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INNOVATIVE MODIFICATIONS TO HIGH DENSITY SLUDGE PROCESS

BERNARD C. AUBÉ Noranda Inc. Technology Centre, 240 Hymus Blvd Pointe-Claire, Québec H9R 1G5, Canada

Abstract: An improved Acid Mine Drainage (AMD) treatment process is currently operating at Noranda Inc., Heath Steele Division. The process is similar to the Conventional High Density Sludge (HDS) Process, without a rapid mix tank nor a flocculation tank. These modifications allow for savings in both capital and operating costs without compromising the treatment efficiency. Following the process start-up, aeration was also discontinued, for savings in operating costs. The final result is a compact, efficient plant which performs as well or better than other HDS plants.

Piloting results leading to this water treatment plant were performed in 1996 and are described in detail. Two distinct processes were examined in the pilot campaign: the Conventional HDS Process and the Geco HDS. Process performances were similar when comparing the resulting effluent and sludge qualities from these two processes. Final sludge densities of 25 to 27% solids were attained while the viscosities remained within acceptable limits. Different flocculants were tested with variations in sludge recycle rates. Aeration was also assessed as high ferrous iron loadings were oxidised by simple pH increase. The primary result of this pilot campaign a demonstration that a Rapid Mix Tank is not a necessary component of an HDS Process.

Key Words: Acid mine drainage, treatment, lime neutralisation, precipitation, high density sludge

INTRODUCTION

A Water Treatment Plant was commissioned at Noranda Inc., Heath Steele Division in October of 1997. A pilot campaign was conducted prior to construction, to better define the most efficient process design for treatment at Heath Steele. Two processes (Conventional HDS Process and Geco HDS Process) and variations thereof were tested and are discussed here. The chosen design is detailed, and the full-scale operation is discussed.



Figure 1. Conventional HDS Process.

The Conventional HDS Process (Kuit, 1980) typically consists of 3 reactors plus a clarifier (or clari-thickener). As shown in Figure 1, the acid mine drainage (AMD) is fed into the Rapid Mix Tank (RMT), a small reactor with a minimum retention of about 5 minutes. A mixture of sludge and lime is used to neutralise the AMD to a controlled pH in this reactor. The resulting slurry cascades to a second reactor where air is often sparged for iron oxidation. This is generally called the Lime Reactor (LR) and has a minimum retention time of 30 minutes to allow for complete precipitation of metals into hydroxides at the increased pH. The third reactor may have a retention time of only 3 minutes to contact a flocculant with the precipitates and form agglomerates. These agglomerates settle in a clarifier to produce an underflow sludge. The clarifier overflow represents treated water which can meet Canadian discharge quality requirements. The sludge is recycled to the Lime/Sludge Mix Tank (L/S) to be contacted with lime and used to neutralise the incoming AMD in the RMT.





A second process tested in the pilot program is the Geco HDS Process (Aubé and Payant, 1997) as shown in Figure 2. Four reactors are used for neutralisation of AMD via the Geco HDS Process. In the first reactor (R#1), the AMD is contacted directly to recycled sludge with a retention time of at least 30 minutes. The second reactor is a RMT used for primary pH control by addition of a lime slurry. The reaction is completed in a larger vessel of at least 30 minute retention (R#2). Here also, aeration can be provided for ferrous oxidation. The fourth vessel is smaller, and used for flocculation just as with the Conventional HDS Process. The clarifier underflow is recycled and the overflow is released.

HEATH STEELE PILOT PLANT CAMPAIGN

Introduction

At Noranda Inc., Heath Steele Division a pilot plant program was conducted from January to April 1996 for a water treatment plant (WTP) to be commissioned by the fall of 1997. This pilot program was to help define the optimum process for Heath Steele's (HS) WTP. The purpose of the pilot campaign was to determine if the RMT made any difference in the chemistry and efficiency of the HDS Process and to compare the Conventional HDS Process with the Geco HDS Process.

HS Raw Water

The raw water used for the tests was taken from an AMD storage pond considered to have chemical properties closest to those expected in the long term. The acidity and metal concentrations at this location varied considerably, but the neutralisation processes seemed not to be affected by these variations. Table 1 shows a summary of the raw water quality treated during this pilot plant campaign.

	1 401	e I. Ileaul S	ICCIC Raw W	ater Chemisuy	(Jan 12 - Apr	u 4, 1990)		
mg/L	Cu	Pb	Zn	Fe	Fe ²⁺	Solids Production		
Average	17.0	0.9	179	259	205	1470		
Maximum	56.5	3.0	490	915	607	5520		
Minimum	4.3	0.4	96	82	71	660		
Temperature: 2.2 to 8.7 °C				Dissolved Oxygen: 12.9 to 13.8 mg/L				
Oxidation/Reduction Potential (Redox Ag/AgCl): 447 to 561 mV						pH: 2.51 to 3.28		

Table 1. Heath Steele Raw Water Chemistry (Jan 12 - April 4, 1996)

Flocculants

The choice of flocculant was not to be included among the parameters to be researched during this campaign. As Percol 727 had been found to be the best polymer for the Geco Division and since another nearby Noranda mine (Brunswick Mining Division) also used Percol 727, it was decided that this polymer would be used. The flocculant optimisation could be done once the full-scale WTP was commissioned. Early tests, however, showed that it was difficult to form a high density sludge (defined as a sludge with a solid content >20%) with Percol 727. The highest density achieved with this flocculant was less than 10% solids. A series of bench scale tests then showed that Percol E10 was the preferred polymer for the optimum density/clarity combination in the pilot scale.

In the early pilot tests, an attempt was made to meet the required effluent limit of 0.5 mg/L Zn concentration with Percol E10. As the pilot clarifier is not as efficient as a full-size clarifier, flocculant dosages were increased until the effluent was of reasonable turbidity. The polymer dosages required for such clarity in the pilot scale were well beyond typical dosages from AMD treatment plants. Treatment plant dosages normally vary from 2 to 10 ppm, with 3 to 5 ppm as a typical dosage (polymer dosage is calculated as mg of polymer fed to the system over kg of influent AMD). For the early tests, the flocculant dosage was in the order of 15 to 20 ppm. To obtain a clear idea of the expected sludge characteristics in the full scale, a compromise needed to be made on the effluent for a representative flocculant dosage. As shown in Table 2, there are therefore two sets of tests compared: HDS-hi and Geco-hi for the high floc dosage, and HDS-lo and Geco-lo for the lower dosage rates.

Process	Flocculant Dosage	Effluent Zn Concentration	Maximum Sludge Density	Average Sludge Viscosity		
	(ppm)	(mg/L)	(% solids)	(cP)		
HDS-hi	18-19	0.64	17.5	73		
HDS-lo	4-5	2.33	27.3	40		
Geco-hi	15-17	0.45	22.5	56		
Geco-lo	4-5	3.07	24.8	32		

Table	e 2.	Heath	Steele	Test	Results.	

Rapid Mix Tank

One of the objectives of this campaign was to determine if removal of the RMT could in any way affect the efficiency of the HDS treatment process. During this investigation, two Conventional HDS tests (Fig. 1) were carried out with a RMT and two were completed without. The criteria used for comparison were primarily the pH in the LR while also monitoring the effluent quality, sludge density, and sludge chemistry. In the pilot plant, the pH was controlled using a rudimentary proportional controller and a peristaltic pump to feed lime slurry to the L/S Mix Tank. The pump operated on/off in response to a signal from the controller.

Even with this simple pH controller, the pH remained stable in the LR. For a test that lasted over two weeks and saw iron concentrations more than triple (from 300 to 915 mg/L) within 18 hours, the pH in the LR was maintained within 0.3 units from the setpoint (pH 9.3). The effluent and sludge qualities were not affected by removing the RMT. The highest achieved density occurred while operating without the RMT, primarily because this test was pursued for a longer duration and at a low polymer addition rate.

Effluent Quality

The primary goal of water treatment is to consistently meet the discharge criteria. The constraining limit for Heath Steele is a maximum Zn concentration of 0.5 mg/L. Table 2 shows average Zn concentrations calculated from effluent samples taken when the sludge density was higher than 10% solids. On average, the only test that met the effluent criteria was the Geco-hi, even though the polymer dosage rate was 2 to 3 ppm lower than that of the HDS-hi. None of the processes consistently met the Zn limit with this inefficient pilot-scale clarifier. For both processes with low polymer dosages, the effluent Zn concentration was well above discharge criteria.

Following a close examination of this effluent data, it was stipulated that the effluent quality may be directly related to the amount of polymer available for agglomeration of the mass of solids in the slurry. Figure 3 shows the effluent Zn concentration in relation to the mass ratio of flocculant (in mg) to solids (in g) in the slurry. This is a logical correlation as the Zn released in the effluent is nearly all contained in the suspended solids. As can be seen in this figure, there is a minimum dosage of 0.1 mg floc/g of solids below which there is little or no effect of flocculation. For the Geco Process, all samples beyond 0.25 mg/g met the required effluent criteria of 0.5 mg/L Zn. Although the HDS Process showed a little more scatter beyond the 0.25 mg/g, it is believed that with the increased efficiency of a full scale clarifier the process could produce an acceptable effluent at this dosage.



Figure 3. Effect of Polymer Dosage on Effluent Zn Concentration.

Sludge Density

The third column in Table 2 shows the maximum density achieved with each of these tests. The HDS Process at low flocculant dosage resulted in the highest density. The highest measured density during the Geco-lo was close, and the Geco-hi was also above 20%. There is a big difference between the HDS-lo and the HDS-hi, the latter being the only one of the four tests which did not attain 20% solids. As the HDS-lo had the highest density and the HDS-hi the lowest, it is evident that the HDS process is very sensitive to flocculant dosage. The Geco Process, although its maximum density was slightly lower than that of the HDS-lo, did not show as much sensitivity to polymer dosage.

Sludge Viscosity

The viscosity of these sludges (also in Table 2) range from 32 cP for the Geco-lo to 73 cP for the HDS-hi. The effect of flocculant is also evident here as the higher polymer dosages resulted in higher viscosities. It is important to note that these viscosities are quite acceptable, as a standard centrifugal pump could easily pump a sludge with less than 100 cP. The Geco Process resulted in the lowest viscosity sludge, but all processes were acceptable in this aspect.

Ferrous Oxidation

Air was sparged in both the Geco and Conventional HDS processes for oxidation of ferrous iron to ferric. The ferrous concentrations were determined by titration at least once per day. As was shown in Table 1, the average ferrous feed concentration was 205 mg/L. While operating the Geco HDS Process, all the ferrous oxidised in the first reactor, prior to aeration. This can be partially explained by the fact that the raw water had a high redox and the increased pH of the reactor accelerated the iron oxidation. The dissolved oxygen (DO) content of the raw water was always saturated, and some oxygen may have been diffusing across the surface of the slurry.

The most widely accepted iron oxidation reaction at alkaline pH is as follows (Aubé and Payant, 1997):

$$Fe(OH)_2 + \frac{1}{2}H_2O + \frac{1}{4}O_2 \rightarrow Fe(OH)_3$$

The stoichiometry of this reaction shows that 0.25 oxygen moles are needed to oxidise 1 ferrous hydroxide mole. The mass equivalent relates to 1 mg of oxygen to oxidise 7 mg of ferrous iron to ferric. During the pilot tests, the raw water was essentially saturated in oxygen at a temperature of about 3°C. This represents approximately 13 mg/L of oxygen or enough to oxidise 84 mg/L. A second source of oxygen could be transferred from the surface of the reactor. Assuming an oxygen transfer coefficient (k_L) of 4.7x10⁻² cm/s results in approximately 17 mg/L of additional oxygen for the pilot reactor geometry. By subtracting the 2.5 mg/L of oxygen that remained in the reactor, this accounts for oxidation of 192 mg/L of ferrous.

This seems sufficient to account for the average ferrous feed, but it is more difficult to determine the oxidant when we consider that concentrations of up to 600 mg/L were oxidised prior to aeration. No other oxidant was ascertained, but biological activity was suspected. Since this discovery, ferrous oxidation without aeration was found to occur at three other Noranda Divisions (Mattabi, Brunswick, and Waite Amulet).

Sludge Chemistry

Summary results of analysis of HS sludges are shown in Table 3. Sulphate, assumed to be in the form of gypsum, is essentially the same for both processes. Total carbon content for both processes was near 1%, with almost half in the form of TIC (Total Inorganic Carbon) for the Geco Process and three quarters as TIC for HDS. Most of the organic carbon is from the flocculant: at 1.5 g/L solid production and a polymer dosage of 10 ppm, the flocculant content of the sludge would be about 0.7%, most of it carbon. The higher organic carbon content (obtained by difference) from the Geco Process can be explained by the fact that this test was started at a high polymer dosage, while the Conventional HDS Process sample was run at low polymer dosage.

The higher inorganic carbon, Ca, and Mg contents of the Conventional HDS Process suggest that more carbonates are formed in this process. When converting these TIC concentrations to carbonate this represents 3.7% and 2.3% for the Geco and Conventional HDS processes respectively. This suggests that the HDS Process will use more lime for neutralisation, but produce a sludge with a superior buffering capacity (Zinck and Aubé, 1999). This was confirmed by slightly lower concentrations of metals leached in sludge stability tests (not presented here). In the Conventional HDS Process, most of the carbonates may be formed in the L/S Mix Tank as this reactor operates at a pH above 11. In the Geco Process, the carbonates are dissolved in the first reactor and serve to partially treat the AMD. A detailed account of the chemical reactions occurring in the Geco Process can be found in Aubé and Payant, 1997.

	SO4	Total C	TIC	Al	Ca	Cu	Fe	Mg	Mn	Zn
	(%)	(%)	(%)	(%)	(%)	(%)	(%)	(%)	(%)	(%)
Geco	5.61	0.97	0.46	4.57	1.58	1.28	22.96	1.29	2.11	16.60
HDS	5.78	0.98	0.74	4.37	2.50	0.95	21.96	2.11	2.31	17.16

Table 3. Sludge Chemistry of Tests at HS

Although the lower carbonate content suggests that the Geco Process would consume less lime, this could not be positively ascertained. The lime consumption of each test was carefully monitored, but the extreme variations in feed water quality masked any significant difference which may have otherwise been measured. As was shown in Table 1, the feed Zn concentrations varied five-fold and Fe concentrations changed by an order of magnitude (min 82 and max 915 mg/L). An attempt to normalise the data to the variations in feed quality failed to show a clear trend.

Particle sizing analysis of sludges showed the HDS to have a D_{50} of 2.92 µm and the Geco particles to have a D_{50} of 2.38 µm. Scanning Electron Microscope (SEM) analysis done on sludge samples from both processes showed apparently spherical particles of 1 to 3 µm in diameter. No difference between the Geco and Conventional processes was apparent.

Pilot Plant Campaign Conclusions

- 1) The pilot results showed that the RMT was not necessary for pH control, at least at a reduced scale. It also showed that the sludge chemistry and effluent quality was not affected by removing the RMT.
- 2) The Pilot Plant Campaign failed to show a significant advantage to the Geco Process. The highest density was attained with the Conventional HDS Process, and only marginal improvements could be found in effluent quality and sludge viscosity using the Geco Process. The lime consumption data was inconclusive due to variations in raw water chemistry.
- 3) Ferrous oxidation results suggested that aeration was not necessary at Heath Steele, although it was recommended to retain the blowers in case of upscaling differences or changes in raw water chemistry.
- 4) It was recommended that Percol E10 be used for flocculation at the plant start-up.

HEATH STEELE WATER TREATMENT PLANT

Following the results of the pilot campaign and because of Noranda's experience with the HDS Process, HS decided to build a WTP based on the Conventional HDS Process but excluding both Rapid Mix Tank and Floc Tank. The plant, commissioned in the fall of 1997, has been operating successfully since start-up.

A raw water retention pond (or Buffer Pond) of $820,000 \text{ m}^3$ was constructed for storage during the spring freshet and for high precipitation events. This pond also serves to equalise the AMD and feed a relatively constant water quality to the WTP. This is important at HS as the varied sources of mine water, waste rock drainage, and site drainage offer the entire range of AMD concentrations.



Figure 4. HDS Process at Heath Steele Division.

The plant consists of two process reactors and a clarifier (Fig. 4). At the peak flowrate of 20,000 L/min, the Lime Reactor offers a little more than 40 min of retention for complete precipitation of heavy metals. A flocculant is added in the launder (trough conducting the slurry) immediately after the LR overflow and in the clarifier feed well. The polymer feed split is approximately 60/40 in favour of the first feed point. Following solid/liquid separation in the clarifier, the underflow sludge is recycled to a 14 m³ Lime/Sludge Mix Tank. The minimum retention supplied in this vessel is of about 5 minutes. Sufficient lime slurry is added in this reactor to control the pH of the treatment process. The L/S Mix cascades into the LR to complete the cycle. Underflow sludge from the clarifier is purged intermittently to control the inventory.

While many water treatment plants have polishing ponds for removal of residual suspended solids from the clarifier overflow, HS opted for a slightly

different scenario. The mine has deposited tailings in three distinct ponds over the years. All ponds are presently submerged and this water cover will be maintained over all three ponds following closure (planned for fall 1999). In order to prevent the acidification of these ponds and to maintain the required water level, the WTP effluent is discharged into the upper pond to cascade through the two others prior to release. The extended retention offered by this system will help settle any residual suspended solid prior to release from the lower pond.

Rapid Mix Tank

To operate the pilot plant without a RMT was not necessarily representative of full scale operation. The only conclusion transferable from the pilot scale is that the process itself was not affected in that the effluent and sludge qualities were

the same. The remaining challenge was pH control in a vessel with a retention time of 40 minutes at high flowrate and over 2 hours at lower flowrates.

One of the concerns was that treatment would be entirely dependent on a single pH probe at the LR overflow. When a RMT is included, there are two pH probes in the system: one controlling the instantaneous pH in the RMT and another at the LR overflow for verification and possible adjustment of the RMT pH setpoint. If the control is only in the LR, the next continuous pH measurement is taken at the clarifier overflow. This suggests a dead time of 2 hours or more, depending on the flowrate, before the pH is verified.

For safe operation, redundancy was added at the plant. Two pH probes were installed to control the lime addition. At start-up in October 1997, the probes were placed at the surface of the reactor and significant variations and differences between the two were measured. This is expected in a large vessel with acid water and lime added from the top. Both probes were then moved to the overflow of the LR, in the riser. The measured pH values stabilised and were the same following calibration. The pH control loop is arranged so that it can use probe A, probe B, or the average of the two. The default setting is on the average, but in case a significant difference is measured between the two probes, the one considered accurate is used alone. This also allows for continued pH control while servicing one probe. Emergency shutdown procedures can be automatically engaged if either probe measures a pH value outside the acceptable range.

Even with this redundancy, the concern about pH control with the long retention of the LR remained. On top of the extended residence time in the LR, there is added dead time in the L/S Mix Tank, as the pH control variable is the valve opening for the addition of lime in this tank. At low flowrates, it can take more than 15 minutes before the added lime even enters the LR.

pH Control

Typical pH control is done using a PID (Proportional Integral Derivative) feedback loop. Only PI are normally used for simple pH control (D=0). This system measures the resulting pH and corrects the lime addition rate to minimise the difference between the setpoint and process variable (measured pH). With dead-time, some oscillations are expected but the concern is on the magnitude of the oscillations. With a pH setpoint of 9.5, for example, oscillations of pH 9.3 to 9.7 are normal, and from 9.2 to 9.8 remain acceptable for this treatment plant.

When a RMT is used, the oscillations in this small reactor can be relatively significant, but the added retention in the LR damps the pH and stabilises it. Without a RMT, there is no backup, no additional damping as the measurement directly reflects the treatment pH. If this value was outside the desired range, it could either result in inefficient Zn precipitation (low pH) or expensive lime wastage (high pH). The positive aspect is that today's process control knowledge and PLC programming versatility make optimisation of parameters (P and I) a simple task.

There also exists the possibility of adding a feed-forward loop (or ratio control) to correct the lime addition rate whenever the flowrate is changed. The lime slurry is delivered to the L/S Mix Tank on a recirculating loop with a fully open/fully closed pinch valve to control the lime addition rate. The lime dosage is fixed by a split-time proportional control of one minute on this valve. This means that for a control variable of 25%, the valve is open for 25% of the time (15 seconds) and closed for the remaining 75% of the sequence (45 seconds). With this control, the mass of lime added is therefore directly proportional to the control variable (linear). As the lime consumption is also directly proportional to feed flowrate, doubling the feedrate should double the control variable (valve opening). The feed rate is defined by the operator and a ratio control would multiply a change in feed rate directly with the set control variable of the lime addition.

pH Control at Start-up

During start-up in the fall of 1997, the pH control parameters (PID) were set and tests were conducted to verify the response of the system without a ratio control. In one trial, the feed flowrate was changed from 6,800 L/min to 10,000 L/min in a single step increase. The pH stayed within 0.2 points from the setpoint, which is more than acceptable. The feed-forward control loop was then never programmed in as there seemed to be no need of it. After operating for more than a year, it is now evident that the pH control without a RMT is achievable.

Solids Flocculation

A Floc Tank is normally used for good contact of diluted polymer with the precipitates from treatment. A few Noranda treatment plants had a Floc Tank with multi-point injection of polymer. The first injection point is always before the Floc Tank, the second may be at the Floc Tank overflow and another at the clarifier feed well. Trials had been done to feed the flocculant at any single feed point to compare the results. The clarifier overflow was generally used to define a change in solid/liquid efficiency. Results showed little or no difference if the polymer was fed only before or after the Floc Tank, but an increase in turbidity was measured when the polymer was dosed in the clarifier feed well only. This suggests that either the retention time and/or the mixing intensity are insufficient in the feed well alone, but sufficient in the launder.

Testing the removal of the Floc Tank could not be done at the pilot scale due to the low flowrates and lack of mixing intensity. It was therefore decided to remove the Floc Tank from the plan based on the results obtained at other Noranda treatment plants. The primary polymer injection point was placed in the launder immediately after the slurry overflows from the LR. The necessary agitation and retention is supplied by the straight launder conveying the slurry to the clarifier. A second injection point was installed immediately before the clarifier feed well. The floc discharge system is arranged so that the fluid is distributed evenly throughout the surface of the slurry in the launder. This is accomplished by feeding the diluted polymer solution into a receptacle with a toothed edge at the overflow point to distribute a dozen streams of solution across the slurry surface.

Based on the pilot results, the WTP used Percol E10 at start-up. While the sludge density was steadily increasing in the first days of operation, the clarifier overflow turbidity was also increasing. The overflow Zn concentrations exceeded the target of 0.5 mg/L. The long retention time in the tailings ponds allowed for settling of the Zn particulates so that the effluent requirements were always met, but the goal was to produce an acceptable water quality from the WTP. HS then changed to Percol 727 and managed a significant improvement in turbidity, but the overflow Zn concentrations remained borderline.

A plant review conducted in February of 1998 examined the issue of clarifier overflow suspended solids in greater detail. One finding was that the mixing in the trough seemed insufficient; the diluted flocculant solution was discharging over a laminar flow of slurry. The plant, designed with a hydraulic capacity of 25,000 L/min, was operating at 6,600 L/min at the time. Flocculant theory suggests that quick mixing at first, followed by less vigorous mixing allows the polymer strands to first make contact with all small precipitates, followed by a general agglomeration of these strands to create large flocs. It is these large flocs that settle quickly and insure efficient clarification. At HS, it was the initial vigorous mixing that seemed to be lacking, which would result in some small particles not being contacted to the polymer.

HS then tried to induce some mixing by adding an obstruction in the launder and breaking the laminar flow. No improvement was noted. The next attempt at improvement was to increase the concentrated polymer solution from a setpoint of 0.15% to 0.30%. This gradually improved the turbidity and brought the overflow Zn concentrations well within the target. When this increase was made, the floc feed rate was decreased to keep the polymer dosage approximately the same. Why the more concentrated solution performs better is not known. The polymer dosage was near 5.7 ppm, or about 0.30 mg of polymer per g of solids. This concurs with the pilot results of a minimum of 0.25 mg/g as presented in Figure 3.

Aeration

Following the ferrous oxidation results of the pilot campaign HS designed the LR agitator and support structure so that it could operate without air. During the plant review of February 1998, the blowers were stopped and the ferrous concentration measured with time. Ferrous titrations determined that approximately 80 mg/L of the 140 mg/L Fe in the feed was in the ferrous form. The ferrous measurements in the LR remained below detection (<1 mg/L) limit. This simple test showed that aeration was unnecessary at Heath Steele. By stopping the 40 HP blower, approximately \$13k per year is saved in power consumption.

In November 1998, one of the mine de-watering systems was decommissioned as this mine is being allowed to flood due to the impending closure. This mine water was the individual source which contained the highest concentration of iron and also had a very high redox potential (E_h of ~630 mV). When the de-watering was shut down, the iron concentrations in the WTP raw water decreased, but the ferrous that remained did not oxidise without aeration. Green sludge was

formed, indicating ferrous hydroxide precipitates. The aeration was re-started at that point and remains in operation. Once the mine water level reaches the upper target, this water will be pumped to the Buffer Pond again and aeration may no longer be necessary.

CONCLUSIONS

- A Rapid Mix Tank is not necessary for efficient pH control when treating Acid Mine Drainage. Even with variations
 in feed water quality and flowrates, the pH can always be adequately controlled in a single reactor with a minimum
 retention time of 40 minutes.
- 2) Aeration is not always necessary, even if significant concentrations of ferrous must be treated.
- 3) Efficient flocculation of precipitates can be achieved without a Flocculation Tank.
- 4) A higher concentration of initial polymer solution resulted in improved turbidity and excellent effluent Zn concentrations.

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