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University of Nevada, Reno

**Barrick Gold Senior Design Project: Treatment and Removal of Excess Lixivants  
from the Tailings Water of Two Parallel Gold Leaching Circuits**

A thesis submitted in partial fulfillment of the  
requirements for the degree of

Bachelor of Science in Chemical Engineering and the Honors Program

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**UNIVERSITY  
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**BACHELOR OF SCIENCE IN CHEMICAL ENGINEERING**

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## **Abstract**

The following project was completed by a Senior Design team of the Department of Chemical and Materials Engineering at the University of Nevada, Reno. The problem statement was given to the team by a representative at Barrick's Goldstrike mine, located outside of Carlin, Nevada. The statement describes the mine's present operation of two gold leaching circuits, with one circuit being a newly constructed process running separately from the existing cyanide leach operation. The statement then communicates a desire to combine the waste streams (tailings) of each circuit into a single holding dam prior to recycling, effectively increasing the throughput of ore in the new circuit significantly. The newly constructed Resin in Leach (RIL) circuit uses Thiosulfate, which is an alternative lixivant that has proven to be more effective in extracting gold from double-refractory ore. However, it is difficult to prevent the competing reagent from negatively affecting gold recovery in the opposite leach circuit. In order to combine the wastewater tailings from the two processes into a single holding dam, these streams must be treated before they are mixed to forestall any deleterious effects caused by the cross contamination of the two lixivants in each circuit. The senior design team investigated and evaluated several designs of unit operations for reagent reclamation; various viable options are presented.

## **Acknowledgments**

The senior design team would like to first thank our thesis advisor and professor, Dr. Alan Fuchs. His experience and insight throughout our investigation proved invaluable. At Barrick Gold, we would like to thank Wayne Douglas and his associates for their encouragement and attention to our progress, as well as their continued support of the University of Nevada. We would especially like to thank Mike Kivistik for sharing his knowledge in conducting experiments and how to conduct them safely; we could not have done this without his help. Lastly, the design team leaders would like to extend their gratitude to the truly dedicated team members who did more than their share of work to complete this project.

## Table of Contents

|   |    |
|---|----|
| 1. Introduction to Gold Leaching .....                  | 1  |
| 2. Problem Statement .....                              | 3  |
| 3. Thermal Decomposition in Autoclave and Roaster ..... | 3  |
| 4. Experimentation .....                                | 6  |
| 4.1 CIL Experimentation .....                           | 6  |
| 4.2 RIL Experimentation .....                           | 9  |
| 5. Alternative Cases .....                              | 12 |
| 5.1 Thiosulfate Treatment with Ion Exchange .....       | 12 |
| 5.2 Thiosulfate with Hydrogen Peroxides .....           | 18 |
| 5.3 Cyanide Treatment with Hydrogen Peroxide .....      | 19 |
| 5.4 Reverse Osmosis .....                               | 22 |
| 5.5 SPARRO Reverse Osmosis .....                        | 23 |
| 5.6 UV Treatment .....                                  | 30 |
| 5.7 Solar Thermal Treatment .....                       | 32 |
| 6. Conclusion .....                                     | 35 |
| 7. Future Work .....                                    | 35 |
| 8. References .....                                     | 37 |

## **Introduction to Gold Leaching**

Precious metals have long captured the attentions of countless generations, from the ancient Bronze Age to the current generation's urgency in obtaining gold plated Apple watches. As old as the fascination is, so too is the obvious profitability of such minerals. Placer mining in the High Sierras brought the United States to the gold standard, and thousands of men and woman set out to make their fortunes in crystal clear streams pouring off the Northern California mountainsides. The largest advance in the lucrative mining business, however, is the development of leaching circuits on an industrial scale. The construction of large scale chemical processes has since put the sluice box and mineshafts to shame, making it possible to move thousands of tons of earth a day to extract the slight increases in gold concentration on the parts per million scale.

Gold is currently worth over one thousand U.S. Dollars per ounce, which is a pretty hefty price tag. Companies like Barrick Gold Corp. can take advantage of this highly valuable product by literally moving mountains to find gold, and still turning a profit. Because of the enormous profit margin, a mining outfit can afford to process thousands of tons of earth every day. Gold-rich ore (comparatively) is ground down into fine particles using ball mills, and then is subjected to high heat and pressure (roasting) to oxidize the gold from the common sulfur compounds it is normally bound in. Once freed, gold can be dissolved ("leached") into solution by a chemical reactant colloquially referred to as a lixivant. The most common lixivant used in the mining industry is Cyanide, which gold readily dissolves in. After the liquid is separated from the remaining solid ore, the gold-cyanide complex is trapped onto granules of activated carbon, which can then be removed from the leaching circuit. The gold can then be solidified (eluted) by the

application of an electrical potential gradient, and can be further purified into the iconic gold bars.

Not all lixivants are equal, however, and while Cyanide is the reliable standby, it is not without its drawbacks. Cyanide is a very toxic chemical (think cyanide pills) and industry must always be vigilant to prevent its release into natural waterways. It also does not perform as efficiently when faced with certain grades of mineral ore, which has prompted scientists to find more suitable alternatives. Within the last two decades, Thiosulfate has been gaining usage in the leaching gold from double refractory ore. Single refractory ore is termed as such to indicate that there is only one component that blocks the dissolution of gold into solution. In the industry, this nomenclature is understood as ore containing Sulfur as the most prevalent refractory component. Cyanide has been used for decades as a dependable leaching agent for this type of ore after a pass through a roaster. Double refractory ore is understood to mean that the gold is refracted mostly by both Carbon and Sulfur; Cyanide does not meet performance standards.

Barrick's Goldstrike mine has been facing increasing levels of double refractory ore as the mine reaches a new chapter of its life, and therefore millions of dollars have been spent by the company to construct and operate a secondary leaching circuit, using Thiosulfate as a lixivant and a specialty resin to extract the gold-thiosulfate complex that forms. The integration of this new process leads to a novel set of circumstances from an operational standpoint. The two separate systems should share as much existing infrastructure as possible to reduce operating costs, but it is not known how the increasingly complex chemical equilibrium of the constantly recycled process water will

affect gold recovery at various stages of the operations. This leads to the problem statement proposed by Wayne Douglass and his associates.

## **Problem Statement**

The team's initial problem statements was provided by a team of metallurgists working at the Barrick Goldstrike mining operation. The problem that Barrick is facing at Goldstrike is the replacement of a cyanide leaching process with a new thiosulfate resin-in-leach process. The problem is that the roaster and the autoclave circuits will be using different reagents for gold leaching that affect each process differently and therefore cannot be recycled directly from the tailings pond. For this reason, Barrick has asked the senior design team to determine the effects of co-mingling the tailings from each process and provide possible treatment options to combat any negative effects to either leaching operation.

## **Thermal Decomposition in Autoclave and Roaster**

In our final meeting at Barrick, it was revealed to us that cyanide and thiosulfate decompose in the autoclave and roaster, but reform further down the process. Even so, experiments [1][2] have been conducted to characterize the kinetics of the decomposition. In alkaline conditions with a  $\text{pH} > 10$ , the decomposition of cyanide and thiosulfate were observed with an elevation in temperature and oxygen pressure. Temperatures varied from 393 – 473 K and 373 – 411 K, and oxygen pressures varied from 2.0 – 8.0 MPa and 0.35 – 1.0 MPa for cyanide and thiosulfate, respectively<sup>[1][2]</sup>. There were two pathways

observed with cyanide and its decomposition through a wet oxidation. The first pathway had  $\text{HCONH}_2$  as a intermediate compound and gave  $\text{NH}_3$  and as by products, while the second pathway produced  $\text{CNO}^-$  as the intermediate compound with  $\text{CO}_2$ ,  $\text{N}_2$ ,  $\text{H}_2$ , and  $\text{OH}^-$  as the final products, as seen in Figure 1.

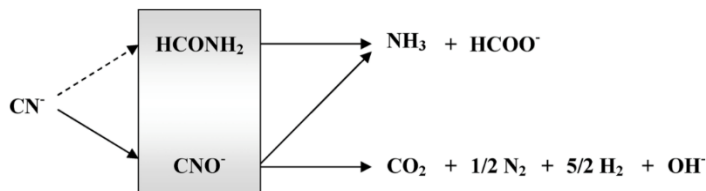


Figure 1: Two different pathways for the degradation of cyanide through wet oxidation<sup>[1]</sup>

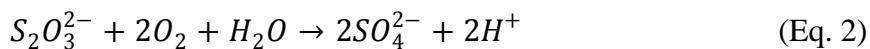
With a pH held at 11, it was found that oxygen pressure contributed a very small change to the rate of decomposition for cyanide. There was a 95.5% decomposition of cyanide after nine minutes and an oxygen pressure of 2.0 MPa, while 8.0 MPa of oxygen pressure resulted in 96.7% decomposition in the same reaction time<sup>[1]</sup>. The rate of decomposition for cyanide can be calculated with equation 1<sup>[1]</sup>, where the values for each variable can be found in Table 1<sup>[1]</sup>.

Under the same pH conditions, 8.0 MPa of oxygen, and higher temperatures, the rate of decomposition increased with the increase in temperature. The formation of  $\text{CO}_2$  and  $\text{N}_2$  was preferred with lower temperatures, while ammonia and formate were the dominant products with higher temperatures. At a temperature of 373 K, it required 315 minutes to achieve a conversion of 93%, while a temperature of 473 K required only 2 minutes to

reach the same conversion <sup>[1]</sup>. The rate for cyanide decomposition may also be calculated using equation 1, and the values from Table 1 <sup>[1]</sup>, where k is the rate of reaction and C is concentration of the given compound.

$$r_{CN^-} = -\frac{dC_{CN^-}}{dt} = -(k_1 + k_2)C_{CN^-} = -K_{ov}C_{CN^-} \quad (\text{Eq. 1})$$

Unlike cyanide, thiosulfate has one pathway for the decomposition, as found in the stoichiometric equation 2. It was noted that the logarithmic plot of k' against oxygen pressure was linear with a slope of 1.66 <sup>[2]</sup>. When calculating the rate of reaction for thiosulfate, equation 3 may be used where it is a first order reaction inside a pressurized vessel, similar to an autoclave <sup>[2]</sup>.



$$-\frac{dC_{S_2O_3^{2-}}}{dt} = kC_{S_2O_3^{2-}}C_{OH^-}(P_{O_2})^{1.66} \quad (\text{Eq. 3})$$

By increasing the temperature of the reaction vessel, the rate of decomposition was found to increase. The reaction rates for the different temperatures varied from  $3.33 \times 10^6 \text{ M s}^{-1}$  to  $21.6 \times 10^6 \text{ M s}^{-1}$  for 373 K and 411 K, respectively <sup>[2]</sup>.

Table 1. Relevant Kinetic Data Results for the Degradation of Cyanide at Different Operating Conditions<sup>[1]</sup>

| effect of oxygen pressure                    |                                  |                                  |                                  |                                  |                                  |
|--|----------------------------------|----------------------------------|----------------------------------|----------------------------------|----------------------------------|
| T (K)  | 453                              |                                  |                                  |                                  |                                  |
| P (MPa)                                      | 2.0 (O <sub>2</sub> )            | 5.0(O <sub>2</sub> )             | 8.0 (O <sub>2</sub> )            | 8.0 (N <sub>2</sub> )            |                                  |
| C <sub>CN<sub>0</sub><sup>-</sup> (mM)</sub> | 3.85 ± 0.01                      | 3.85 ± 0.02                      | 3.85 ± 0.03                      | 4.40 ± 0.03                      |                                  |
| C <sub>O<sub>2</sub></sub> (M)               | 1.0 × 10 <sup>-2</sup>           | 3.8 × 10 <sup>-2</sup>           | 6.7 × 10 <sup>-2</sup>           |                                  |                                  |
| k <sub>ov</sub> (s <sup>-1</sup> )           | (5.98 ± 0.03) × 10 <sup>-3</sup> | (6.27 ± 0.07) × 10 <sup>-3</sup> | (6.42 ± 0.07) × 10 <sup>-3</sup> | (6.32 ± 0.03) × 10 <sup>-3</sup> |                                  |
| r <sup>2a</sup>                              | 0.9991                           | 0.997                            | 0.997                            | 0.994                            |                                  |
| k <sub>1</sub> (s <sup>-1</sup> )            | (2.27 ± 0.06) × 10 <sup>-3</sup> | (2.35 ± 0.16) × 10 <sup>-3</sup> | (2.40 ± 0.16) × 10 <sup>-3</sup> | (2.10 ± 0.07) × 10 <sup>-3</sup> |                                  |
| k <sub>2</sub> (s <sup>-1</sup> )            | (3.71 ± 0.03) × 10 <sup>-3</sup> | (3.92 ± 0.09) × 10 <sup>-3</sup> | (4.02 ± 0.09) × 10 <sup>-3</sup> | (4.22 ± 0.04) × 10 <sup>-3</sup> |                                  |
| φ  | 0.620 ± 0.002                    | 0.626 ± 0.003                    | 0.626 ± 0.006                    | 0.667 ± 0.002                    |                                  |
| effect of temperature                        |                                  |                                  |                                  |                                  |                                  |
| P (MPa)                                      | 8.0                              |                                  |                                  |                                  |                                  |
| T (K)  | 393                              | 408                              | 423                              | 453                              | 473                              |
| C <sub>CN<sub>0</sub><sup>-</sup> (mM)</sub> | 3.85 ± 0.01                      | 3.85 ± 0.02                      | 3.85 ± 0.03                      | 3.85 ± 0.03                      | 3.85 ± 0.02                      |
| k <sub>ov</sub> (s <sup>-1</sup> )           | (1.35 ± 0.01) × 10 <sup>-4</sup> | (4.53 ± 0.05) × 10 <sup>-4</sup> | (1.04 ± 0.02) × 10 <sup>-3</sup> | (6.42 ± 0.07) × 10 <sup>-3</sup> | (2.02 ± 0.02) × 10 <sup>-2</sup> |
| r <sup>2a</sup>                              | 0.997                            | 0.997                            | 0.9992                           | 0.997                            | 0.998                            |
| k <sub>1</sub> (s <sup>-1</sup> )            | (8.57 ± 0.02) × 10 <sup>-5</sup> | (2.30 ± 0.08) × 10 <sup>-4</sup> | (4.37 ± 0.03) × 10 <sup>-4</sup> | (2.40 ± 0.16) × 10 <sup>-3</sup> | (4.50 ± 0.5) × 10 <sup>-3</sup>  |
| k <sub>2</sub> (s <sup>-1</sup> )            | (4.95 ± 0.01) × 10 <sup>-5</sup> | (2.23 ± 0.03) × 10 <sup>-4</sup> | (5.98 ± 0.01) × 10 <sup>-4</sup> | (4.02 ± 0.09) × 10 <sup>-3</sup> | (1.57 ± 0.03) × 10 <sup>-2</sup> |
| φ  | 0.366 ± 0.002                    | 0.491 ± 0.002                    | 0.578 ± 0.003                    | 0.626 ± 0.006                    | 0.778 ± 0.004                    |
| effect of initial cyanide concentration      |                                  |                                  |                                  |                                  |                                  |
| P (MPa)                                      | 8.0                              |                                  |                                  |                                  |                                  |
| T (K)  | 453                              |                                  |                                  |                                  |                                  |
| C <sub>CN<sub>0</sub><sup>-</sup> (mM)</sub> | 3.85 ± 0.03                      | 5.77 ± 0.02                      | 7.69 ± 0.02                      | 9.62 ± 0.03                      | 25 ± 0.02                        |
| k <sub>ov</sub> (s <sup>-1</sup> )           | (6.42 ± 0.07) × 10 <sup>-3</sup> | (6.55 ± 0.10) × 10 <sup>-3</sup> | (6.62 ± 0.07) × 10 <sup>-3</sup> | (6.68 ± 0.07) × 10 <sup>-3</sup> | (6.90 ± 0.03) × 10 <sup>-3</sup> |
| r <sup>2a</sup>                              | 0.997                            | 0.9991                           | 0.998                            | 0.997                            | 0.998                            |
| k <sub>1</sub> (s <sup>-1</sup> )            | (2.40 ± 0.16) × 10 <sup>-3</sup> | (2.42 ± 0.18) × 10 <sup>-3</sup> | (2.44 ± 0.13) × 10 <sup>-3</sup> | (2.44 ± 0.14) × 10 <sup>-3</sup> | (2.52 ± 0.06) × 10 <sup>-3</sup> |
| k <sub>2</sub> (s <sup>-1</sup> )            | (4.02 ± 0.09) × 10 <sup>-3</sup> | (4.13 ± 0.08) × 10 <sup>-3</sup> | (4.18 ± 0.06) × 10 <sup>-3</sup> | (4.24 ± 0.07) × 10 <sup>-3</sup> | (4.38 ± 0.03) × 10 <sup>-3</sup> |
| φ  | 0.626 ± 0.006                    | 0.630 ± 0.003                    | 0.632 ± 0.003                    | 0.634 ± 0.004                    | 0.635 ± 0.001                    |

<sup>a</sup>Referred to the goodness of the k<sub>ov</sub> fitting.

## Experimentation

### CIL Experimentation

In order to get an idea about how the contamination of the CIL circuit with thiosulfate ions would affect overall gold yield, the design team conducted a set of comparison experiments. The parameters of the experiments are listed in Table 2. These experiments were performed using the bottle roll method of leaching gold and adsorbing onto an activated carbon matrix in the form of 8x16 mesh coconut charcoal giving a minimum

surface area of 1200 m<sup>2</sup>/g [3]. Initially, gold wire was used to provide a source of gold to be dissolved by the 0.5% cyanide solution; however, it was determined that the dissolution rate of the gold from the wire was too low (due to low surface area of the wire) in order to perform the experiments in sufficient time. Therefore, a gold potassium cyanide complex was used as the precursor to get gold into solution efficiently. Despite the use of a precursor for gold dissolution, all of the parameters in Table 2 were maintained.

The experimental procedure for gold leaching in cyanide starts with the preparation of the cyanide solution. A 500 mL volume of deionized water is brought to a pH of 11 with sodium hydroxide in order to prevent the formation and volatilization of hydrogen cyanide gas. Next, 0.5% sodium cyanide is added to the solution and stirred until dissolved. Finally, the gold potassium cyanide, K[Au(CN)<sub>2</sub>], is added at 760 micromolar concentration in gold ions. The solution is stirred for 15 minutes to allow completely dissolution. For the comparison bottle, the procedure is the same as above with the addition of 50 ppm of sodium thiosulfate. After complete dissolution, a sample of the solution is taken in order to obtain a baseline gold loading measurement. Finally, the activated carbon is added at the minimum loading of 1 pound of carbon to 0.1 ounces of gold present in solution<sup>[3]</sup>. With the addition of carbon, the experiments are rolled on a bottle roller for 16 hours in order to achieve maximum adsorption.

Table 2: Reagent Parameters for CIL experiments

|                           |
|---------------------------|
| Experimental Parameters   |
| 0.5% Sodium Cyanide       |
| 50 ppm Sodium Thiosulfate |
| 25 g Activated Carbon     |
| ~760 $\mu$ M Gold         |

Following rolling, the solutions were stirred to maintain uniform mixture of gold in the solution in order to achieve a more representative sample of gold from the solution. A 7 milliliter sample of each solution was taken to be compared to the baseline sample taken previously. The solutions were analyzed using an inductively coupled plasma optical emission spectrometer (ICP-OES). Using a 10,000 K plasma, the ICP atomizes and “burns” the solution measuring the optical emission of the individual elements in the solution due to relaxation of excited states unique to each element. The results of the analysis are displayed in Table 3. The difference in adsorption of the gold complexes onto the carbon as indicated by the amount of gold left in solution is shown to be negligible or within the error of the ICP-OES detection limits at 0.0767 ppb or a 0.0077% difference. This means that while the ICP-OES is capable of measuring gold at ppb levels the accuracy of the system diminishes significantly as concentration decreases below ppm levels. Therefore, it can be concluded that the addition of 50 ppm thiosulfate, as can be expected from the recycled tailings, will have little effect on the adsorption of gold onto carbon.

Table 3: ICP-OES results for both samples

| Sample #1: 0.5% CN with 50 ppm NaTS |                 |             |        |
|-------------------------------------|-----------------|-------------|--------|
| Sample                              | Wavelength (nm) | Conc (ppm)  | RSD    |
| Baseline #1                         | Au 267.595      | 147.8032696 | 1.054% |
| After Charcoal #1                   | Au 267.595      | 0.154267109 | 0.163% |
| Percentage Not Adsorbed onto Carbon |                 | 0.1044%     |        |
|                                     |                 |             |        |
| Sample #2: 0.5% CN with no NaTS     |                 |             |        |
| Sample                              | Wavelength (nm) | Conc (ppm)  | RSD    |
| Baseline #2                         | Au 267.595      | 152.5577818 | 0.536% |
| After Charcoal #2                   | Au 267.595      | 0.147532586 | 0.596% |
| Percentage Not Adsorbed onto Carbon |                 | 0.0967%     |        |
| Difference                          | Percentage      | 0.0077%     |        |
|                                     | Conc (ppm)      | 7.66723E-05 |        |

## RIL Experimentation

A lab scale experiment was designed to test the effects of cyanide contamination in the RIL process. The results of this experiment were necessary to determine if trace amounts of cyanide would have a negative effect on the RIL process and if so, the amount of gold lost in comparison to an uncontaminated RIL test would give us a base case for an incremental economic analysis of different purification processes. The following procedure was used to test an uncontaminated RIL process as well as two different cyanide contamination levels.

The resin used in this process is very expensive so it is important to use only the amount needed to leach all the gold in solution. A 500:1 volume ratio of our gold thiosulfate solution to resin was determined to be the necessary amount, however, the resin is solid spherical particles so measuring volumes is not simple. To make following measurements easier, the density of our resin was determined experimentally using water displacement

and a very accurate gram scale. With all the necessary physical properties known, the lab solutions were prepared. A 100 mL solution of gold thiosulfate was prepared using .1191g of gold thiosulfate powder in 100mL of water. For safety reasons, this solution was then made to be very basic using .42g of NaOH. This solution was set aside as our uncontaminated test and two more solutions were made using .2404g gold thiosulfate in 200mL of water and made basic using .82g NaOH. This 200mL solution was split into two 100mL solutions and contaminated with .0021g of cyanide and .0038g of cyanide respectively. Note, all cyanide was handled under a hood. With all three of these solutions prepared, the resin for each solution was measured. Using a 500:1 volume ratio, .2578g of resin was added to the uncontaminated solution, .2584g of resin was added to the lower cyanide contamination, and .2580g of resin was added to the higher cyanide contamination. All three of these vessels were then sealed and placed on a shake table at 120 RPM to continuously agitate them.

Samples of the solutions were taken prior to adding the resin to measure the amount of gold in solution before the leaching process began. Knowing that this leaching process is a slow one, the solutions were allowed three days to reach equilibrium and samples were taken from each solution periodically to measure the progress of each reaction. These samples were all measured using Inductively Coupled Plasma Mass Spectrometry (ICP-MS).

While carrying out these experiments, safety was a huge concern. Cyanide is a very dangerous chemical, so extra precautions were taken to assure that there was no harmful exposure. Long pants and sleeves were used to avoid skin contact as well as gloves and goggles. This basic PPE is necessary for any lab, however, cyanide is also volatile in

solutions with pH less than 10 and is deadly if inhaled. To avoid any possible chance of cyanide inhalation, our solutions were made to be above pH 10 and all solutions containing cyanide were always handled underneath a hood. To assure that no trace amounts of cyanide were remaining on our hands and later ingested from food contact, everyone thoroughly washed their hands before leaving the lab. With these safety consideration taken into account, the experiments were carried out with no worry of danger.

After sufficient time intervals had passed, samples were analyzed using an ICP-Optical Emission Spectrometer in order to quantify the amount of gold absorbed into the resin. After the resin has absorbed the gold in solution, the resin should be flushed to quantify the amount of gold absorbed by the resin. The following table shows the results from the first half of the experimentation discussed above.

Table 4: ICP-OES Results for CN contamination in RIL simulated experiment

| Sample                 | 1      | 2      | 3      |
|------------------------|--------|--------|--------|
| CN Contamination (ppm) | 0.00   | 21.00  | 38.00  |
| Au Initial (ppm)       | 415.93 | 164.71 | 164.71 |
| Au Intermediate (ppm)  | 239.72 | 103.29 | 74.51  |
| Au Final (ppm)         | 237.82 | 102.95 | 69.95  |
| Fraction Au Absorbed   | 0.43   | 0.36   | 0.58   |

Some conclusions can be drawn by the data in the uncontaminated sample, and the sample with a lower amount of contamination. It appears that the data for the sample with 38.00 ppm of Cyanide contamination had errors, or the procedure in preparing the sample was incorrect. One significant source of error likely occurred in a large difference in pH

between the 21 and 38 ppm cyanide contaminated samples. Due to the importance of safety when working with cyanide, higher amounts of NaOH was added to the 38 ppm CN sample, which provided a larger amount of free cyanide ions to leach the gold. When looking at the first and second sample however it is apparent that cyanide contamination in the resin in leach circuit produces negative effects on gold absorption into the resin. In order to effectively quantify the repercussions of tailing stream mixing, further empirical data should be recorded. The following suggestions should be implemented in order to achieve better results for RIL experimentation:

1. The pH of contaminated samples should be constant at a safe level
2. More samples should be taken at different time intervals in order to get a better understanding of where steady state occurs
3. A single stock solution with the same amount of initial gold concentration should be divided and contaminated with more variations of cyanide concentrations

## **Alternate Cases**

### **Thiosulfate Treatment with Ion Exchange**

Ion exchange is one viable option for the removal of Thiosulfate (TS). Ion exchange uses a column packed with resin which absorbs ions from the feed solution. Charge neutrality is maintained because the resin is initially filled with different ions which are desorbed. The resin has a specific capacity for ion absorption, and once the resin is full, the resin must either be back flushed or removed and flushed, known as regeneration. *Bio-Rad* anion exchange resin AG1-X8 with a 1.2 meq/ml capacity has been chosen to treat

thiosulfate because of its high selectivity in respect to the ion thiosulfate, as shown in the Figure 2 below <sup>[4]</sup>:

| Ion   | Selectivity with respect to OH <sup>-</sup> |        |
|---|---|--------|
|   | AG1-X8                                      | AG2-X8 |
| Cl <sup>-</sup>   | 22  | 2.3    |
| SO <sub>4</sub> <sup>2-</sup> , HSO <sub>4</sub> <sup>-</sup> | 85  | 15     |
| S <sub>2</sub> O <sub>3</sub> <sup>2-</sup>                   | ≈275  | ≈90    |
| S <sub>x</sub> O <sub>6</sub> <sup>2-</sup> (n > 2)           | Fully retained                              | ≈500   |

Figure 2: Selectivity of the ion exchange resin used in this study<sup>[4]</sup>

This resin presents dry mesh size of 50-100 mesh, wet bed diameter of 180-425 μm and nominal density of 0.75 g/ml <sup>[5]</sup>. Based on previous experiments performed at Yellowstone National Park and Green Lake, NY, the resin has been shown to be very effective in the extraction of thiosulfate ion out of water streams. When sampling solutions containing ion thiosulfate passed through the resin columns and eluted with KCl 1-3 M, recoveries of 89 to 100% were achieved <sup>[4]</sup>.

Based on previous experiments, precautions should be taken to avoid exceeding the capacity of the resin, otherwise incomplete capture of the ions causing poor efficiency will result <sup>[4]</sup>. The figure below, taken from Bio-Rad's instruction manual, shows the recommended flow rate for ion exchange resins <sup>[5]</sup>.

| <b>Application</b>                                    | <b>Linear Flow Rate (cm/min)</b> |
|---|----------------------------------|
| Removing trace ions                                   | 5-10                             |
| Separations with very few components                  | 1-3                              |
| Separations of multi-component samples                | 0.3-1.0                          |
| Using high resolution resins with small particle size | 1-2                              |

Figure 3: Suggested flow rate for ion exchange resins<sup>[5]</sup>

Based on the above table, for laboratory scale, the indicated flow rate for this resin is between 5-10 cm/min in order to improve sample ion and resin interaction. Also was found that the higher the flow rate, the less ions will be retained, due to short contact time.

### Ion Exchange Design Calculations

The following demonstrates design calculations required for assessing required resin volume and resin effectiveness.

The resin volume can be calculated as follows <sup>[6]</sup>:

$$\text{Resin Volume} = \frac{\text{Feed salinity} \left( \frac{\text{eq}}{\text{m}^3} \right) \times \text{Throughput} (\text{m}^3)}{\text{Resin operating capacity} \left( \frac{\text{eq}}{\text{m}^3} \right)}$$

Using the following data,

$$\text{Flow rate} = 1500 \text{ gpm} = 340.69 \text{ m}^3/\text{h}$$

$$\text{Feed salinity} = 50 \text{ ppm} = 0.0657 \text{ eq}/\text{m}^3$$

$$\text{For one day of operation: Throughput} = 340.69 \left( \frac{\text{m}^3}{\text{h}} \right) \cdot 24 (\text{h}) = 8.175 * 10^3 \text{ m}^3$$

The flow rate of water to be treated is equal to:

$$\text{Resin Volume} = \frac{0.0657 \left( \frac{\text{eq}}{\text{m}^3} \right) \cdot 8.175 \cdot 10^3 (\text{m}^3)}{1200 \left( \frac{\text{eq}}{\text{m}^3} \right)} = 0.447 \text{ m}^3$$

$$\text{Resin density} = 750 * 10^3 \text{ g}/\text{m}^3$$

$$\text{Resin mass} = 750 * 10^3 \times 0.447 = 335.25 * 10^3 \text{ g}$$

$$\text{Removal capacity} = \frac{0.09132 \text{g}(\text{CaS}_2\text{O}_3)}{\text{mL}(\text{resin})} = \frac{91320 \text{g}(\text{CaS}_2\text{O}_3)}{\text{m}^3(\text{resin})}$$

The removal capacity is calculated as follows:

$$\text{Exchange capacity} = \text{Removal capacity} \left( \frac{\text{g}}{\text{m}^3} \right) \cdot \text{Resin Volume} (\text{m}^3)$$

$$\text{Exchange capacity} = 91320 \left( \frac{g}{m^3} \right) \cdot 0.447 (m^3) = 40.82 \cdot 10^3 g (CaS_2O_3)$$

The water treatment capacity is calculated as follows:

$$\text{Hardness} = 0.05 \frac{g}{L} = 50 \frac{g}{m^3}$$

$$\text{Water treatment capacity} = \frac{\text{Exchange capacity}(g)}{\text{Hardness} \left( \frac{g}{m^3} \right)} = \frac{40.82 \cdot 10^3 g}{50 \left( \frac{g}{m^3} \right)} = 816.4 m^3$$

Then, using the previous results, we can calculate the operating time as follows:

$$\text{Operating time} = \frac{\text{Water treatment capacity}(m^3)}{\text{Flow rate}(m^3/h)}$$

$$\text{Operating time} = \frac{816.4}{340.69} = 2.4 h$$

The operation time should be 24 h, therefore we can achieve this time with 10 times more resin. Then:

$$\text{Resin Volume} = 4.47 m^3$$

$$\text{Operating time (h)} = 24 h$$

### **Ion Exchange Vessel Design**

The volume of the vessel must allow all the resin sites to be contacted by the fluid. For this reason, it is reasonable that the vessel volume should be about two times bigger than the resin volume. The calculations are as follows:

$$\text{Vessel volume} = 2.3 \cdot \text{Resin volume} = 10.28 m^3$$

For a vessel, the L/D ratio is usually 2.5 <sup>[7]</sup>. With this ratio, we can calculate the vessel diameter as follows:

$$Diameter = \left[ \frac{4 \cdot Vessel\ volume}{\pi \left(\frac{L}{D}\right)} \right]^3 = \left[ \frac{4 \cdot 10.28}{2.5\pi} \right]^3 = 1.74\ m$$

Supposing a dead height of 10%, the height can be calculated as follows:

$$Height = Diameter * \frac{L}{D} * 1.1 = 1.74 * 2.5 * 1.1 = 4.78\ m$$

### Ion Exchange Vessel Cost

In order to achieve a continuous system, two vessels are necessary. The cost of these vessels was calculated using the CAPCOST program provided by Turton (2012). The results are shown in Table 5.

Table 5: Vessel design parameters and cost

| Quantity | Height (meters) | Diameter (meters) | Vessel MOC      | Pressure (barg) | Purchased Equipment Cost | Bare Module Cost |
|----------|-----------------|-------------------|-----------------|-----------------|--------------------------|------------------|
| 2        | 2.22            | 0.81              | Stainless Steel | 2               | \$ 16,460.00             | \$84,200.00      |

### Ion Exchange Operating Cost

The operating cost is basically the regeneration cost associated with the use of KCl which is usually priced at 2.44 \$/(mol of solid). Considering that the cost of water is negligible, the price of the KCl 1 molar solution is 2.44\$/(mol/L). The volume of KCl needed to regenerate the column is assumed to be 1.2 times the resin volume.

With the presented data, we can calculate the operating cost as follows:

$$\text{Volume needed per batch} = 1.2 * \text{resin volume} = 1.2 * 4.47 = 5.36 \text{ m}^3/\text{day}$$

$$\text{Volume needed per year} = 5.36 * 365 = 1957.86 \frac{\text{m}^3}{\text{year}}$$

$$\text{Cost of KCl per year} = \text{Price of KCl per m}^3 * \text{Volume needed per year}$$

$$\text{Cost of KCl per year} = 2436 * 1957.86 = \$ 4,769,346.96$$

### Possible gold loss

In 2014, according to information found in the Barrack official website [8], the gold production in the Gold Strike mine was 900,000 ounces. Supposing a gold loss of 2% of gold, this represents 15,300 ounces. As the gold is priced at about 1204.6 \$/ounce, a loss of 2% could represent 16,798,523.44 \$/year.

Non-discounted Cash Flow:

Using the operating cost, the possible loss of gold as a profit and the resin cost, a non-discounted cash flow diagram was made which is showed in the Figure 4. The payback period is about 1 year.

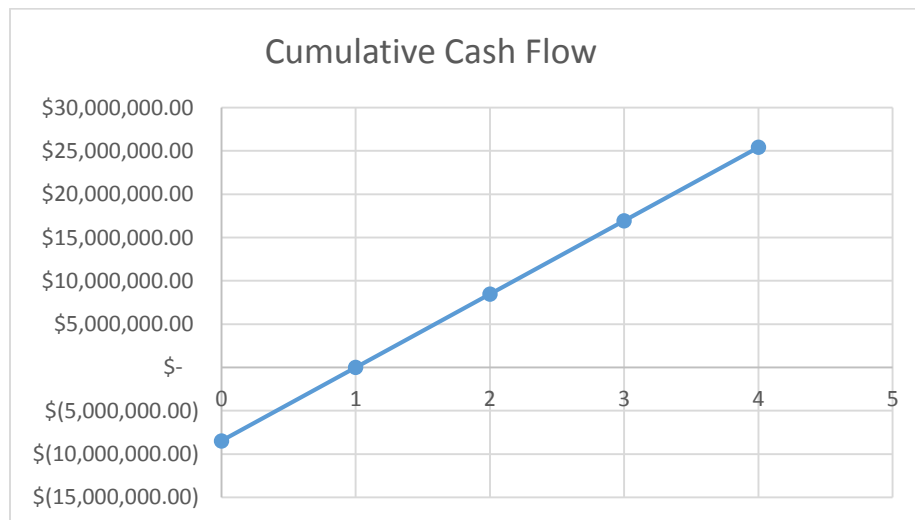


Figure 4: Non-discounted cash flow diagram

## TS Treatment with Hydrogen Peroxide

An alternative case for the treatment of thiosulfate in solution is found in hydrogen peroxide. Thiosulfate reacts quantitatively with hydrogen peroxide in an alkaline solution. The chemical equation of the hydrogen peroxide-thiosulfate reaction is given below<sup>[9]</sup>.



It was found by FMC chemicals that when the reaction was allowed to run for 10 minutes, the reaction went to 90% completion, and after 30 minutes the reaction had reached 98% completion. Fig 5 shows results from a typical experiment when running the system at a pH of greater than 8<sup>[9]</sup>.

| Sodium Thiosulfate<br>Pentahydrate Conc.<br>mg/l | H <sub>2</sub> O <sub>2</sub> Conc.<br>mg/l | BOD <sub>5</sub><br>mg/l             |                                     |
|--|---|--------------------------------------|-------------------------------------|
|  |   | Before H <sub>2</sub> O <sub>2</sub> | After H <sub>2</sub> O <sub>2</sub> |
| 20,000   | 11,000                                      | 4,000                                | 40                                  |

Figure 5: Typical results of a thiosulfate-hydrogen peroxide reaction run at a pH of 8

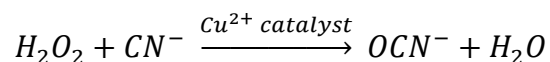
To find a cost estimate of using hydrogen peroxide, a purchasing cost of \$1.50/kg for hydrogen peroxide, a flowrate of 340 m<sup>3</sup>/hr, and a concentration of 50 ppm thiosulfate are assumed. The cost per hour of hydrogen peroxide is found to be approximately \$14, bringing the cost of hydrogen peroxide to be \$120,000 per year<sup>[9]</sup>. Further economic analysis was not pursued because Barrick informed the team that TS does not need to be

treated because tailings water will never be recycled to the CIL circuit due to corrosion concerns.

## **CN Treatment with Hydrogen Peroxide**

Within mining industries, cyanide treatment of the process flow can be reduced by using hydrogen peroxide ( $H_2O_2$ ) to oxidize the cyanide ions ( $CN^-$ ) into cyanate ions ( $OCN^-$ ).

This process is carried out following the reaction given below: <sup>[10]</sup>



By turning cyanide into a less reactive substance, cyanate, the resulting slurry or solution can be recycled for other uses or legally released into the sewage. In certain mining applications, turning  $CN^-$  into  $OCN^-$  limits the build-up of dissolved solids in the solution, thus effectively reducing the amount of reaction between  $CN^-$  and gold ores as well as with thiosulfate. This process would reduce the amount of thiocyanate formed in the solution, which is a harmful substance to have in the industry. The reduction in rate of binding to gold ores also reduces the interference from  $CN^-$  in producing gold through the CIL process and effectively reduces the amount of gold lost from combining the CIL and RIL processes. <sup>[11]</sup>

In a lab testing experiment, the effective usage of  $H_2O_2$  is about 1.3 grams of  $H_2O_2$  to oxidize about 1 gram of  $CN^-$ . However, the effective usage range in a slurry is about 2.0 to 8.0 grams of  $H_2O_2$  per gram of  $CN^-$ . The reaction required copper ions as a catalyst, however, no study which suggests the definitive amount for the best rate of conversion can be found.

Whether or not the destruction of cyanide is necessary for the process will be determined by the profit obtained when the destruction by  $H_2O_2$  is implemented. For Barrick, largest problem with cyanide is the interaction between cyanide and gold ores as the process slurry goes through the autoclave. The following table shows the data obtained from Barrick during an accidental recycle of cyanide without treatment to the autoclave circuit.

Table 6: Process flow data through Autoclave <sup>[12]</sup>

|                             |                   |
|-----------------------------|-------------------|
| Flowrate                    | 4000 gal/min      |
| Gold ores                   | 12 to 14 tons/day |
| Cyanide                     | 5 ppm             |
|                             |                   |
| Reduced gold leached by 1/3 |                   |

From the data given, it is suggested that the amount of gold extracted from the ores is reduced by about 33% from the highest extraction amount possible. This means that the revenue gains per year from obtaining 33% more gold has to be higher than the cost of materials and labor of oxidation by  $H_2O_2$  to be profitable.

For the profitability analysis, the average extractable gold per ton of ores is about 7.12 kg of gold per 1000 kg of gold ores <sup>[12]</sup>. Considering the 33% loss, the amount of gold generated with cyanide in the slurry is about 4.3 kg. With these data combined with the average gold price for the last 5 years, the revenue generated from various gold amount produced is:

Table 7: Revenue from Gold Yield

| Gold yield (per year) | revenues (\$ million/year) |
|-----------------------|----------------------------|
| 100% (36385 kg)       | 1.519                      |
| 66% (24000 kg)        | 1.002                      |

The data from Table 7 suggests that the loss of 33% of the maximum gold yield means the loss of about \$500 million in revenue.

A calculation has been done to estimate the amount of cyanide going through the autoclave. With 12 to 14 tons of ores per day, there are approximately 60 to 70 grams of cyanide in the slurry per day. The profitability calculation is to determine the highest theoretical cost for implementing and operating the oxidation process based on using either 2g of H<sub>2</sub>O<sub>2</sub> or 8g of H<sub>2</sub>O<sub>2</sub>. The cost of 100% H<sub>2</sub>O<sub>2</sub> is estimated to be about \$1.50 per kg. The amount needed, as well as price calculation for H<sub>2</sub>O<sub>2</sub> is shown in the table below: <sup>[444]</sup>

Table 8: Material cost per year

| For 70 grams CN- per day                                      |         |         |
|---|---------|---------|
| \$60 per gallon of 70% H <sub>2</sub> O <sub>2</sub> (4.9 kg) |         |         |
|   | 2 grams | 8 grams |
| H <sub>2</sub> O <sub>2</sub> required in                     |         |         |
| per day   | 140.00  | 560     |
| per month   | 4200    | 16800   |
| per year  | 51100   | 204400  |
| Price per year (\$)   | 56224   | 224896  |

For the economic analysis, the sheer value of gold obtained alone by eliminating cyanide from the autoclave alone make it worth to install and operate the cyanide destruction process. Depending on the storage unit as well as the piping and other installation and maintenance fee, the capital cost could be the theoretical value of revenue gains from more gold yield produce more than enough to settle the capital cost and greatly improve the profit margin. The following table shows the capital cost, including the rough estimate of the storage tanks, piping, and other modifications:

Table 9: Total cost of implementation

| Cost items          | Cost (\$)     |
|---------------------|---------------|
| Equipment           | 164000        |
| Piping/modification | 143000        |
| Miscellaneous       | 60000         |
| Maintenance         | 2310          |
| Process Control     | 8640          |
| Material, H2O2      | 204400        |
| <b>Total Cost</b>   | <b>582350</b> |

For the total cost of \$582,350 in which able to increase the revenue of gold to \$500 million, it may be wise to implement the hydrogen peroxide storage and process to oxidize cyanide in the slurry.

## Reverse Osmosis

The process of Reverse Osmosis utilizes a non-permeable or semi-permeable membrane to “filter” materials. The system runs at high pressures to overcome osmotic pressure and move the permeate to a location of lower salt concentration. For this design project, the removal of thiosulfate from waste water was analyzed.

The membrane used in this design was a thin film composite membrane <sup>[13]</sup>. The rejection level of the membrane is 96-98%, although for this process a rejection level of 98% was assumed. Since reverse osmosis runs at high pressures a pressure of 600 psi was assumed and a pressure of 100 psi for the permeate side was assumed. Since the concentration of thiosulfate in the water is low the osmotic pressure is assumed to be negligible. A concentration of 0.000657 mol/m<sup>3</sup> was assumed to be a safe estimate for the desired permeate concentration, which is 2% of the feed concentration of 5 ppm. It was assumed that the desired permeate flow rate was assumed to be 2/3 the feed flow rate of 4000 gpm.

Assuming a transmembrane flux of 25 gal/ft<sup>2</sup>d, the necessary membrane area needed is 153,600 ft<sup>2</sup>. The calculations for these values can be seen below. These calculations only take into account a once through system and does not consider further processing of the water.

$$R = 1 - \left(\frac{C_p}{C_f}\right) = 98\% \quad (1)$$

$$C_p = 0.02 * C_f = 0.000657 \frac{mol}{m^3} = 1ppm \quad (2)$$

$$4000 \text{ gpm} * \frac{2}{3} = 2666.67 \text{ gpm} \quad (3)$$

$$M_A = \frac{Q_P}{J_V} = \frac{2666.67 \text{ gpm}}{25 \text{ gal/ft}^2 \text{d}} = 153,600 \text{ ft}^2 \quad (4)$$

A FILMTEC RO unit designed by DOW Chemicals has roughly 380 ft<sup>2</sup> of membrane area and will be used to determine the necessary number of units to achieve the required membrane area.

$$\frac{153,600 \text{ ft}^2}{380 \text{ ft}^2/\text{unit}} = 405 \text{ units} \quad (5)$$

This RO unit cost roughly \$300.00. It is assumed that the capital cost required per unit is \$10,000 per unit. At 405 units the total capital cost is \$4,050,000.00. With little information on operating costs an operating and maintenance cost will be \$100.00 per unit annual amounting to \$40,500.00.

### **SPARRO Reverse Osmosis**

There is another type of reverse osmosis that has been used in mining industries before. This process is called the Slurry Precipitation and Recycle Reverse Osmosis (SPARRO)

that was done in a mine South Africa. This process takes the precipitation of metals by increasing the effluent pH as a pretreatment step. This is followed by cooling the slurry, then filtering, and finally readjusting the pH to the desired amount for the process. This was able to recover 96% water and reducing the sulfate salts from about 6600mg/L to 150g/L. This week I will talk more about this process since I did not completely cover all the design parameters and economics of this process.

The SPARRO process is the development and evaluation of seeded membrane process. The main objected for SPARRO was to overcome serve shortcoming of existing seeded systems. These include a high energy consumption and high slurry recirculation of water. There are several design parameters for this process which was address to allow for the best treatment.

- Membrane Stack and module configuration
  - Tube velocity: At the mine in South Africa, one problem was the high power consumption and this was due to the linear tube velocities and the requirement for inter-stage pumping. Having a certain velocity can effect the particle size of material of removal, so calculating minimum velocity can help save money and the membrane
  - Tapered configuration: It was found that the slurry flow through the membrane stake was diminishing as a result of the permeation of the product water through the membrane walls. To avoid the slurry flow, different

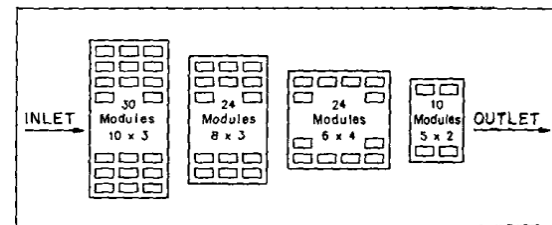


Figure 6: Tapering of membrane modules

size modules, with different configurations were found to improve flow while achieving the same effect. This can be seen in Fig. 6.

- Pressure drop and concentration factor: It was found that having a small pressure drop would be better than a large, which would be more economical for the design. The design values can be seen in Table 10.

Table 10: Design Values for Different Process Variables

| <b>Design values for different process variables</b> |                  |               |                |
|--|------------------|---------------|----------------|
| <b>Process variable</b>                              | <b>Condition</b> |               |                |
|  | <b>Minimum</b>   | <b>Design</b> | <b>Maximum</b> |
| Corrected membrane flux (Um <sup>2</sup> d)          | 400              | 550           | ~600           |
| Inlet pressure (kPa(g))                              | 3,600            | 4,000         | 4,500          |
| Product water recovery (%)                           | 80               | 90            | ~99            |
| Product extraction (conversion) (%)                  | 40               | 50            | unknown        |

- Feed pumping system: It was found that there were two problems with seeded RO systems. This was mainly from slurry recycle systems. One

causes was that the volume of the slurry recycle was too high. This meant excessive power usage and a need for large pumps. And the other cause was the excessive pump wear. The SPARRO process adopted using a novel pumping procedure. Instead of feeding raw mine water into the reactor vessel, a separate pumping system was used, incorporating a standard multistage centrifugal pump. This allowed for a longer pump life and the volume would decrease by half.

- Other parameters for RO that weren't mention in SPARRO
  - Tank volume
  - Surface Area
  - Fouling and scaling of membrane and equipment
  - Flux

The results of the process looked great. The product water recovery was high with an average of 93.3%. The membrane performance was good, but not perfect, because the membrane was not completely stable. That is the salt rejection and membrane flux values could not be maintained simultaneously. Depending on the module design, the salt rejections were different. This can be seen in Figure 7.

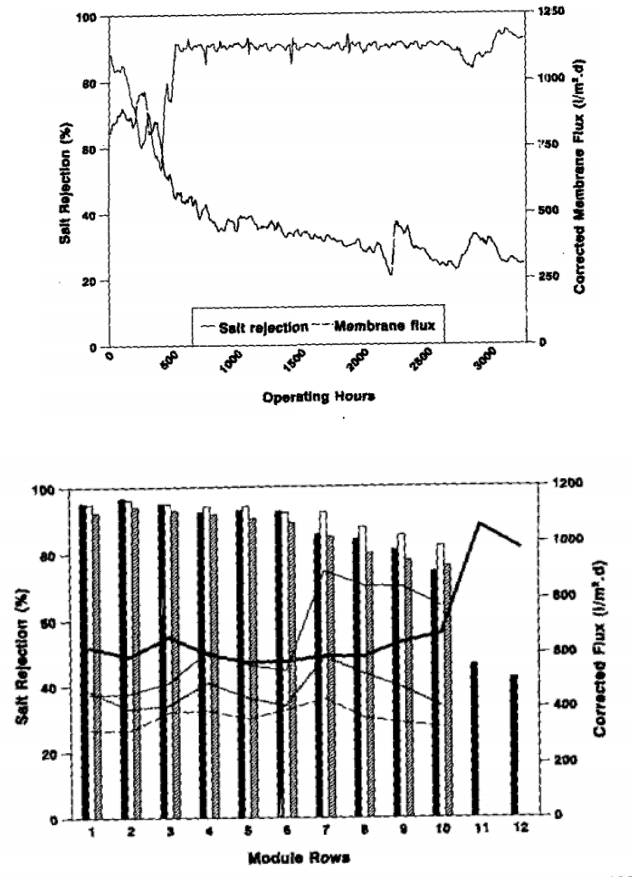


Figure 7: Top - Variation of salt rejection and membrane flux for SPARRO plant. Bottom - Variation of salt rejection and membrane flux for SPARRO for each row of modules.

## Economics

In this paper, it gave some economics that was calculated back in 1994. For part of the economics, the equipment is an important part. Table 11 is what was used for the design parameters of the SPARRO plant vessel. Within this vessel, there could be different variables that could be calculated and accounted for in the economics, which can be seen in Table 12. Along with equipment design, the operating costs, this can be seen in Table

13. After the plant is built, the annual costs and unit cost for capital and operating is important, which can be seen in Table 14.

Table 11: Design Information for SPARRO Plant Reactor Vessel.

| <b>Description</b>                       | <b>Value</b> |
|--|--------------|
| <b>Reactor</b>                           |              |
| Maximum capacity (m <sup>3</sup> )       | 4.5          |
| Design capacity (m <sup>3</sup> )        | 3.5          |
| <b>Dimensions</b>                        |              |
| Diameter (m)                             | 1.5          |
| Height (m)                               | 2.5          |
| <b>Mixer</b>                             |              |
| Power (kW)                               | 0.75         |
| Speed (rpm)                              | 290          |
| Hydraulic Residence time (h) (at design) | 1            |

Table 12: Design values for different process variables

| <b>Design values for different process variables</b> |                  |               |                |
|--|------------------|---------------|----------------|
| <b>Process variable</b>                              | <b>Condition</b> |               |                |
|  | <b>Minimum</b>   | <b>Design</b> | <b>Maximum</b> |
| Corrected membrane flux (Um <sup>2</sup> d)          | 400              | 550           | ~600           |
| Inlet pressure (kPa(g))                              | 3,600            | 4,000         | 4,500          |
| Product water recovery (%)                           | 80               | 90            | ~99            |
| Product extraction (conversion) (%)                  | 40               | 50            | unknown        |

Table 13: Estimated operation cost.

| <b>Estimated operation costs for SPARRO plant installation to produce 46.3 Us of stabilized and disinfected product water (1994)</b> |                                      |
|--|--------------------------------------|
| <b>Description</b>   | <b>Annual cost (x10<sup>3</sup>)</b> |
| <b>Pretreatment</b>  |                                      |
| Alum @ 10 mg/L maximum   | 33                                   |
| Polymer @ 0.5 mg/L   | 10                                   |
| Acid/alkali for pH correction  | 18                                   |
| Chlorine @ 2 mg/L maximum  | 10                                   |
| Electrical power @ 10 c/kWh  | 11                                   |
| General maintenance based on capital cost  | 22                                   |
| <b>SPARRO plant</b>  |                                      |
| Membrane replacement – 2-year life   | 1,404                                |
| Electrical power @ 4.82 kWh/m <sup>3</sup>   | 701                                  |
| Membrane washing   | 20                                   |
| Plunger/packing/valve replacement  | 24                                   |
| General maintenance based on capital cost  | 186                                  |
| <b>Post-treatment</b>  |                                      |
| Lime addition @ 25 mg/L  | 11                                   |
| Chlorine @ 2mg/L maximum   | 10                                   |
| Electrical power @ 10 c/kWh  | 8                                    |
| General maintenance based on capital cost  | 2                                    |
| <b>Staff</b>   |                                      |
| One plant manager – time   | 144                                  |
| Four plant operators – ½ time  |                                      |
| One plant cleaner – full time  |                                      |
| <b>Total estimated operating cost (excluding VAT)</b>  | <b>2,614</b>                         |

Table 14: Summary of total annual costs and unit costs for capital and operating requirements

| <b>Summary of total annual costs and unit costs for capital and operating requirements (1994)</b> |                                      |                                    |
|---|--------------------------------------|------------------------------------|
| <b>Description</b>  | <b>Annual Cost (x10<sup>3</sup>)</b> | <b>Unit Cost (c/m<sup>3</sup>)</b> |
| Interest and redemption on capital  |                                      |                                    |
| Civil (25 years @ 15%)  | 677                                  | 48                                 |
| Mechanical and electrical (15 years @15%)   | 2,076                                | 148                                |
| Estimated operating costs   | 2,614                                | 187                                |
| <b>Total</b>  | <b>5,367</b>                         | <b>383</b>                         |

At a capital cost of \$5 million, the SPARRO system is not the most inexpensive but is a viable alternative.

## **UV Treatment**

UV light can be used to destruct cyanide. The cost of UV destruction is dependent on several factors such as site location, water characteristics, and plant capacity <sup>[14]</sup>. Because the kinetics of cyanide destruction under UV light are relatively unstudied, and factors such as the opacity of the tailings stream, and how that opacity will effect UV light diffusion in the system are unknown, it is difficult to reasonably carry out design calculations and to fully characterize the costs of a UV system. Costs will be based off of an EPA study of the overall effectiveness of UV disinfection of a wastewater stream <sup>[14]</sup>. Table 15 provides an EPA estimate of capital costs of a UV system. This table does not

include the costs of materials transportation, or the cost of building new facilities to house the UV system.

Table 15: Capital Cost of Building a UV destruct system

| <b>Cost item</b>         | <b>UV system Cost (\$)</b> |
|--------------------------|----------------------------|
| Capital Costs            |                            |
| Equipment                | \$183,158                  |
| Structural Modifications | \$97,684                   |
| Electrical               | \$30,526                   |
| Miscellaneous            | \$61,053                   |
| <b>Total:</b>            | <b>\$372,421</b>           |

The most expensive upkeep cost of the system will be replacing the UV lamps. Either low, or medium-pressure lamps will be used for the system. Medium-pressure lamps can cost up to four to five times as much as low-pressure lamps, however, the reduced number of lamps necessary for adequate disinfection could make medium-pressure lamps cost-effective<sup>[14]</sup>. Using flow as a basis for which lamps to purchase, an estimate for lamp cost can be found. Table 16 provides cost comparisons of different lamps based on system flowrates. Overall, UV destruct seems reasonable, but due to the fact that no applications similar to mining have been studied, designing a properly sized system is impossible.

Table 16: Cost Comparison of different UV lamp designs

| Item       | Range     | Typical   |
|------------|-----------|-----------|
| UV         |           |           |
| lamps      | (\$/lamp) | (\$/lamp) |
| 1-5 mgd    | 605-2,083 | 877       |
| 5-10       |           |           |
| mgd        | 523-906   | 724       |
| 19-100 mgd | 418-897   | 610       |

### **Solar Thermal Treatment**

Solar thermal technology converts light from the sun directly into heat that can be used for chemical processes. The sun only showers the earth with so much radiation (about 8 kWh/m<sup>2</sup>/day in Nevada), and due to thermodynamic considerations, sun light can only be converted to heat at a certain efficiency (about 70%). Also, heat energy from the heating medium (in this case molten salt) to the process fluid (tailings water), will be limited by thermodynamic efficiency<sup>[15]</sup>.

There are a number of different designs available, the most studied design being the parabolic trough. The principle of operation for a parabolic trough is that a parabolic mirror reflects sunlight onto an axis running down its middle. In this axis lies a pipe which contains the working fluid, which is usually molten salt. This working fluid can now directly exchange heat with a process fluid.

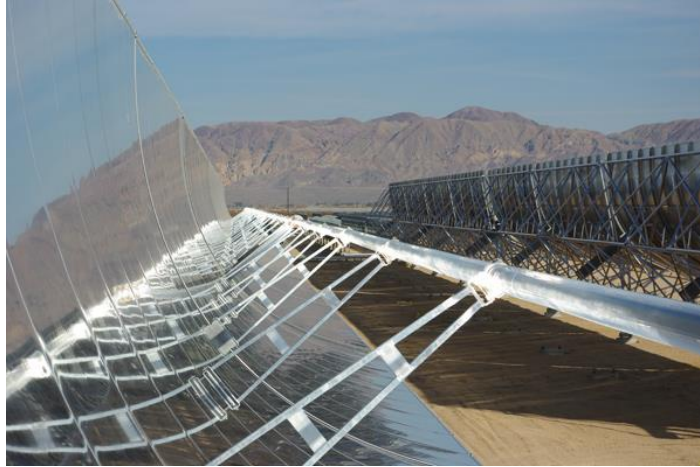


Figure 8. Photograph of a parabolic trough installation.

Estimation of system size is essential for design. For design purposes, we will assume the following design variables:

Table 17. Design variables

| Design Variable  | Value |
|--|-------|
| $\varepsilon$ , overall system efficiency (sunlight into process heat) | 0.35  |
| F, flow rate (gpm)   | 1000  |
| $L_e$ , 'land efficiency' (MW/hectare)                                 | 2.00  |

Using these values, to find the land required to collect enough solar power to distill 1000 gpm of water, the following performance equation is used:

$$L_{req} = \frac{L_e \times F \times \Delta h_{vap} \times \rho}{\varepsilon}$$

Using the above design parameters, this results in a bare minimum 1,987 acres. Our gut tells us that this will not be feasible to design and build. Next we must estimate a cost. A solar thermal power plant near Las Vegas cost about \$2.2 billion and it occupied about 600 acres. This gives a per acre cost as \$3.65 million per acre. Since that was a power generating facility, and our design is only for process heat, we will assume that the facility can be built for half of that, or \$1.825 million per acre. Assuming 1987 acres, that brings the installed cost to \$3.63 billion. This is used for the basis of the economic analysis. Assuming a 3 year construction period, and a cost savings in the form of lower water use, increased gold recovery, and government incentives totaling \$15 million/year. The payback period for this project would be 242 years!

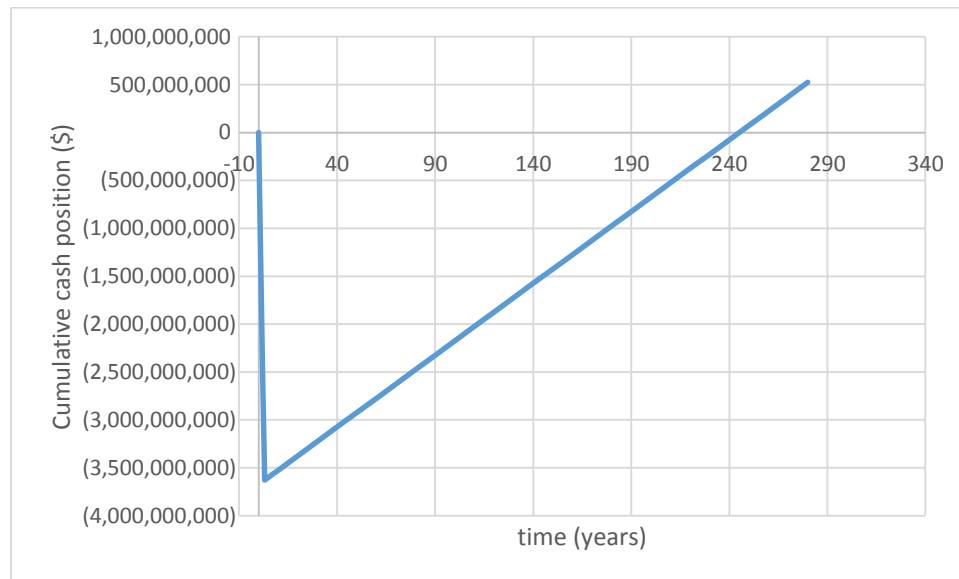


Figure 9: Cumulative cash flow diagram for assumed economics of solar thermal project.

In reality, it seems that this project must be abandoned in favor of other, cheaper options available for the treatment of tailings water. Although in this age of changing climate and

new environmental awareness, it is necessary to investigate options such as solar thermal. As technology improves, or environmental regulation becomes stricter, such options may yet become more viable in the future.

### **Conclusion**

Based on experimentation and feedback from Barrick, the problematic contamination occurs when cyanide enters the RIL circuit even at undetectable levels. Based on Barrick's estimates, as much as 20% gold loss occurs even at undetectable cyanide concentrations. Additionally, the presence of thiosulfate critically accelerates corrosion of carbon steel used in the CIL circuit. Therefore, the only viable alternative treatment options that will meet the needs of Barrick are solar thermal tailings purification and hydrogen peroxide destruct of cyanide. However, currently the solar thermal purification process for this scale of operation is far too expensive to justify. The most economic option would be the use of hydrogen peroxide to destroy cyanide.

### **Future Work**

Feedback from Barrick engineers has led to the conclusion that future senior design classes could explore possible ways of removing thiosulfate from the tailings. The motivation of this is to prevent corrosion of the carbon steel of the roaster's CIL circuit, rather than preventing gold losses. Currently, the most pressing matter for the Goldstrike mining operation is the removal of cyanide contamination in concentrations of < 5ppm which has shown to have detrimental effects on the RIL circuit's gold recovery.

Additionally, methods of extracting aurocyanide complexes from carbonaceous ore may be explored in the event that cyanide is introduced to the RIL circuit.

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