

***Reverse Osmosis Pilot Test Report***

***SD026 Active Treatment Evaluation***

***Prepared for  
Cliffs Erie LLC and PolyMet Mining Inc.***

***June 2013***



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## SD026 Active Treatment Evaluation

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# 1.0 Introduction

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The water in the southeast corner of the former LTV Steel Mining Company (LTVSMC) tailings basin (Cell 1E) is the likely driving force for the spring that forms the headwaters of Second (aka Knox) Creek. From the headwaters, Second Creek flows to the southwest into the Partridge River, and eventually into the St. Louis River and into Lake Superior.

The discharge at the headwaters of Second Creek is administered under Minnesota Pollution Control Agency (MPCA) National Pollutant Discharge Elimination System (NPDES) / State Disposal System (SDS) Permit MN0042536 (Permit). In the Permit, this discharge is designated SD026 and is routinely monitored at the culvert where Second Creek passes beneath Dunka Road, approximately 1,500 feet downstream of the source spring. The general site layout is shown on Figure 1. The Permit is currently held by Cliffs Erie LLC (CE), but PolyMet Mining Inc. (PolyMet) is collaborating with CE on the reissuance of the Permit.

A key aspect of the Permit renewal process is the development of a plan to address sulfate and parameters of concern that have had elevated concentrations in the discharge. The ‘parameters of concern’ (as defined in the April 6, 2010 Consent Decree between MPCA and CE) are total dissolved solids (TDS), bicarbonates, total hardness, and specific conductivity. Collection and pump-back of the source area seep from the tailings basin to SD026 was identified as the preferred short term mitigation in the *Short-Term Mitigation Evaluation and Implementation Plan for SD026*, which was submitted to the MPCA in June 2010. The pump-back system was installed and put into operation in 2011 and is still operating.

A *Long-Term Mitigation Evaluation and Implementation Plan for SD026 (Long-Term Plan)* was submitted to the MPCA in April 2012. The *Long-Term Plan* described the testing of passive treatment technologies (natural attenuation, enhanced natural attenuation, and lagoon/surface-flow wetland) for reducing the concentrations of the parameters of concern in the discharge. As part of their review of the *Long-Term Mitigation Evaluation and Implementation Plan for SD026*, the MPCA indicated that testing of active treatment technologies would also be required.

In September 2012, a work plan (*Work Plan for Investigation of Membrane Treatment at SD026*) was submitted to the MPCA to investigate the feasibility and cost of using reverse osmosis (RO) to reduce the concentrations of sulfate and parameters of concern to at or below the applicable water quality standards for Second Creek. The primary objectives of the testing proposed in the work plan

were to collect sufficient information to (1) determine the requirements for successful implementation of reverse osmosis to meet the water quality goals, including permeate stabilization and concentrate management, and (2) collect sufficient information to refine the capital and operating costs for treatment systems necessary to meet the water quality goals. The work plan was accepted by the MPCA in January 2013.

This report presents the data collected during the testing program, evaluates its applicability to the discharge at SD026, evaluates concentrate management approaches, and provides preliminary estimates of costs to implement the evaluated technologies. The report is structured as follows:

1. A presentation of the project background (Section 1.0)
2. An overview of the testing program structure (Section 2.0)
3. A review of the historical, tested, and target water qualities (Section 3.0)
4. A presentation and evaluation of the reverse osmosis pilot testing results (Section 4.0)
5. A presentation and evaluation of the concentrate management testing results (Section 5.0)
6. A presentation and evaluation of the permeate stabilization testing results (Section 6.0)
7. A presentation of the preliminary cost estimates for the evaluated treatment approaches (Section 7.0)
8. A discussion of final conclusions from the testing program (Section 8.0).



## 2.0 Testing Overview

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In order to meet the pilot testing objectives, the pilot testing program was conducted in phases, to provide periods of time for investigation and optimization and time for collection of data to assess the longer term performance of the processes under investigation. Each of the testing phases and its objectives are briefly described in the following sections.

### 2.1 Phase 1 – Startup and Commissioning

Phase 1 consisted of the startup and commissioning of the reverse osmosis and greensand filter pilot units. This period provided an opportunity for pilot unit installation and assembly, tuning of control systems, implementation of the data collection procedures, and initiation of operation and the initiation of the process of determining operating conditions. Operator training by the vendor was provided during this phase.

### 2.2 Phase 2 – Membrane Selection, Pretreatment Investigations, and System Optimization

The purpose of Phase 2 was to identify pretreatment requirements and RO operating conditions that optimize the treatment train (balancing capital costs, operating costs, and reliability). During this phase, greensand filter operation as well as the recovery and flux of the RO system were adjusted and monitored to determine an operating approach for use in Phase 3.

### 2.3 Phase 3 – Steady-State Operation

During Phase 3, the treatment train and operating conditions optimized during the Phase 2 investigations were used. The treatment system was operated, largely unaltered, for the duration of Phase 3 under steady-state conditions. The purposes of this test were to gain longer-term operating data on the proposed system to evaluate system reliability, system performance with respect to water quality targets, life cycle cost, ability to effectively clean the membranes, and to generate permeate and concentrate for use in the effluent stabilization and concentrate management investigations.

### 2.4 Effluent Stabilization Investigation

Were RO to be implemented for the treatment of the discharge from SD026, the future treatment plant effluent will be a blend of very low-TDS water (e.g., permeates and distillates). The effluent blend will be void of alkalinity and hardness, making the water corrosive to piping and materials near the outfall. The objective of the effluent stabilization investigation was to identify a stabilization

method (e.g., addition of minerals) that will reduce the corrosiveness of the blended effluent, while maintaining compliance with the effluent water quality targets (Section 3.0).

## 2.5 Concentrate Management Investigation

Mechanical evaporation and crystallization of the RO concentrate was the concentrate management method determined to be potentially feasible at SD026 in the *Short-Term Mitigation Evaluation and Implementation Plan for SD026*. The primary objectives of this concentrate management investigation were to verify the technical feasibility of this approach to concentrate management, and to better quantify the associated costs. For this investigation, three “pretreatment” approaches for the evaporator and crystallizer were evaluated:

1. Greensand filtration followed by reverse osmosis (with the RO concentrate being sent to the evaporator without further treatment), and
2. Greensand filtration followed by reverse osmosis, with the RO concentrate being further reduced in volume by a specialty RO membrane (vibratory sheer enhanced processing (VSEP)).
3. Greensand filtration followed by reverse osmosis, with the RO concentrate being further reduced in volume by intermediate concentrate chemical precipitation (ICCP) and RO.

Once steady-state operation of the RO pilot was established, a study of further reduction of the concentrate volume reduction was initiated via routing the RO concentrate through the VSEP system, manufactured by New Logic Research. The objective of this investigation was to evaluate the recovery, fluxes, and operational requirements for the VSEP equipment, and to characterize the resulting concentrate and permeate quality.

To investigate the ICCP process, bench testing of lime soda-ash softening of the RO concentrate was conducted, followed by an evaluation of the secondary RO system by GE, which included modeling of the secondary RO system performance, preliminary equipment sizing, and assessing pretreatment requirements.

For all of the three pretreatment options listed above, data was provided to GE for their use in evaluating the use of evaporation and crystallization in this application.

## 3.0 Water Quality

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### 3.1 Historical Water Quality

The quality of the water being discharged via the outfall at SD026 has been monitored regularly over the past several years. Table 1 includes a statistical summary of the water quality data from SD026 from January 2005 through December 2011. These data are based on the analytical results from the monthly Discharge Monitoring Reports (DMRs) as reported by CE in accordance with the Permit. For parameters that had concentrations below their corresponding reporting limit values, concentrations were reported as the reporting limit values. These reporting limit values, along with the actual analytical data results, were used to calculate the average concentrations.

In 2012, additional water samples were collected from SD026 and analyzed for parameters of importance to evaluate the RO treatment systems. These data are presented in Table 2.

### 3.2 Comparison to Tested Water

Following the submission of the work plan, bench testing and analytical work were conducted to compare the water at SD026 to the water proposed to be used as feed water for the RO pilot test. The proposed feed water was a blend of water from another seep at the Tailings Basin, SD004, and a newly installed well in the northwest corner of the Tailings Basin (locations shown on Figure 1). The comparison of the water qualities is presented in Appendix B. It was determined that SD026 had similar water quality to that of SD004 and the new well, particularly after treatment with greensand filtration. Where differences in the water quality were such that different treatment considerations exist, those are identified and described within the evaluations presented in Sections 4.0 through 6.0.

### 3.3 Treated Water Quality Goals

The allowable concentrations of parameters in permitted discharges are derived from the beneficial use classification(s) of the receiving water. The receiving water for SD026 is Second (aka Knox) Creek, which is an unlisted water with default beneficial use classifications of 2B, 3C, 4A, 4B, 5, and 6, as described in Minnesota Rule Chapter 7050.0430. However, because a detailed evaluation of actual uses of the receiving water has not yet been completed, the potentially applicable Minnesota water quality standards were used as a guide in the selection of reverse osmosis for testing and for development of this work plan.

### 3.3.1 Sulfate

The primary use classifications pertaining to sulfate are the Class 1, Class 4A, and Class 4B water quality standards, as described in Minnesota Rules Chapters 7050.0221 and 7050.0224. Waters with a beneficial use classification of 1 are used for drinking water consumption and, as noted above, are not applicable to Second Creek. Class 1 waters must meet the U.S. EPA primary and secondary drinking water standards. The secondary drinking water standard for sulfate is 250 milligrams per liter (mg/L). For waters with a beneficial use classification of 4A (irrigation), the sulfate concentration in those waters can be no greater than 10 mg/L for waters where wild rice is produced “during periods when the rice may be susceptible to damage by high sulfate levels.” For waters with a beneficial use classification of 4B (livestock and wildlife consumption), while no numeric standard is given, it has been interpreted by the MPCA to mean that the sulfate concentration in those waters cannot be greater than 1,000 mg/L. Given the potentially wide range of applicable water quality targets for sulfate discharge to Second Creek, a conservative value of 10 mg/L was used as a treatment target when evaluating the data collected during the pilot testing program.

### 3.3.2 Other Parameters of Concern

For the parameters of concern, the following in-stream water quality standards, from Minnesota Rules Chapter 7050.0222 through 7050.02, were used as a guide for setting treatment goals for the pilot test:

- Class 2B (fishing and aquatic life): no specific requirements for the parameters of concern
- Class 3C (industrial cooling and materials transport): Hardness – 500 mg/L
- Class 4A (irrigation): Bicarbonates – 5 millequivalents per liter (meq/L) (250 mg/L as CaCO<sub>3</sub>), specific conductance – 1,000 micromhos per centimeter (µmhos/cm), total dissolved salts – 700 mg/L
- Class 4B (livestock and wildlife consumption): Total salinity – 1,000 mg/L
- Class 5 (aesthetic enjoyment and navigation): no specific requirements for the parameters of concern
- Class 6 (other uses): no specific requirements for the parameters of concern

## 4.0 Reverse Osmosis Pilot Test Results

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### 4.1 Pretreatment

#### 4.1.1 Greensand Filtration

The greensand filter pilot unit provided by GE for the pilot test was a pressure filter (Figure 2). This filter is a 30-inch diameter unit filled with coarse gravel (5 inches), greensand filter media (30 inches), and anthracite (12 inches). The greensand media is silica sand coated with manganese oxide. Technical information on the greensand used during the pilot test and information on the GE pilot unit systems can be found in Appendix A.

For the pilot test, the influent was dosed continuously with potassium permanganate in order to (1) oxidize iron and manganese for removal by filtration and (2) regenerate the greensand media.

##### 4.1.1.1 Filter Loading

Over the duration of the testing program, the influent flow rate ranged from 19 to 22 gpm. The resultant range of hydraulic loading to the filter was 3.5 to 4.9 gpm per square foot ( $\text{gpm}/\text{ft}^2$ ) of filter bed area.

##### 4.1.1.2 Filter Removal Rates

The greensand filter removal rates for total suspended solids (TSS), iron, and manganese are presented in Table 3. Overall (including startup and optimization phases of testing), the removal of TSS across the filter averaged  $>87\%$  (to less than the method reporting limit in the filtrate). During Phase 3, the removal of TSS was  $>90\%$  on average. Iron removal by the filter consistently averaged  $>99.7\%$ . Table 5 displays the greensand filtrate water quality.

During Phases 2 and early in Phase 3, it was noted that, at times, manganese was breaking through the filter (Table 3). Breakthrough was particularly noticeable at start-up following system idle periods, such as on Mondays or at the beginning of the day. Because of this, a trial to improve manganese removal was initiated during Phase 3 (at the end of August 2012). For this optimization, the permanganate dose was increased every other day, with daily monitoring of filter influent and effluent manganese. The greensand filter influent and effluent samples were analyzed by Pace Analytical during this time in order to obtain the data more quickly to inform daily permanganate dose adjustments. (Weekly routine sample analysis by Legend was also continued during this optimization period.) In order to protect the RO membranes from potential damage from excess permanganate (a strong oxidant), sodium bisulfite was dosed immediately ahead of the RO unit.

Figure 3 provides an overview of the manganese removal results obtained during this optimization. A final potassium permanganate dose of about 4.5 mg/L was selected as the optimal dose for manganese removal based on the filtrate dissolved manganese concentration. As can be seen in Figure 3, manganese removal was significantly improved from an average of 81% prior to optimization to an average of 97% after optimization. The results suggest that the breakthrough of manganese observed during Phase 2 and 3 was likely due to the incomplete oxidation of dissolved manganese and/or insufficient regeneration of the greensand media at the permanganate doses initially applied during testing.

#### **4.1.1.3 Residuals**

Periodically, accumulated solids must be removed from the filter bed to maintain hydraulic capacity and performance. A filter backwash can be triggered based on filter run time, or more commonly, an increase in pressure drop across the filter. For the pilot unit, pressure drop was used to trigger backwash events. When the pressure drop across the unit reached approximately 10 psi, feed water was pumped up through the filter bed at a rate of 60 to 70 gpm (12 gpm/ft<sup>2</sup>) to remove solids from the bed. During Phase 3 operations, the filter backwash frequency was approximately once every two days. Samples of the spent backwash water were collected and analyzed periodically. Greensand filter backwash water quality results are summarized in Table 4. In addition to containing elevated concentrations of TSS, iron, and manganese—the targeted constituents—the spent backwash water also contained elevated concentrations of organic material (as chemical oxygen demand), silica, and a number of other trace metals.

#### **4.1.1.4 Discussion**

The primary purpose of the greensand filter was to protect the RO membranes by removing particulate matter, iron, and manganese. The filter removed TSS and iron to concentrations below the method reporting limits. Manganese was also significantly reduced, especially after optimization of the potassium permanganate dose during Phase 3. The RO membranes did not exhibit signs of fouling during the 7 month pilot test (as discussed in more detail in Section 4.2). The greensand filter was a simple-to-operate, effective means of pretreatment for the feed water from SD004 and the pilot test well.

Greensand filtration is expected to perform similarly for the discharge from SD026. Over the duration of the pilot test, for the measured dissolved iron and manganese concentrations at the feed tank, the theoretical potassium permanganate dose required ranged from approximately 1.8 to 4.5 mg/L. A limited number of data points exist for dissolved iron and manganese at SD026, but the

available historical water quality data for SD026 suggest that similar or slightly greater potassium permanganate doses may be required, due to greater concentrations of dissolved manganese that have been observed at times in the SD026 discharge.

In full-scale application, one of the primary design criteria for greensand filters is the hydraulic loading rate. The loading rate for greensand filters has the potential to affect the manganese removal efficiency, the backwash frequency, and the number of filters required for filtration. For this pilot test, the hydraulic loading rate was fixed by the pilot unit supplied by GE, and was higher than typical hydraulic loadings for this type of filter (approximately 4.5 compared to 3 gpm/ft<sup>2</sup>), particularly given the concentrations of iron and manganese in the influent. However, higher-than-typical loading rates can be acceptable if demonstration testing shows acceptable treatment performance and backwash frequency, which was case during this pilot testing program.

#### **4.1.2 Chemical Pretreatment**

The greensand filter effluent was treated with 1 ppm sodium bisulfite (to quench any excess permanganate from the filter and prevent membrane oxidation) and 2.2 ppm of Hypersperse MDC150, a scale inhibitor.

As can be seen in Figure 4, at times (including 2012 grab samples), the water at SD026 contains higher concentrations of bicarbonate alkalinity than was observed in the feed water during the pilot testing program. For implementation of RO for the treatment of water from SD026, additional pretreatment of the water with a mineral acid is recommended to mitigate scaling from calcium carbonate.

## **4.2 Reverse Osmosis**

The RO pilot unit was provided by GE. A picture of the pilot test unit employed for the project is shown on Figure 5. Manufacturer's information on the pilot unit can be found in Appendix A. The RO pilot unit provided by GE used 18 4-inch-diameter RO modules housed in six vessels, in a 2-2-1-1 array. The membranes employed were low-pressure RO membranes (GE model AK90-LE).

The pilot unit was operated continuously for approximately 8 hours per day, typically 5 days per week. At the end of each 8-hour shift, the RO system was flushed with permeate and shut down.

### **4.2.1 Flux and Recovery**

During Phase 2 of the pilot test, a number of operating conditions were tested to optimize the RO system operation. The primary operating variables adjusted were recovery (the percentage of feed

water volume that becomes permeate) and flux (the flow rate through the system per unit of membrane in service). In general, the higher the membrane flux, the lower the membrane area required for a given treatment capacity. However, operation at higher flux rates has the potential to increase the fouling rate of the membranes.

Phase 2 lasted approximately 8 weeks and the conditions tested were as follows:

- Condition 1 – 75% recovery, flux of 14 gfd – 3 weeks
- Condition 2 – 80% recovery, flux of 16 gfd – 3 weeks
- Condition 3 – 80% recovery, flux of 18 gfd – 2 weeks

The RO pilot unit performed well at all conditions tested. Condition 3 was considered a “stress condition” because the flux was at the upper end of what is generally used in the design of RO groundwater treatment systems (US EPA, 2005). Nevertheless, for the short duration test of this operating condition, no operational problems were encountered. The feed-to-concentrate pressure drop across the RO system was stable at all three conditions and was well below the threshold to initiate membrane cleaning (> 50 psi per stage). Changes in recovery and flux can also impact the salt rejection of the membranes. Over the conditions tested in Phase 2, no unacceptable or significant changes in permeate water quality were observed. For Phase 3, a flux of 16 gfd and recovery of 80% were selected. This combination of operating conditions was determined to provide an acceptable performance and reliability. The small increase in pressure drop at the 18-gfd flux condition further demonstrated the selected flux (16 gfd) is not an operational maximum.

During Phase 3, the RO membrane system operated continuously at a recovery of 80% and a flux of 16 gfd. The feed-to-concentrate pressure drop throughout Phase 4 was approximately 25 to 30 psi with little upward movement. The feed-to-concentrate pressure drop and the feed pressures experienced over the course of pilot testing are shown on Figures 6 and 7. The absence of any substantial change in feed pressure or feed-to-concentrate pressure drop suggests that very little scaling or fouling of the membranes occurred during the pilot testing program.

#### **4.2.2 Permeate Water Quality**

The RO feed (greensand filter effluent), permeate, and concentrate water quality data collected during Phases 2 and 3 are summarized in Table 5, Table 6, and Table 7, respectively.



#### 4.2.2.1 Removal Rates

Average removal rates were estimated for those parameters with detectable concentrations in the greensand filter effluent (RO feed) and are displayed in Table 8. The average sulfate removal was 99.8% during the pilot test (Figure 8). The average sulfate concentration in the RO permeate was 0.57 mg/L, and the highest sulfate concentration observed was 0.98 mg/L, well below the 10 mg/L water quality target. During Phase 3, the average salt passage through the membranes, as represented by the percentage of TDS in the feed that passed through the membrane to the permeate, was <0.6% with no reported TDS in the permeate, at a reporting limit of 10 mg/l (Figure 9).

Many other parameters, particularly the major anions and cations, were reduced by greater than 95%. However, in many instances the upper limit of removals were not determined in the routine testing because (1) the concentrations measured in the permeate were less than the method reporting limit and/or (2) the concentrations in the influent were low and close to the method reporting limit.

For some constituents, removal by RO membranes is highly pH-dependent. Examples of this are ammonia and boron. For these compounds, over a range of pH values, they are present as un-ionized species. The un-ionized species are not well-removed by membranes. For this pilot test, the following observations were noted:

- Ammonia: At pH values below 7, most of the ammonia is present as the ammonium ion and can be removed by the RO process. However, the pH of the feed water to the pilot RO system is approximately 7.5, reducing the amount of ammonia that can be removed. In addition, the concentration of ammonia in the influent was relatively low. The low concentration in the influent limited the ability to accurately estimate the ammonia removal by the RO system.
- Boron: It is well known that boron removal at pH values below the pKa (pH = 9.2) of boric acid is limited due to the lack of charge on the species. The boron removal during the pilot-testing program, while limited, was sufficient to maintain permeate concentrations below 0.5 mg/L, the Class 4A water quality standard. Specialty membranes or pH adjustment are typically required for greater boron removal.

#### 4.2.2.2 Comparison to Equipment Supplier Models

The suppliers of RO membranes commonly use models in their system design and to estimate the permeate water quality. Each supplier typically has developed their own models for their membranes, and each supplier has significant operating data collected over many years for validation

of the model output. The model water quality input and output is generally limited to the major anions and cations, pH, boron, and certain constituents of concern with respect to membrane fouling or scaling (e.g., aluminum, barium, silica, strontium). Because the water tested during the pilot program was other than the SD026 discharge, modeling of treatment by RO of the SD026 discharge was conducted. The average water quality from the greensand filter bench test conducted using SD026 water (Appendix B) was provided to GE for input into the model used for preliminary pilot test planning. GE's RO model results are presented in Table 9. As can be seen, the modeled treatment of SD026 water by RO resulted in:

- 80% recovery across the RO system
- Pretreatment with antiscalant and acid
- Permeate water quality very similar to that demonstrated during the pilot test for sulfate and parameters of concern

#### **4.2.3 Cleaning Requirements**

Inorganic and organic scale and foulants build up on RO membranes over time and reduce performance. Membranes are chemically cleaned-in-place (CIP) to remove the foulants and restore performance. CIPs are triggered either when the system pressure drop reaches a predetermined value or increases by a certain percentage, if salt passage increases beyond a certain percentage, or on a regular time interval, if other parameters have not triggered a CIP. GE generally recommends that membranes be cleaned every 3-4 months (of continuous operation) if a CIP has not been initiated for other reasons.

Significant increases in pressure drop from the RO feed to the concentrate were not seen in any phase of the pilot testing. A CIP was conducted on July 30, 2012 to test the cleaning procedures recommended by GE. A low pH cleaner (citric acid) and a proprietary high pH cleaner from GE were used to clean the membranes during the CIP. The cleaning solutions were recirculated through the membranes in a two-step cleaning process and samples of the spent cleaning wastes were collected for analysis (Table 10).

The analytical results from the chemical cleaning wastes can provide insight into the fouling or scaling constituents on the membranes and which cleaner removes them. The following were elevated following treatment of each cleaner:

- low pH cleaner – chemical oxygen demand (COD, from the cleaner), TDS, aluminum, barium, calcium, iron, magnesium, manganese, and sodium

- high pH cleaner – Sodium and COD (both from cleaner) and magnesium

In the low pH cleaning solution waste, iron and manganese were the metals present in the highest concentrations. This finding was one of the reasons for conducting the greensand filter optimization study described in Section 4.1.1.2

#### **4.2.4 Discussion**

The selection of RO for pilot testing of mechanical treatment of SD026 water was driven primarily by its potential to produce treated water containing less than 10 mg/L of sulfate. Throughout Phases 2 through 4, the RO membranes produced a permeate water quality that consistently met that and other treated water quality goals described in Section 3.0. The average sulfate concentration observed in the RO permeate was 0.57 mg/L (with 0.98 mg/L being the highest concentration observed), which is an average sulfate removal efficiency of 99.8% across the membranes. It is expected that sulfate removal may change over time as the membranes age, but it is also expected that, even with some degradation of performance, water quality goals are likely to be met.

Throughout the duration of the pilot testing program, no significant operational or maintenance problems were encountered. Based on influent water chemistry and RO treatment modeling conducted by GE, the recovery selected for the RO pilot unit was primarily a function of the solubility limits of calcium carbonate and silica, which become saturated or supersaturated at the membrane surface during treatment. During the pilot test, a scale inhibitor (a phosphonic acid salt solution) was used to manage the formation of scale and silica on the membranes. The membrane system did not experience a significant increase in pressure drop from the RO feed to the concentrate. This stability indicates that scaling and fouling were not significant during the pilot test and that the pretreatment systems in place were effective.

The feed pressures observed during the pilot were stable and were lower than many brackish water RO applications, averaging 123 psi. The low feed pressures translate to lower operational (energy) costs for pumping into the system.

## 5.0 Concentrate Management

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Concentrate management is a key component that must be considered for successful implementation of membrane treatment for any application. Concentrate management options for the use of RO at SD026 were reviewed in *Short-Term Mitigation Evaluation and Implementation Plan for SD026* in 2010. That report concluded that evaporation and crystallization of the RO concentrate, followed by disposal of the solid salts, was the most likely technically feasible option. The objectives of this study were to evaluate these technologies in more detail, and to determine if other complementary volume-reducing technologies could reduce the cost of producing a solid salt waste product for disposal.

The evaluation consisted of bench testing, pilot testing, and input from GE, a manufacturer of evaporation and crystallization equipment. The sections below describe the results of technical evaluations completed. Updated estimates of implementation costs are provided in Section 7.0.

### 5.1 Evaporation and Crystallization

The use of evaporation (brine concentration) and crystallization immediately after the RO system, as shown in Figure 10, is the approach that was reviewed originally in the *Short-Term Mitigation Evaluation and Implementation Plan for SD026*. For the evaluation conducted in this work, more complete water quality data from SD026 was provided to GE for their use in assessing the use of RO to treat the water from SD026. Using this data, and comparing it to the pilot test data, GE modeled what the resulting RO concentrate water quality would be, using the same membrane that was employed during the pilot testing program. The modeling results are presented in Table 11. The resulting modeled RO concentrate was then used as the theoretical feed source for evaluation of the mechanical brine concentrator and crystallizer systems.

For this application, GE suggested the following equipment:

- Falling-film, mechanical vapor recompression evaporator with seed slurry, followed by a
- Mixed-salt, calandria crystallizer
- Further dewatering of the salt solids using a centrifuge

Some pretreatment of the RO concentrate prior to the evaporator is necessary. The recommended pretreatment consists of pH adjustment to reduce carbonate scaling, followed by deaeration to

remove the resulting carbon dioxide. The feed to the evaporator is heated, in part, by the waste heat from the system distillate (via a heat exchanger).

During the evaporator cycle, concentrated brine slurry is continuously withdrawn and sent to the crystallizer. In the crystallizer, the brine slurry becomes supersaturated and salt crystals will precipitate. The precipitates are separated from solution by a filter press or centrifuge.

The water from SD026, if similar to the pilot feed water, may contain some naturally occurring organic matter (as total organic carbon, TOC). The TOC may impact the performance of the evaporator and crystallizer units in two specific ways: foaming; and decreased solids separation efficiency in the dewatering process. To manage foaming, antifoaming agent would be continuously added to the crystallizer body. To manage the effects of the TOC on the solids separation process, a purge may be necessary from the crystallizer. This concentrated, liquid purge stream would require detailed evaluation of disposal options. Alternatively, additional pretreatment may be required upstream of the RO process to remove or mineralize the TOC prior to treatment. Additional treatment processes could include options such as adsorption with granular activated carbon, or advanced oxidation processes such as ozone/peroxide.

The distillates generated by the evaporator and crystallizer units will be low in TDS (< 50 mg/L). These distillates would be blended with the RO permeate for stabilization and discharge. The distillate may contain some volatile components such as ammonia or organic compounds that have been cycled up throughout the treatment train, however, their dilution by the primary RO system permeate will likely result in only minor increases compared to the feed water concentrations.

## **5.2 Concentrate Volume Reduction using VSEP**

The capital and operating costs of evaporators and crystallizers can be a significant percentage of overall project costs. The VSEP process was evaluated as a possible method to reduce or eliminate the evaporator and reduce overall operating costs of the treatment train. The VSEP configuration evaluated is illustrated in Figure 11.

### **5.2.1 VSEP Pilot Test Results**

The VSEP pilot unit was provided by New Logic Research. A picture of the pilot test unit that was used in the pilot testing program is shown on Figure 12. Manufacturer's information on the pilot unit can be found in Appendix A. The unit can be operated in batch mode or single-pass (continuous) mode, and both operating modes were tested during the pilot testing activities. For the pilot test, RO membranes (ESPA series by Hydranautics) were used.

### **5.2.1.1 Pretreatment and Optimization**

During the initial phase of testing for the VSEP unit, a number of methods for optimizing performance of the system were investigated:

- operational mode selection—batch versus single-pass operation—to maximize system recovery
- antiscalant dose selection to maximize system recovery
- acidification of the VSEP feed water to maximize system recovery
- cleaning chemical selection and cleaning procedure refinements to maximize the restoration of membrane flux

The preliminary investigations related to each of these are described in the sections that follow.

### **5.2.1.2 Operational Mode**

The initial startup and optimization of the VSEP unit was led by a New Logic Research field engineer with assistance provided by PolyMet staff. New Logic Research operated the unit in both batch and single-pass mode and determined that greater flux stability could be achieved by operating the unit in batch mode. In batch mode, the VSEP system uses a constant cross flow along with vibration to reduce fouling and polarization at the membrane surface. For the batch process, a fixed volume of concentrate from the GE RO system is fed to the VSEP system. The concentrate from the VSEP unit is returned to the VSEP feed tank and the VSEP permeate is discharged (as illustrated on Figure 12). As a result, the concentration of total dissolved solids in the feed tank increases over the duration of batch processing. This process continues until the target recovery has been achieved or until the flow through the membrane falls below a predetermined threshold. The flow through the system decreases as the osmotic pressure increases and scalants and foulants accumulate on the membrane. When the terminal flow is reached, the membranes must be cleaned. It is possible to process more than one batch of concentrate before a cleaning is required.

### **5.2.1.3 Chemical Pretreatment**

During New Logic Research's initial startup and optimization of the VSEP pilot unit, RO concentrate was initially processed without the use of any chemical additives. Without chemical addition, the recovery achieved by the VSEP pilot unit was only 10%. A single antiscalant (NRL 759) was added to the batch feed tank and the performance of the unit was re-evaluated. When NRL 759 was dosed at 10 ppm, the VSEP recovery improved to 65%. Higher doses of the antiscalant did not result in noticeable improvement.

Additional improvement in recovery was achieved by lowering the pH of the VSEP feed to approximately 6 to 6.5. At this pH range, the scaling potential of calcium carbonate is reduced. Using acid addition, the recovery across the VSEP unit was improved to 80 to 90%. Figure 13 illustrates the results of the initial pretreatment investigations. The membrane flux was sustained over the batch most effectively using a combination of antiscalant and pH adjustment.

After the initial optimization was completed, a second phase of optimization was conducted in which the following aspects of VSEP operation were investigated:

- Use of hydrochloric or sulfuric acid
- Timing of acid addition for pretreatment
  - A single acid addition event at the beginning of a batch
  - Adjustment of pH at the beginning of the batch, and again once a recovery of 50-65% was reached
  - Adjustment of pH during the batch only when the recovery reached 50-65%.
- Degree of pH adjustment necessary (pH 6.0 versus 6.5)

#### **5.2.1.3.1 Acid Type**

Over the duration of the VSEP pilot test, two types of acid were used for pH adjustment (pretreatment): 31.7% hydrochloric (muriatic) acid and 40% sulfuric acid. Figure 14 provides examples of two batches in which the VSEP feed water was pretreated with sulfuric and hydrochloric acids. The feed water was adjusted to pH 6 at the beginning of the batch and again midway through processing. As can be seen in the figure, the acids are similarly effective in maintaining the membrane flux throughout the batch. With respect to VSEP permeate water quality, when hydrochloric acid was used, the average sulfate concentration in the VSEP permeate was 12 mg/L. When sulfuric acid was used, under similar operating conditions (80-85% recovery and pH 6), the average VSEP permeate sulfate concentration was 19 mg/L.

#### **5.2.1.3.2 pH Adjustment Method**

The initial optimization of the VSEP pilot unit demonstrated that pH adjustment of the feed water improved recovery. The method for pH adjustment was further refined in subsequent investigations. Figure 15 shows some of the results of the pH adjustment trials in which acid was added to the feed tank:

- Only once a recovery of 50 to 65% had been reached

- At the beginning of the batch, and again when a recovery of 50 to 65% was reached to maintain a pH of approximately 6 in the feed tank
- At the beginning of the batch only

As Figure 15 illustrates, all three approaches were able to achieve 80% recovery, however, the flux was more stable throughout the batch and higher at the end of the batch for Batches 16 and 20, which used pH adjustment initially. During Batch 20 pH was also adjusted again at a recovery of 60%. Throughout the numerous batches processed, the approach of adjusting pH initially consistently resulted in a more stable flux throughout the batch and a higher terminal flux at the end of the batch. Adjusting the pH again later in the batch did not provide significantly different or better results than a single, initial pH adjustment. Maintaining a higher flux rate over more of the batch, as is achieved by adjusting the pH at the beginning of the batch, results in less membrane area required (i.e., less capital cost) to treat the same volume.

#### **5.2.1.3.3 Degree of pH Adjustment**

The amount of acid used per 1,000-L batch typically ranged from 1,500-2,500 mL (of 40% sulfuric acid). For a full-scale system, the cost of chemicals for the system operation must be balanced with the capital costs of the VSEP membranes (membrane area required based on flux). For this reason, several runs were completed to compare the performance of the system at pH 6 versus pH 6.5. Some of these runs are presented in Figure 16. For these runs, the pH was only adjusted at the beginning of the batch. While the trends in flux over the batch were similar at pH 6 and 6.5, the flux for pH 6.5 was generally lower than that achieved for pH 6. The pretreatment acid dose was approximately 30% lower to achieve a pH of 6.5 compared to that needed to achieve pH 6. In addition to lower chemical consumption, operation at pH 6.5 requires less acid, which results in less sulfate in the feed water and less sulfate in the VSEP permeate. The capital and operational trade-offs resulting from the degree of acid adjustment would need to be considered during detailed engineering, as would the selection of inorganic acid for the pH adjustment. Other acids, such as carbonic acid, may provide a more cost-effective means of pH adjustment.

#### **5.2.1.4 Recovery**

In general, higher recovery results in less final VSEP concentrate volume, which has the advantage of minimizing the volume of VSEP concentrate that must be processed by an evaporator and/or crystallizer. A range of recoveries were tested during the pilot test, based on the results of the pretreatment investigations. Figure 17 shows the results from batches ranging from 80 to 90% recovery. The batches in the figure were pretreated with 10 ppm NLR 759 and sulfuric acid. The pH



was adjusted to pH 6 at the beginning of each batch and again at approximately 60% recovery. The system flux was stable at all recoveries tested, however at 90% recovery, a noticeable decline in flux was observed and the membranes required more chemical cleaning after every batch to restore the system flux.

#### **5.2.1.5 Cleaning**

The VSEP membranes must be cleaned on a regular basis. As part of the optimization investigations, several different cleaning strategies were evaluated. Typically for membranes, including standard RO membranes, a two-step cleaning procedure is employed: an acid clean and a basic clean. The acid clean removes scale and foulants such as carbonate minerals and some metals. The basic cleaning step removes organic materials, silica, and biofilms. For the VSEP, three types of cleanings were tested:

- Hot water flush – no chemicals
- Acid clean – using a proprietary cleaning solution from New Logic Research, NLR 404
- Basic clean – using a proprietary cleaning solution from New Logic Research, NLR 505

When only antiscalant was used for chemical pretreatment, the membrane flux was shown to be restored most effectively by NLR 404, suggesting that acid-soluble minerals were limiting the recovery of the membrane. When both antiscalant and acid were used for pretreatment of the batch feed solution, NLR 505 was most effective in restoring membrane flux, suggesting that different components, possibly organic compounds or silica, were limiting recovery under those operating conditions.

Samples of spent cleaning solutions were collected and analyzed during pilot testing. Table 12 summarizes the resulting analytical data for two cleanings with NLR 505 and one hot water flush using RO permeate. For all cleanings, the spent cleaning solution contained elevated concentrations of chemical oxygen demand (COD). NLR 505 is an organic surfactant and expected to exhibit some COD, however elevated COD was also observed in the hot water flush waste. This indicates some possible accumulation of organic material on the membranes. Additionally, barium was also elevated in the hot water flush waste, indicating potential accumulation of barium sulfate on the membranes.

Three critical observations can be made about the VSEP membrane cleaning process:

- The cleanings were able to consistently restore the membrane permeability to the original (new membrane) flux (70 gfd). This suggests that irreversible fouling, which reduces membrane life, did not occur.
- Cleaning temperature is an important variable for effective cleanings. New Logic Research recommended that the chemical cleaning solutions be 50°C for the cleaning process. During piloting, cleanings at that temperature and at colder temperatures were tested. Cleanings at 50°C were much more effective at restoring membrane flux.
- Pretreatment with acid and antiscalant may reduce the cleaning frequency required. When this pretreatment is applied, hot water flushes without cleaning chemicals between batches were sometimes sufficient to restore the flux.

#### **5.2.1.6 Removal Rates**

A summary of the VSEP permeate water quality is presented in Table 13. A preliminary estimate of average removal rates is shown in Table 14 and Table 15 (concentration and mass-based, respectively). Removal rates were estimated for those parameters with detectable concentrations in the RO concentrate (VSEP feed). Many parameters are reduced by greater than 90% on average. Similar to the primary RO unit, in many instances the upper limit of removals were not determined in the routine testing because (1) the concentrations measured in the permeate were less than the method reporting limit and/or (2) the concentrations in the influent were low and close to the method reporting limit.

With the exception of sulfate, the VSEP permeate met the treatment targets listed in Section 3.3. However, as shown on Figure 11, the VSEP permeate will be blended with the RO permeate and distillate prior to discharge. Excluding additional dilution with the distillates, blending of the pilot permeates would have a combined sulfate concentration of approximately 4 mg/L, based on 80% recovery across the primary RO system, 85% recovery across the VSEP, a primary RO permeate sulfate concentration of 1 mg/L and an overall average VSEP permeate sulfate concentration of 16 mg/L.

The VSEP concentrate quality was analyzed during the pilot test and those results are presented in Table 16.

### **5.2.2 Crystallization of the VSEP Concentrate**

The modeled RO concentrate chemistry from the GE models for the SD026 primary RO system (Table 11) was provided to New Logic Research for use in estimating the size, recovery, and

performance of a VSEP system for SD026. Based on that chemistry and on the results of the pilot test, New Logic Research indicated that a VSEP system for SD026 would consist of:

- Eight i48 modules containing RO membranes (Hydranautics ESPA flat sheet membranes)
- Pretreatment with acid and antiscalant
- Feed tanks
- CIP system

The estimated recovery for system for SD026 is 80%, similar to what was observed on the pilot scale.

The concentrate from the VSEP unit would be further treated using crystallization to produce a solid salt product for disposal. The VSEP concentrate chemistry projected for SD026 was provided to GE for their assessment of the crystallization process. GE suggested that a steam-driven, mixed-salt crystallizer with a thermocompressor be used for the VSEP concentrate. Because the VSEP modeling assumed hydrochloric acid was used for the pH adjustment of the VSEP feed (to minimize the addition of sulfate to the system), the resulting concentrate was projected to contain elevated concentrations of magnesium chloride. As a result, the materials of construction of the crystallizer require upgrading, which is reflected in the capital costs presented in Section 7.0. However, were another acid, such as carbonic acid, used for VSEP pretreatment, this materials upgrade may not be necessary.

Similar to the considerations discussed in Section 5.1, foam control and management of TOC in the system will also apply to this concentrate management configuration.

### **5.3 Intermediate Concentrate Chemical Precipitation**

Another method to reduce RO concentrate volume (and therefore also reduce the size of downstream concentrate management processes) is the use of intermediate concentrate chemical precipitation (ICCP). ICCP involves treatment of the RO concentrate to remove certain constituents such as calcium and silica that contribute to RO fouling, typically by processes such as lime softening. After treatment by ICCP, the treated concentrate can be passed through a secondary RO system to reduce the concentrate volume further and maximize permeate production. The process is illustrated conceptually in Figure 18. Both the pilot test feed water and SD026 contain concentrations of calcium carbonate and silica high enough to limit the recovery of an RO system. For this reason, these compounds were targeted for removal by the ICCP process.

Silica and silicates are generic names given to compounds derived from the polymerization of silicic acid ( $\text{Si}(\text{OH})_4$ ). In neutral pH waters (pH of 6-8), silicic acid is common and has a propensity to polymerize, eventually forming colloidal polymers of many silicon dioxide molecules linked together. Metal hydroxides, if available, can be incorporated into the polymers to form more complex silicates (Ning, 2002).

In RO systems, the polymers of silicon dioxide and silicates coagulate with themselves and other organic matter and foul membranes, reducing recovery (Ning, 2002). To remove silica from feed water, lime-soda ash softening is commonly used, which increases the pH to 10-11, allowing magnesium hydroxide and calcium carbonate to precipitate. Silica is removed during this process by adsorption onto the surface of magnesium precipitates and by precipitation of the mineral forsterite ( $\text{Mg}_2\text{SiO}_4$ ) (Sheikholeslami and Bright, 2002 and Parks and Edwards, 2007).

The rate of silica polymerization decreases at a pH above 9.5 and below 5.5. At a pH of 9.5, silicic acid mostly ionizes, preventing polymerization and minimizing fouling potential. Ionized silicic acid (present at pH of 9.5) in the presence of cations like magnesium and calcium causes particulate fouling, but a majority of the cations should be removed during the softening process (Sheikholeslami, Al-Mutaz, and Young, 2001).

The amount of silica removal via lime-soda ash softening reportedly ranges from 70% to 90% (Sheikholeslami, Al-Mutaz, and Young, 2001, Sheikholeslami and Bright, 2002, and Parks and Edwards, 2007).

### **5.3.1. Bench Test Results**

As introduced in Section 2.0, bench testing of the ICCP process was conducted to verify that scalants such as silica and calcium carbonate could be removed from the RO concentrate to allow for additional volume reduction by a second stage of RO membranes. The protocols followed can be found in Appendix A.

Table 17 presents a summary of the ICCP bench test analytical results. Table 18 displays the percent removal rates for the three final samples that were dosed with soda ash compared with the untreated RO concentrate water quality. All three final samples met or exceeded the maximum silica removal rate recorded in the literature (90%).

The optimal lime dosage, which was that used to maximize magnesium removal, resulted in the lowest final silica concentration (4.63 mg/L). Figure 19 shows the silica concentrations measured

during lime flocculation and the concentration following soda ash addition. Most of the silica that was removed during the lime addition occurred within the first 10 minutes of flocculation. Between 30 and 45 minutes of flocculation, less than 5% of additional silica (compared to the untreated RO concentration) was removed.

Along with silica and calcium carbonate, barium sulfate, calcium fluoride, calcium sulfate, and metal oxides are all known scalants or foulants that can reduce membrane recoveries. Significant barium, strontium, and fluoride removal were observed from the ICCP bench testing (Table 18), which, along with silica removal, may improve RO recovery. As expected with lime softening, calcium and magnesium were also significantly reduced.

Some boron removal was also observed during the bench test. Boron removal has been reported during the formation of magnesium silicates during lime-soda ash softening. The removal mechanism is either co-precipitation or adsorption to the magnesium silicate solids. The optimal pH for boron removal has been reported to be 10.8 (Rahman, 2009).

### **5.3.2 Secondary Reverse Osmosis System**

The ICCP bench test results for silica removal were provided to GE for their use in further evaluating lime-soda ash softening and the secondary RO system for concentrate volume reduction. GE estimates that for SD026, the required lime and soda ash dosage would be approximately 4.9 mg/L and 3.3 mg/L, respectively. These doses are similar to what was used during the bench test. Based on the available data, it is estimated that the recovery of the secondary RO system would be approximately 70%. The blended permeates are estimated to meet the water quality goals presented in Section 3.0, including 10 mg/L sulfate. pH adjustment and antiscalant would also be required after the softening process, prior to treatment by the secondary RO unit.

### **5.3.3 Crystallization of the Secondary RO System Concentrate**

The concentrate from the secondary RO system would be sent to a crystallizer to produce a solid salt product for disposal. GE evaluated crystallizers for this water, based on their modeling conducted in support of the secondary membrane system. For this flow and feed water chemistry, either a mechanical vapor recompression or steam-driven, mixed-salt crystallizer were recommended. The materials of construction recommended are different than the crystallizer proposed for the VSEP option, due to the removal of magnesium from the water in the ICCP process. As discussed in Section 5.1, pH adjustment and foam control would be required. Some removal of TOC is expected

across the lime softening process, but a purge or other pretreatment may still be necessary to manage the effects of organic materials on the precipitation process.

## 5.4 Implementation Considerations

When evaporators and crystallizers are used for concentrate management, they are typically the most expensive pieces of equipment in the treatment train. Pretreatment to optimize the water they receive is an important consideration in the overall engineering of the treatment train. In the case of SD026, two important factors to be considered in the design of a full-scale treatment facility are: (1) selection of the acids used upstream of the evaporator and crystallizer and (2) managing the effects of the total organic carbon in the feed water on the crystallization process.

Acid use (for pH reduction to manage scale) will be necessary prior to the RO system and VSEP system. The acid selected for this process will ultimately report to the evaporator and/or crystallizer and may impact the materials of construction required for this equipment. For example, if hydrochloric acid is selected for use (to minimize the sulfate concentration in discharged water), the chloride may dictate higher grade materials be used in the crystallizer. The cost of the acid, the cost of the materials of construction for the crystallizer, and the final effluent water quality must be balanced.

During their evaluation of SD026 treatment, GE indicated that a purge stream off the crystallizer may be necessary to remove TOC that has been cycled up in the treatment process and which may interfere with salt precipitation and dewatering. A purge stream, regardless of volume, may be difficult to dispose, given its high concentrations of salts. Strategies to manage the purge stream or alternatives to eliminate it (e.g., removal of the TOC upstream of the RO process by adsorption or use of an advanced oxidation process) should be considered prior to detailed engineering.

## 6.0 Permeate Stabilization

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Because RO removes dissolved constituents from water, the permeate is virtually void of minerals, and contains low amounts of calcium and alkalinity. Additionally, RO permeate often contains elevated concentrations of dissolved carbon dioxide. The carbon dioxide is formed from the reaction of antiscalant chemicals, which are added to RO feed water to prevent calcium carbonate scaling on the membranes, with bicarbonate alkalinity already present in the feed water. The resulting permeate, with low buffering capacity and low pH, is corrosive. Prior to discharge, RO permeate must be stabilized to meet the discharge water quality targets.

An effluent stabilization bench testing experiment was designed and executed with two main objectives: (1) identify a stabilization method (e.g., addition of minerals) that will reduce the corrosiveness of the blended RO and VSEP permeates and maintain compliance with the effluent water quality targets in Section 3.0, and (2) produce a non-toxic effluent. For the purposes of the bench test, “non-toxic” was defined as water that was neither acutely or chronically toxic to *C. dubia*. The measure of chronic toxicity used for this evaluation was the ability of the test organisms to produce at least 75% of the number of young as those organisms in the control water. Two known treatment technologies were tested to meet the above objectives:

- Hydrated lime ( $\text{Ca}(\text{OH})_2$ ) and carbon dioxide ( $\text{CO}_2$ ) addition
- Limestone contactors (LC)

The permeate used for testing was a blend of RO and VSEP permeate generated by the RO and VSEP pilot unit, blended at a 5:1 ratio (representing recoveries of 80% for the RO unit and 80% for the VSEP unit). The stabilization bench testing was conducted at Barr’s wastewater laboratory.

In addition to the final water quality targets for the stabilized water discussed in Section 3.0, the following additional targets were used in this evaluation to measure the corrosiveness and toxicity of the blended effluent:

- Langelier Saturation Index (LSI)  $\geq 0$
- Calcium carbonate saturation index (SI)  $> 0$
- 7-day chronic whole effluent toxicity (WET) test young reproduction  $\geq 75\%$  young reproduction of the laboratory control water sample (moderately hard laboratory water)
- $6.5 < \text{pH} < 8.5$

LSI and SI are both indices used to measure the scaling potential of calcium carbonate. Positive values for both indices indicate scale forming water (versus negative values that indicate corrosive water). The treatment targets for the stabilization tests were to obtain slightly positive values for each measure.

## **6.1 Lime Addition Bench Test**

The lime and carbon dioxide stabilization process was first modeled using PHREEQC, an aquatic equilibrium model by the United States Geological Survey (USGS). The simulation was used to estimate the lime and carbon dioxide dosages that would be required to achieve the target SI, and the resulting final pH. Table 19 displays the modeling results of the estimated optimal lime dose.

An experimental protocol was then developed using the PHREEQC model dose as a guide. The protocol included the addition of lime to the blended effluent to increase the total hardness concentration of the blended permeates, followed by addition of carbon dioxide to achieve the target SI value. The lime dose would raise the SI value of the blended effluent above the target (0.1) and the carbon dioxide would reduce it to the target value. This approach results in water with minimal carbon dioxide fugacity, which lends stability to the effluent pH and provides stable water for WET testing.

Based on the modeling results shown in Table 19, a range of hydrated lime doses were added to the blended permeates and then the water was titrated down to a pH of approximately 7.3 using carbon dioxide during the bench tests.

### **6.1.1 Experimental Setup**

The lime addition tests were conducted in a 4-L Erlenmeyer flask. A range of hydrated lime doses (Table 20) were added to 3-L aliquots of the blended effluent and were mixed vigorously on a stir plate. The samples were then titrated to a pH of 7.3 using a 5%:95% carbon dioxide and nitrogen gas mix. Final titrated blend samples were submitted to external laboratories for analytical and WET testing.

The hydrated lime used in the bench testing experiments was 94.3%  $\text{Ca}(\text{OH})_2$ .

### **6.1.2 Results**

#### **6.1.2.1 Stabilized Water Chemistry**

Table 20 presents a summary of the stabilization bench test results. Doses 4, 5, and 6 all met the calcium carbonate scaling potential water quality targets described in Section 6.1. Dosages 1, 2, and



3 did not have enough hardness and alkalinity to result in a positive LSI or SI value, indicating the final samples were still corrosive. The following observations can be made about the results of lime addition:

- turbidity - dosages 4, 5, and 6 may exceed potential future turbidity requirements
- TSS – doses 4 and 6 may exceed the potential future total suspended solids requirements
- aluminum – doses 3, 4, 5, and 6 may exceed potential future aluminum requirements (Class 2B water quality standard)

The results for these three parameters were likely affected by the grade of hydrated lime, lime contact time, and dosing methods. Excess turbidity and TSS likely, in part, resulted from the experimental setup and can be mitigated. Section 6.1.3 contains additional discussion of these issues.

#### **6.1.2.2 Whole Effluent Toxicity**

Based on the results from the bench testing, Dose 4 would likely produce the most stable blended effluent for the system. The LSI and SI values indicate the water would not be corrosive and the WET testing suggests the stabilized blended effluent would meet the WET requirements.

Figure 20 displays the mean number of young produced per female for each dose compared to 75% of the control. Note that the, unstabilized water achieved a mean young production that was 53% of the control (i.e., an observable toxic effect). Doses 2-6 produced effluent that achieved a mean number of young produced per female of at least 75% of the control, suggesting that the stabilization approach reduced toxicity as intended, despite the introduction of aluminum described in the previous section. Dose 4 resulted in a mean young production higher than the control.

#### **6.1.3 Implementation Considerations**

Dose 4 was identified as the best dose for the blend of permeate that was tested. Residual turbidity is a known operational challenge of using a lime addition to stabilize RO effluent (AWWA, 2007). As listed above in Section 6.1.2.1, lime doses 4 through 6 all had elevated turbidities. If lime addition is the chosen method of effluent stabilization, turbidity could be managed using the following techniques:

- High quality lime – Using high quality lime reduces the amount of inert material present to contribute to TSS and turbidity. For project implementation, the lime product used should be greater than 94% hydrated lime (purity used for bench testing) if available. High quality lime

also has a high specific surface area which helps to maximize reactivity and minimize grit (Hart, 2007).

- Liquid lime dosing – Dosing the lime as a liquid slurry rather than a solid provides minimal turbidity increases as less inert materials are present in liquid lime, and it avoids maintenance issues associated with dry lime (Lozier, et al., 2010).
- Lime contact chamber – Contact chambers provide the necessary turbulent mixing time for the lime to fully dissolve into the blended effluent. The mixing or contact time is a key design parameter and is typically between 5-10 minutes (AWWA, 2007)

When the lime is initially dosed to the blended effluent, some of the dissolved carbon dioxide reacts with the lime and calcium carbonate precipitates and turns the mixture cloudy. As additional mixing time is allowed in the lime contact chamber, the remaining carbon dioxide reacts and dissolves the newly formed calcium carbonate, thereby reducing the turbidity.

The aluminum measured in the stabilized water from the bench tests originated from the hydrated lime product. Using the measured aluminum and calcium concentrations, it is estimated that the lime product that was used contained approximately 0.23% aluminum by weight. In order to achieve the 125 ug/L effluent aluminum concentration using Dose 4 (Table 20), the lime product would have to contain less than 961 mg aluminum/kg hydrated lime product. Below is a list of the closest lime suppliers to the project area and the standard aluminum concentration in their lime product:

- Graymont – hydrated lime product contains 1,059-2,118 mg aluminum/kg hydrated lime product
- Carmeuse Lime & Stone – hydrated lime products contain on average 963 mg aluminum/kg hydrated lime product
- Linwood Mining & Minerals – does not test for aluminum separately

The above concentrations indicate that identifying a supplier that can provide a lime product with consistently less than 961 mg aluminum/kg hydrated lime within a reasonable shipping distance will be an important consideration for this stabilization option.

## **6.2 Limestone Contactor Bench Test**

The limestone contactor (LC) system is a semi-passive stabilization option that passes the blended effluent through a crushed limestone bed. As the blended effluent contacts the limestone media, it dissolves the limestone ( $\text{CaCO}_3$ ) increasing both the hardness and alkalinity of the blended effluent.

The rate of limestone dissolution is an important design parameter for an LC system. Three different hydraulic loading rates were tested on three identical LCs to identify the rate that would result in adequate introduction of hardness and alkalinity to the blended permeate.

As the effluent from the LC columns was anticipated to still have a low LSI, due primarily to remaining dissolved carbon dioxide, air stripping and caustic addition were tested for final pH adjustment.

The objectives of this bench test were as follows:

- identify the maximum hydraulic loading rate that would achieve the treated water quality targets outlined in Section 6.1
- identify the best post-LC treatment to achieve the treated water quality targets outlined in Section 6.1

### **6.2.1 Experimental Setup**

The LCs were constructed as 6-foot long, 2-inch diameter upflow columns (Figure 21). The tests were conducted using two types of limestone media:

- $\frac{3}{4}$ -inch crushed landscaping limestone
- Columbia River Carbonates' Puri-Cal RO product with a particle size range of 2-3.4 mm

Before the tests were conducted, the media was washed to remove fines. Also, for both tests, the blended effluent was pumped at three different hydraulic loading rates through three identical upflow LCs using a peristaltic pump.

The test program is illustrated in Figure 22. The first 2-L of effluent from each LC was discarded and the next 6-L of sample from each LC was collected for analysis. 2-L of the collected sample was sparged with compressed air, 2-L was dosed with caustic soda, and the final 2-L was not amended. All samples were submitted for analytical and WET testing. Turbidity values were measured upon collection using a field turbidimeter.

## **6.2.2 Results**

### **6.2.2.1 Stabilized Water Chemistry**

The  $\frac{3}{4}$ -inch crushed landscaping limestone media resulted in an insufficient amount of alkalinity and hardness in the LC effluent. The Puri-Cal RO product has a higher specific surface area and allowed

for more CaCO<sub>3</sub> dissolution. Table 21 presents a summary of the results from the testing using the Puri-Cal RO product.

The following observations can be made regarding the results of the LC bench tests:

- turbidity – Only the caustic dosed Rate 3 sample exceeded the goal
- total suspended solids – Only the caustic dosed Rate 3 sample exceeded the goal
- metals – None of the samples exceeded any listed targets

Samples collected from the ¾-inch limestone testing were subjected to low-level mercury analysis. None of the samples had a detectable amount of mercury present, and therefore mercury was not tested for in the second round of LC testing.

#### **6.2.2.2 Whole Effluent Toxicity**

Figure 23 displays the mean number of young produced per female for the LC treatments, compared to 75% of the control sample's reproduction. As shown in the figure, the unstabilized permeate would not likely pass the IC25 criterion. The Rate 1 no treatment and sparged samples and the Rate 2 sparged samples produced effluent that achieved a mean number of young produced per female of at least 75% of the control.

#### **6.2.3 Implementation Considerations**

The LC bench test results suggest that a limestone bed hydraulic loading rate (HLR) of 2.4 gpm/sf (using the Puri-Cal RO product), followed by air sparging, is able to produce a stabilized effluent that meets the treatment targets. However, in addition to HLR, there are other factors that will need to be considered for full-scale stabilization, such as residence time and bed depth.

For upflow contactors, HLRs ranging from 1.0-17.2 gpm/sf are typical (Shih, et al., 2012). The HLR is related to the flow rate of the LC system required for a given reactor diameter. The highest HLR that achieves the treated water quality targets minimizes the number of LCs required to stabilize the blended effluent flow. However, HLRs that are too high can cause media blowouts causing turbidity and TSS.

The residence time of the system is related to the dissolution rate of the limestone. Typical empty bed contact times (EBCT) range from 3.6 to 30 minutes for LC systems (Shih, et al., 2012). Required residence times are related to the limestone media size. Larger diameter media has lower specific surface area which requires longer residence times to allow for adequate dissolution of the media.

After the residence time and the HLR are defined, the volume and therefore the bed depth of the LC can be calculated. The calculated bed depth represents the minimum depth of media required to meet the treatment targets that must always be maintained.

As mentioned above, LC systems are semi-passive. The limestone will need to be replaced periodically as it dissolves. If the blended permeate is applied at 2.4 gpm/sf to the LCs and the system is operated 24 hours/day, then 3.38 pounds of limestone per day per square feet of LC will need to be replaced. The frequency of media replacement would need to be considered in the amount of limestone that is provided beyond the minimum depth that must be maintained for stabilization.

Sparge systems are commonly added as a post treatment following LCs to strip any excess dissolved carbon dioxide remaining in the effluent. The dissolved carbon dioxide will likely off-gas at the discharge point if not removed at the treatment site. Off-gassing will cause a pH increase which is known to contribute to failed WET tests. Stripping the carbon dioxide before it reaches the final discharge point will produce a more pH-stable water.

Upflow contactors were constructed for this bench test and are the most common LC, but downflow contactors are also used. Upflow reactors typically result in a lower effluent turbidity and do not require backwashing, but an internal top screen does need to be used to prevent calcite from blowing out of the reactor. Downflow reactors provide calcite dissolution and sediment filtration.

Disadvantages of downflow configurations include required backwashing, high turbidity waste streams, increased risk of TSS in the treated effluent from fines breakthrough, and higher capital and operational and maintenance costs (Shih, et al., 2012).

The upflow configuration was selected for this application because of the typically lower turbidity effluent and no backwashing requirement.

## 7.0 Preliminary Estimates of Implementation Costs

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### 7.1 Basis of Cost Estimates

The pilot and bench testing work conducted as part of this project, along with the associated literature and modeling studies of concentrate management, were used to refine the RO treatment train originally presented in the *Short-Term Mitigation Evaluation and Implementation Plan for SD026*, and to update the estimated capital costs for the system. The three treatment train concepts used to develop cost estimates are shown in Figures 10, 11, and 18. Table 22 provides an overview of the basic treatment information used to develop the capital and operating costs for the systems.

The cost estimates provided in this report represent updates to the costs of treatment presented in the *Short-Term Mitigation Evaluation and Implementation Plan for SD026*. The costs of site work and utilities have not been updated. Additionally, the flow rates used as the basis of the cost estimates assume that some equalization of the flow at SD026 will be provided to reduce the peak flow requiring treatment, and therefore reduce the cost of the required installed treatment system capacity. The historical average flow rate from SD026 is approximately 450 gpm. This value was used to estimate the annual operating costs for each of the options. The installed capacity for capital costs assumes a peak influent flow rate of 600 gpm. The cost of equalization is not included in the estimates.

For the selected treatment options presented in this report, the capital, operation and maintenance, and present worth costs have been developed for the purpose of evaluating and comparing alternatives. The cost estimates developed for this report are considered conceptual level costs or Class 5 estimates, according to the Association for the Advancement of Cost Engineering International. The typical associated level of accuracy of Class 5 cost estimates is  $\pm 25$  to 100%.

The capital and operation and maintenance costs were developed using a number of sources listed below, recent local pricing for similar equipment, and general resources:

- RO system and evaporator and crystallizer capital and operating costs were obtained from GE budgetary quotes
- VSEP capital and operating costs were obtained from New Logic Research budgetary quotes

- Lime costs were obtained from the United States Geological Survey on-line mineral commodity summaries (<http://minerals.usgs.gov/minerals/pubs/commodity/lime/>)
- Acid prices were obtained from ICIS on-line (<http://www.icis.com/>)
- The cost of electricity used was \$0.055/kW-h.
- Office of Management and Budget (1992), Guidelines and Discount Rates for Benefit Cost Analysis for Federal Programs, Circular A-94, Appendix C, updated December 2012.
- U. S. EPA (2000), A Guide to Developing and Documenting Cost Estimates During the Feasibility Study, U.S. EPA 540-R00-002.

The following assumptions were used in developing the opinion of probable cost:

- The capital costs assume a 20-year equipment life for mechanical equipment.
- No redundant capacity is included in the costs for the treatment equipment.
- A real discount rate of 0.8% and a time frame of 20 years were used in calculating the net present value of the operation and maintenance costs. Present value analyses are typically conducted for the duration of the project (U.S. EPA, 2000). In this case, the duration of the project has been assumed to be the life of the equipment, which is approximately 20 years. The conceptual treatment facilities presented in this report were developed with the assumption that they could be incorporated into a long-term treatment strategy, if necessary.
- A 30% contingency has been included in the capital and O&M costs to account for items not detailed in the estimate but known to be part of the project such as process pumps, piping and supports, painting and protective coatings, process ancillary equipment, spare parts, operation and maintenance consumables, contractor mobilization and demobilization, and demolition.
- A 10% contingency has been included for professional services and reflects the lesser degree to which changes in capital items impact the cost of required engineering services.
- As discussed earlier, the range of accuracy for the costs presented is  $\pm 25$  to 100%. This reflects the uncertainties associated with the scope of the project at this time, including: site and subsurface conditions, costs of materials and services, and utility requirements. This degree of accuracy falls within the level of accuracy suggested for screening and conceptual development of alternatives by the U.S. EPA (U.S. EPA, 2000).

As noted, the preliminary cost estimates were made using a variety of references, along with professional experience from qualified professionals familiar with the project, and with the assistance of potential water treatment vendors. The estimates are based on project-related information, including the water quality as described in this report and the general characteristics of the site (based on field observations) and include a contingency factor to account for items that are unknown at this time. While the costs of some specific items for a specified set of conditions can be determined with precision, the factors controlling the design conditions, namely the actual water quality and site conditions, are still highly variable. The high potential for changes in these controlling values precludes a lower contingency in cost estimates at this stage of the project.

The potential actual costs for implementation of any technology selected based on the evaluation described in this plan would be expected to change as more detailed information (site survey, geotechnical testing, etc.) becomes available, further design is completed, or as the project needs change. PolyMet, CE, and Barr cannot and do not guarantee that proposals, bids, or actual construction costs will not vary from the preliminary cost estimates prepared in this report as any approved mitigation/treatment plan is implemented.

## **7.2 Preliminary Estimates**

### **7.2.1 Reverse Osmosis with Evaporation and Crystallization**

A summary of the required treatment equipment is provided below, and Table 23 provides a summary of the estimated capital and operating costs for this treatment option.

Major equipment required includes:

- Greensand filtration:
  - 3 filter units
  - Potassium or sodium permanganate storage and feed system
  - Filtrate tank
- RO system:
  - Acid storage feed system
  - Antiscalant feed system
  - Sodium bisulfite feed system
  - RO skid with feed pumps, membrane housings, and membrane elements
  - Clean-in-place (CIP) skid
- Evaporator system:



- Evaporator feed tank and mixer
- Evaporator feed pump
- Heat exchangers
- Feed deaerator
- Evaporator vessel
- Vapor compressor
- Distillate tank
- Distillate pump
- Evaporator recirculation pump
- Seed pump, tank, and mixer
- Seed recycle system
- Acid pumps
- Caustic pumps
- Antiscalant and antifoam pumps
- Crystallizer system:
  - Calandria crystallizer with internal heater
  - Recirculation pump
  - Crystallizer feed tank, pump, and mixer
  - Dewatering feed pump
  - Filter press
  - Filtrate pump
  - Distillate tank and pump
  - Antifoam pump
  - Caustic pump
- Permeate stabilization system
  - 3 limestone contactor units
  - 2 degassifiers and blowers

The estimated building area for this treatment equipment is approximately 6,200 sf. The revised preliminary capital cost estimate for this option is \$19.7M, with an annual operation and maintenance cost of \$1.5M. This treatment train is similar to that first evaluated in the Short Term Mitigation Plan. The revised cost estimate is substantially higher than estimated in that report. The capital cost increase is driven largely by the higher cost of the evaporator and crystallizer equipment. The costs for that equipment were previously developed based on published cost curves (Mickley, 2006) and

escalated using Engineering News Record (ENR) indices. The current estimate is based on budgetary estimates from GE, obtained in May 2013. The cost of this type of equipment has increased at a rate significantly greater than the ENR Construction Cost Index (CCI).

### **7.2.2 Reverse Osmosis with VSEP and Crystallization**

A summary of the required treatment equipment is provided below, and Table 24 provides a summary of the estimated capital and operating costs for this treatment option.

Major equipment required includes:

- Greensand filtration:
  - 3 filter units
  - Potassium or sodium permanganate storage and feed system
  - Filtrate tank
- RO system:
  - Acid storage feed system
  - Antiscalant feed system
  - Sodium bisulfite feed system
  - RO skid with feed pumps, membrane housings, and membrane elements
  - CIP skid
- VSEP system:
  - 2 feed tanks
  - Antiscalant pumps
  - Acid storage tank and feed system
  - Eight i48 modules containing RO membranes
  - CIP skid
- Crystallizer system:
  - Steam-driven, mixed-salt crystallizer vapor body
  - External crystallizer heater
  - Recirculation pump
  - Crystallizer feed tank, pump, and mixer
  - Dewatering feed pump
  - Filter press
  - Filtrate pump
  - Distillate tank and pump

- Antifoam pump
- Caustic pump
- Permeate stabilization system
  - 3 limestone contactor units
  - 2 degassifiers and blowers

The estimated building area for this treatment equipment is approximately 6,200 sf. The revised preliminary capital cost estimate for this option is \$24M, with an annual operation and maintenance cost of \$1.6M. For this treatment train configuration, the VSEP unit acts as the brine concentrator, replacing the evaporator. However, a larger crystallizer is required because the VSEP does not reduce the volume of RO concentrate to the same degree as the evaporator.

### **7.2.3 Reverse Osmosis with ICCP, Secondary RO System, and Crystallization**

A summary of the required treatment equipment is provided below, and Table 25 provides a summary of the estimated capital and operating costs for this treatment option.

Major equipment required includes:

- Greensand filtration:
  - 3 filter units
  - Potassium or sodium permanganate storage and feed system
  - Filtrate tank
- Primary RO system:
  - Acid storage feed system
  - Antiscalant feed system
  - Sodium bisulfite feed system
  - RO skid with feed pumps, membrane housings, and membrane elements
  - CIP skid
- ICCP system:
  - Lime storage
  - Soda ash storage
  - Slurry tanks and pumps
  - Solids contact clarifier
  - Solids dewatering equipment
- Secondary RO system:

- Feed tank
- Acid feed pump
- Antiscalant feed pump
- RO skid with pump, membrane housing, and membrane elements
- Crystallizer system:
  - Steam-driven, mixed-salt crystallizer vapor body
  - External crystallizer heater
  - Recirculation pump
  - Crystallizer feed tank, pump, and mixer
  - Dewatering feed pump
  - Filter press
  - Filtrate pump
  - Distillate tank and pump
  - Antifoam pump
  - Caustic pump
- Permeate stabilization system
  - 3 limestone contactor units
  - 2 degassifiers and blowers

The estimated building area for this treatment equipment is approximately 5,600 sf. The revised preliminary cost estimate for this option is \$14.8M, with an annual operation and maintenance cost of \$1.9M.

### **7.3 Discussion**

The use of mechanical evaporation and/or crystallization for concentrate management is a capital-intensive approach; however, concentrate disposal options in the project area are limited. Therefore, other methods to reduce the capital, operating, or overall lifecycle cost were investigated for this project. The 20-year lifecycle costs for each option are presented in Table 26. The baseline concentrate management option – evaporation and crystallization – relies largely on heat energy to reduce the concentrate volume for disposal. The use of the VSEP process replaces some of the heat energy required for volume reduction with mechanical energy, and the use of ICCP and secondary RO membranes replaces some of the heat energy with chemical and mechanical energy. One of the focuses of this investigation was to determine the relative efficiencies of each of these forms of energy for concentrate volume reduction. From a capital cost perspective, the replacement of the

evaporator with ICCP and secondary RO membranes is the most cost-effective of the three options. However, its operating costs are higher, largely due to the costs of the chemicals required and solids disposal. For all of the treatment options, it is important to note that the cost of chemicals is a significant component of the overall operation and maintenance cost. The unit prices for chemicals such as mineral acids can be volatile. Because of this volatility and because the annual operation costs for the options are of similar magnitude, the relative lifecycle costs can change, depending on the market pricing of these commodities at any given time. Were one of these treatment options to be selected for implementation, special consideration should be given to the selection of acids used for pH adjustment, as their selection can have multiple implications, including:

- Final discharge water chemistry – The type of acid and the grade of acid selected can impact the discharged water quality.
- Operating costs – The pricing of certain mineral acids, such as hydrochloric acid, is volatile and other acids may provide more stable and predictable pricing.
- Capital costs – The use of hydrochloric acid, which was used in this study, may necessitate higher grade materials of construction for concentrate management equipment to reduce chloride corrosion.

## 8.0 Conclusions

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The pilot testing program and associated evaluations and modeling have confirmed that low pressure RO membranes can effectively reduce the concentration of sulfate and parameters of concern of the discharge at SD026 to less than or equal to the water quality goals listed in Section 3.3. The RO system requires pretreatment by the following process:

- Greensand filtration for TSS, iron, and manganese removal
- Antiscalant to mitigate mineral scaling
- pH adjustment with acid to reduce the formation of calcium carbonate scale

The pilot testing program yielded several important results, including the following for the RO system:

- throughout the testing program, the RO system consistently produced permeate with sulfate concentrations less than 10 mg/L
- the pretreatment methods selected for the RO system—greensand filtration and antiscalant addition— were effective in maintaining stable RO performance
- the RO system did not experience significant fouling or scaling during the testing program
- the RO system operated stably and effectively at 80% recovery on the pilot feed water, similar to what is projected for water from SD026.

Concentrate management and permeate stabilization were assessed as part of the evaluation of active treatment options for SD026. The ability to manage RO concentrate is a critical component of the RO technology's feasibility for any project. Based on the initial evaluations conducted in the *Short-Term Mitigation Evaluation and Implementation Plan for SD026*, creation of a solid salt for disposal from the RO concentrate was selected for more detailed evaluation. In order to produce the solid salt, three volume-reduction approaches were investigated:

- Evaporation and crystallization
- VSEP and crystallization
- ICCP, secondary RO, and crystallization

The VSEP product was pilot tested for over 2 months. The VSEP pilot test yielded the following results:

- The VSEP sulfate removal efficiency averaged 99.3%. Under the pilot test conditions, when the VSEP and RO permeates are blended, the sulfate concentration is less than 10 mg/L.
- The VSEP system has demonstrated recoveries ranging from 80 to 90%.
- No irreversible fouling was observed during the course of testing. Once cleaning optimization was complete, the membrane flux was restored to its original flux after each cleaning.
- No decline in sulfate removal was observed over time.

ICCP testing was conducted on the bench scale. The bench testing demonstrated silica removal similar to that which has been reported in the scientific literature. Other potential scalants (barium, calcium, fluoride, and strontium) were also removed during the lime-soda ash softening process. Modeling of the secondary RO process suggests that 70% recovery is achievable for the secondary RO system and that blended permeates would produce a final effluent that meets the water quality goals.

Evaluations of evaporation and crystallization were conducted with assistance from GE, a manufacturer of this equipment. Products capable of dewatering solutions of mixed salts were recommended by GE. Technical considerations for their implementation included selection of the materials of construction (based on the salts present), the need for pH control, foam management, and the need for a purge stream due to the presence of organic materials in the water. The organic materials, which are natural organic matter that would be cycled up in concentration, can interfere with the precipitation process. The management and disposal of a purge stream from the crystallizer may present significant technical obstacles. Additional treatment upstream of the RO system to remove or mineralize those materials prior to treatment may be necessary.

Effluent stabilization is also necessary after treatment to adjust the pH and reduce the corrosiveness of permeates and distillates prior to discharge. The permeate stabilization bench testing results produced the following conclusions:

- Lime addition
  - Lime addition was able to adjust the pH and meet most water quality targets, including measures of corrosiveness
  - Two important factors were identified in the test that would need to be considered on a full-scale design:

- Quality of lime used (to reduce turbidity from inert materials and minimize unwanted aluminum in the discharge)
  - Method of lime addition and reaction to minimize residual turbidity
- Limestone contactor
  - The limestone contactor was able to adjust the pH and meet all water quality targets, including measures of corrosiveness.
  - Additional treatment after limestone contactor was needed to remove remaining carbon dioxide (e.g., air sparging).

Preliminary capital and operating costs were developed for the three treatment approaches evaluated. Capital costs ranged from \$14.8M to \$24M and operating costs ranged from \$1.5M to \$1.9M annually. These costs are higher than previously estimated for the *Short-Term Mitigation Evaluation and Implementation Plan for SD026*. The costs differences reflect (1) the availability of more complete water quality data for SD026, (2) pilot and bench testing results which informed not only the equipment requirements but also the chemical treatment needs for the processes, and (3) updated budgetary estimates from the major equipment suppliers.

Were any of the investigated approaches to be selected for implementation at SD026, the following additional tasks are recommended:

- Work with the evaporator and/or crystallizer supplier to select the chemicals used throughout the treatment train (and which report to this equipment) to optimize capital and operating costs, as well as to meet water quality objectives;
- Work with evaporator and/or crystallizer supplier to investigate alternatives for managing necessary purge streams, including addition of equipment upstream of the RO to remove organic compounds;
- Work with the evaporator and/or crystallizer supplier to optimize heating options for the equipment to minimize operating costs;
- Evaluate the level of equipment redundancy required to reliably treat the discharge;
- Evaluate regional disposal options for the solid, mixed salt that will be produced.



## 9.0 References

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