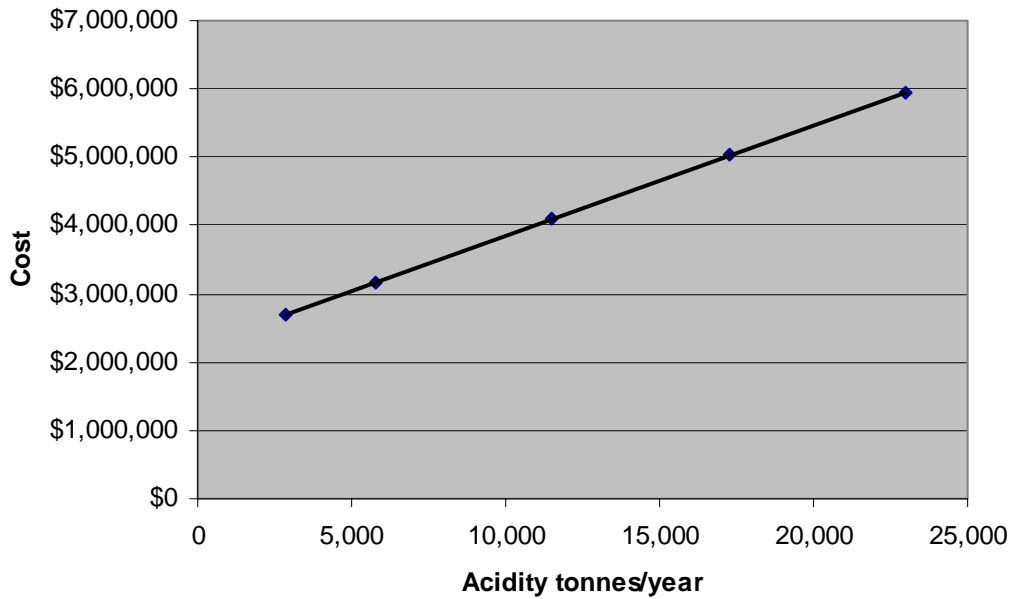
4.3.2.2 Option 2-Source Treatment-HDS Plant Treating Mine Water Only

For Option 2, a new HDS plant would be built to treat 0.41 billion galllons per year of mine water. It is assumed that the mine water would be seasonally stored in one of the open pits and treated over the same period as WTP#2.

This would not require winter operation of the plant and would allow for sharing of staff between the WTP#2 and the new facility. It was assumed that there would be an increase in the operating labor of about 10 % to account for increased labor demands during the annual startup and shutdown of the plants. Under normal operating conditions, the WTP#2 operator could handle all routine operations. Sludge from the plant would likely be stored in the pit. The estimated operating cost for the combined treatment facilities is shown on Table 4.5. The operation of a second treatment plant treating mine water would add about $375,000 to the annual treatment costs for the site.

Table 4.5: Option 2 Operating Costs (WTP#2+HDS Plant Treating Mine Water)

Cost Item	Unit	Usage	Unit Rate	Total Option 2
Operating Labor				$396,000
Maintenance				$472,800
Power	kWh	9,553,846	$0.104	$993,600
Consumables				
Lime	Tonnes	10,695	$174	$1,860,930
Flocculant	Tonnes	48	$3,008	$145,168
Sodium Sulfide	Tonnes	236	$722	$170,392
Capital Replacement Allowance	1.5% capital			$256,500
Total				$4,295,390
Water Treated at New HDS Plant	billion gal	0.41		
Water Treated at WTP#2	billion gal	1.75		
Effluent discharged	billion gal	1.50		
Cost/1000 gal treated	$/1000 gal			$2.45
Cost/1000 gal discharged	$/1000 gal			$2.86

4.3.2.3 Option 3-Source Treatment-HDS Plant Treating Waste Stockpile Drainage Only

For Option 3, a new HDS plant would be built to treat 0.21 billion gallons of waste stockpile drainage per year. It is assumed that the waste stockpile water would be seasonally stored in a new collection basin (not included in the costs) and treated over the same period as WTP#2. This would not require winter operation of the plant and would allow for sharing of staff between the WTP#2 and the new facility. Sludge from the plant would likely be stored in the pit. The estimated operating cost for the combined treatment facilities is shown on Table 4.6. The operation of a second treatment plant treating waste stockpile drainage would add about $235,000 to the annual treatment costs for the site.

Table 4.6: Option 3 Operating Costs (WTP#2 + HDS Plant Treating Waste Stockpile Drainage)

Cost Item	Unit	Usage	Unit Rate	Total Option 3
Operating Labor				$396,000
Maintenance				$453,100
Power (Note 1)	kWh	8,669,231	$0.104	$901,600
Consumables				
Lime	Tonnes	10,695	$174	$1,860,930
Flocculant	Tonnes	44	$3,008	$131,726
Sodium Sulfide	Tonnes	236	$722	$170,392
Capital Replacement Allowance	1.5% capital			$240,000
Total				$4,153,748
Water Treated at New HDS Plant	billion gal	0.21		
Water Treated at WTP#2	billion gal	1.75		
Effluent discharged	billion gal	1.50		
Cost/1000 gal treated	$/1000 gal			$2.37
Cost/1000 gal discharged	$/1000 gal			$2.77

4.3.2.4 Option 4-Source Treatment-HDS Plant Treating Mine water and Waste Stockpile Drainage

For Option 4, a new HDS plant would be built to treat 0.62 billion gallons per year of waste stockpile drainage and mine water. It is assumed that the mine water and waste stockpile drainage would be seasonally stored in one of the open pits and treated over the same period as WTP#2. This would not require winter operation of the plant and would allow for sharing of staff between the WTP#2 and the new facility. Sludge from the plant would likely be stored in the pit. The estimated operating cost for the combined treatment facilities is shown on Table 4.7. The operation of a second treatment plant treating mine water would add about $537,000 to the annual treatment costs for the site.

Table 4.7: Option 4-Operating Costs-WTP#2 + HDS treating Mine Water +Waste Stockpile Drainage

Cost Item	Unit	Usage	Unit Rate	Total Option 4
Operating Labor				$396,000
Maintenance				$492,500
Power	kWh	10,482,692	$0.104	$1,090,200
Consumables				
Lime	tonnes	10,695	$174	$1,860,930
Flocculant	tonnes	53	$3,008	$159,281
Sodium Sulfide	tonnes	236	$722	$170,392
Capital Replacement Allowance	1.5% capital			$286,500
Total				$4,455,803
Water Treated at New HDS Plant	billion gal	0.62		
Water Treated at WTP#2	billion gal	1.75		
Effluent discharged	billion gal	1.50		
Cost/1000 gal treated	$/1000 gal			$2.55
Cost/1000 gal discharged	$/1000 gal			$2.97

4.4 Total Treatment Costs

4.4.1 Total Treatment Costs

Table 4.8 provides a summary of the total costs for treatment for the four options.   The following are key observations:

- The long term operation of WTP#2 to manage effluents from the Red Dog site has a NPV of about $98 million.

- The treatment of the waste stockpile drainage and mine sump water separately in a new plant with final polishing at WTP#2 would increase the long term treatment costs by $20 million. The overall cost would be less if WTP#1 or the new WTP#3 could be used, thereby reducing the capital cost.

Table 4.8: Total Treatment Costs

Option	1	2	3	4
Description	WTP #2 Only	WTP #2 Plus Treatment of Mine water	WTP #2 Plus Treatment of Waste Stockpile Drainage	WTP #2 Plus Treatment of Mine Water and Waste Stockpile Drainage
Operating Cost	$3,918,981	$4,295,390	$4,153,748	$4,455,803
NPV 4%	$98,000,000	$107,000,000	$104,000,000	$111,000,000
Cost $/1000 gal	$2.61	$2.86	$2.77	$2.97
Capital Cost	existing plant	$5,100,000[1]	$4,000,000[1]	$7,100,000[1]
Total Cost	$98,000,000	$112,100,000	$108,000,000	$118,100,000
Differential Cost	0	$14,100,000	$10,000,000	$20,100,000

Note 1: The capital may be reduced if WTP#1 or the proposed WTP#3 was used in lieu of a new plant.

4.4.2 NPV Cost versus Acidity Loading

The estimated net present value (NPV) of the long term treatment costs for Option 1 is $98 million. Future changes to acid production from the site can have significant effects on the long term cost. The estimates in Table 4.8 assume a total annual lime requirement similar to the current 10,500 tonnes per year. If acidity levels increase or decrease, the primary effect on treatment cost will be an increase or decrease in lime use. Figure 4.2 illustrates the effect of acidity production at the site on the NPV treatment cost for Option 1. NPV costs for the other options are similarly sensitive to changes in acid production.
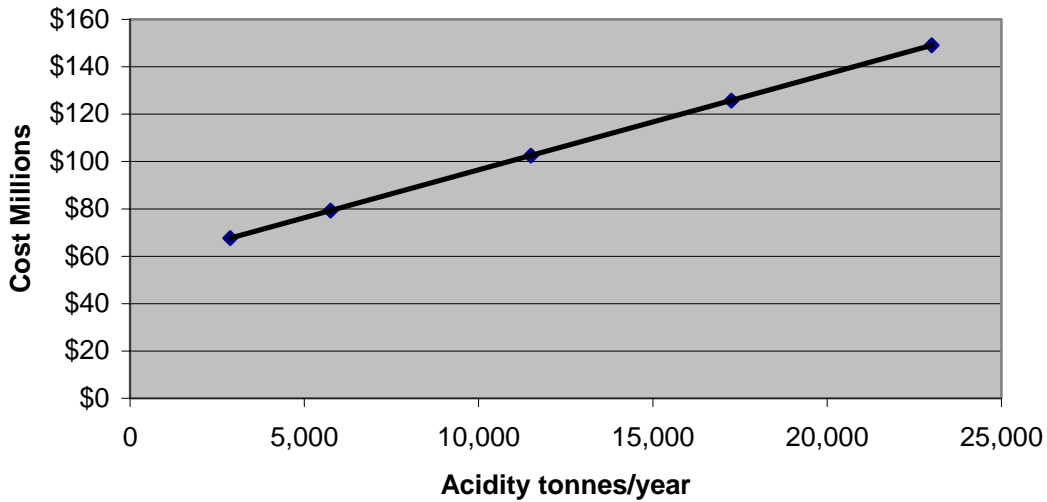

Figure 4.2: NPV of Total Treatment Cost versus Acidity Loading

If source controls are applied (e.g. covers over the waste stockpile) and a 50% reduction in the acidity loading is achieved, the NPV of the wastewater treatment cost would reduce by about $23,000,000. A 75% reduction would decrease the NPV by approximately $34,000,000. Conversely, if acidity levels in the mine sump and waste stockpile drainage were to rise and double the acidity load, the NPV for wastewater treatment would increase by approximately $45,000,000. These results indicate the range of uncertainty in the current estimates and the potential value of source control methods. Source control options are being addressed in separate reports.

4.5   Sludge Production

The estimated sludge production for Option 1, i.e. only WTP#2 operation, at closure are shown in Table 4.9. The following key observations are noted:

- At closure with WTP#2 in operation and sludge densities of 20%, the site will produce about 18,000 t/year of sludge or a sludge volume of about 105,000 yd^3.

- If source reduction methods are applied and the acidity levels decrease by 75%, future sludge production would be reduced to about 2,000 tonnes per year. Sludge densities would be lower, in the order of 2-5 % solids, because gypsum would not be precipitated. The lower densities mean that sludge volumes would remain high.

Table 4.9: Estimates of Sludge Production

	Primary Sludge Components (tonnes/year)				Volume yd^3
	Gypsum	Zn(OH)$_2$	Other Metals	Total	
Option 1-HDS WTP#2	10,250	6,375	1,675	18,200	104,000[1]
Source Reductions-75% reduction of acidity levels	0	1600	400	2,000	50,000-130,000[1]

Note 1- Assume density would be 20% for Option 1 and reduced to 2 -5% solids if source controls reduce acidity by 75%.

4.6   Projected Effluent Quality

It is expected that all treatment systems will provide effluent quality similar to the existing WTP#2. The one change will be in effluent TDS and sulfate levels. For the four options considered herein, estimates of the final effluent sulfate and TDS levels are shown below in Table 4.10. Only Option 4, treating both mine water and waste stockpile drainage at source, is predicted to result in a material reduction of TDS and sulfate levels in the final effluent.

Table 4.10: Projected Effluent TDS and SO4 Levels

	TDS mg/L	SO4 mg/L
Option 1- WTP#2	3,400	2,400
Option 2-WTP#2+New HDS Plant for Mine Water	3,400	2,400
Option 3-WTP #2+new HDS Plant for Waste Stockpile Drainage	3,400	2,400
Option 4-WTP#2+ new HDS Plant for Mine Water + Waste Stockpile Drainage	2,100	1,700

5   Key Findings

1) There are no reasonable alternatives to the current HDS system for long-term wastewater treatment at Red Dog. The alternatives reviewed herein are more complex than HDS, less cost effective in achieving TDS control, and/or unproven on Red Dog wastewaters

2) The preferred option for supplemental TDS control, if it is required in the long term would be HDS treatment of the mine sump water and waste stockpile drainage prior to mixing with the tailings pond. Treatment of these streams has the potential to reduce TDS levels in the final discharge to about 30% below current levels.

3) Separate treatment of the mine water and waste stockpile drainage could also reduce TDS levels during the remaining mine life. TCAK is currently constructing WTP#3. The new plant should be on-line in the spring of 2005.

4) Reduction of the acidity loadings at source (e.g. covers for the waste stockpile) could be a more cost effective method of TDS control than any treatment system. It is estimated that a 75 % reduction of acidity from current levels would result in a total treatment cost savings of about $34 million NPV. TDS levels would be reduced to about 1800 mg/L without implementation of any additional treatment. In addition, sludge production would be reduced from about 18,000 t/year to about 2000 t/year. Options for source control are being investigated in other closure plan tasks.

5) Long term sludge production from the site will become an important consideration to be addressed in the closure plan. Assuming high density sludge is produced at 20% solids, at the current acidity loadings, the site would produce about 104,000 yd^3/year of sludge. Options for long-term sludge management are discussed in a separate report.

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