Presented and published at the 41st Conference of Metallurgists, in Montréal on August 11-14, 2002.

Commissioning a New Water Treatment Process: The Geco Experience

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ABSTRACT

Noranda Inc. Geco Division was a copper-zinc mine that operated for 40 years in Northern Ontario. In 1995, its last year of operation, a new treatment plant was designed and commissioned to treat acid mine drainage (AMD) from the affected area. A new treatment process developed at the Noranda Technology Centre was to offer higher sludge density than existing processes. Modifications made during detailed engineering resulted in a significant difference in the process. Plant start-up posed some important challenges particularly due to sludge pumping problems and an under-sized clarifier. Through a pilot campaign, these problems were resolved. With a different sludge recycle strategy and sand filtration, the plant could not only meet the design guidelines but also surpass them. Over the years, changes in AMD quality and scaling potential made the sand filters unusable. The plant is currently operating well enough to meet today's modified objectives, but significant lessons were learned concerning implication of specialists throughout development of a new process and with the importance of predicting future requirements.

INTRODUCTION

The original Geco ore body was staked, promoted, and brought to life by the *G*eneral *E*ngineering *Co*mpany of Toronto (origin of the "Geco" mine name) in conjunction with Mining Corporation of Canada. The first ore concentrates were shipped by rail in October 1957 with copper to Noranda Horne smelter and zinc to smelters in the United States.

The Geco ore body was made up of a core of massive sulphides consisting of pyrite, pyrrhotite, sphalerite, chalcopyrite, galena, and marcasite. Appreciable silver was associated with the chalcopyrite. Over the course of the mine life there was approximately 56 million tons of muck developed and hoisted to surface. Of the mined ore, approximately 50 million tons were generated as tailings.

Total ore concentrates are as follows:
COPPER…...…. 950,000 Tonnes
ZINC…………. 1,650,000 Tonnes
SILVER ……… 64,730,502 oz

A mine closure plan was initiated in 1990 as a result of the forthcoming depletion of the ore body. The mine was closed in December of 1995. A 4-man environmental crew was retained to maintain water treatment operations, surface and ground water monitoring, and site supervision for reclamation. At the end of 2001, all assets had been removed and all buildings (except those associated with water treatment) were demolished. Today, a single person runs the plant site and treatment operations.

GECO WATER MANAGEMENT HISTORY

Prior to the 1970's, excess water was simply released from the tailings impoundment without control. In 1972, a recycle system was put into place as tailings water was pumped back to the concentrator to be used in the milling process. This system was implemented to reduce the volume of untreated tailings waters being discharged to the environment.

Mine dewatering was also rerouted from the mill circuit and pumped directly into a drainage ditch. This ditch was separated down its length with a clay dike to intercept clean drainage waters and maintain the clean runoff separate from tailings seepage and mine waters. Although the tailings and seepage waters, commonly known as "Acid Mine Drainage" (AMD) were being limed through the mill circuit and again prior to discharge, they still contained important concentrations of heavy metals.

Phase 1

In 1973 and 1974, Geco took on Phase 1 of the Waste Water Treatment Plant (WWTP). This included the construction of a main building, lime silo, and pilot plant. Various studies and pilot plant testing were performed during this period to establish the best method of treating AMD. Geco was among the first Canadian mine sites to install a lime treatment system for controlling AMD discharge.

Phase 2

In 1975, Phase 2 was initiated with a 75' (22.9 m) diameter reactor clarifier, raw water holding pond, and associated piping being commissioned. Treatment plant feed water (influent or raw water) was collected via seepage collection ditches and pumping facilities strategically located at the toe of the tailings dams. All contaminated waters reported to the main holding pond adjacent to the WWTP.

A certain amount of recycle water (tailings pond water reporting to mill) was bled off, upstream of the holding pond as make-up water to supply the WWTP with a minimum amount of flow to operate effectively. This practice also helped control the water level in the tailings pond. At this time, only "Low Density Sludge" (LDS) was created and pumped back directly on the tailings beach and consequently into the tailings pond. During the next 20 years of operation, ongoing changes to equipment, reagents and automation was implemented to further improve quality of effluent, manpower requirements, and to decrease operating costs.

NEW TREATMENT PROCESS DESIGN

In 1994, with the forthcoming mine closure in November 1995 and the age of the WWTP, it was clear that the time was well chosen to upgrade the plant. As long term sludge disposal costs could be important, it was decided to design the plant to produce high density sludge. This upgrade consisted of completely revamping the treatment building, installing new pumps, blowers, compressors, reactor tanks, quicklime paste slaker, electrics, and automation. The original reactor clarifier was remodelled to a standard clarifier with centre feed well and thixopost rake arms and drive.

Plant Design Basis

The plant was to treat AMD at a rate of 2,000 usgpm (7,571 L/min) to remove heavy metals and control pH. The raw water had a pH of about 3.0 and contained approximately 300 mg/L iron, 40 mg/L Zn and lower concentrations of Al, Cu, and Mn. The neutralisation would be completed using quicklime (CaO) slaked and slurried on-site. The final pH was to be 9.2, as was the case for the older LDS plant. The desired criterion for HDS was set to a minimum target of 20% solids in the sludge.

This design flowrate was based on the historically treated volumes plus other expected flows. Noranda has partial environmental responsibility for Willroy, a nearby decommissioned mine. Although the level of responsibility was disputed at the time, it was thought likely that some AMD from the Willroy site would eventually be treated at this plant. Also, site runoff and maintaining a low level in the underground mine workings were to impact the yearly treated volume.

A process that had been researched extensively at the Noranda Technology Centre, the NTC Process was proposed (Kuyucak et al. 1995 [1]). This process is described in Figure 1 and the following section.

NTC Process

The Acid Mine Drainage (AMD) is fed into Reactor 1 (R#1), where it is contacted with recycled sludge only. According to process design, a pH of 4.5 is to be controlled by the rate of sludge added to this reactor. Sludge is also to be contacted with lime in the Lime/Sludge Mix Tank. Sufficient lime is added into the Mix Tank to control the pH in R#2 at the desired setpoint. The overflows from both R#1 and the L/S Mix Tank are contacted with aeration in R#2. A flocculant is fed to the overflow stream from this reactor and a small Flocculation Tank allows for proper contact of a polymer with the precipitates.

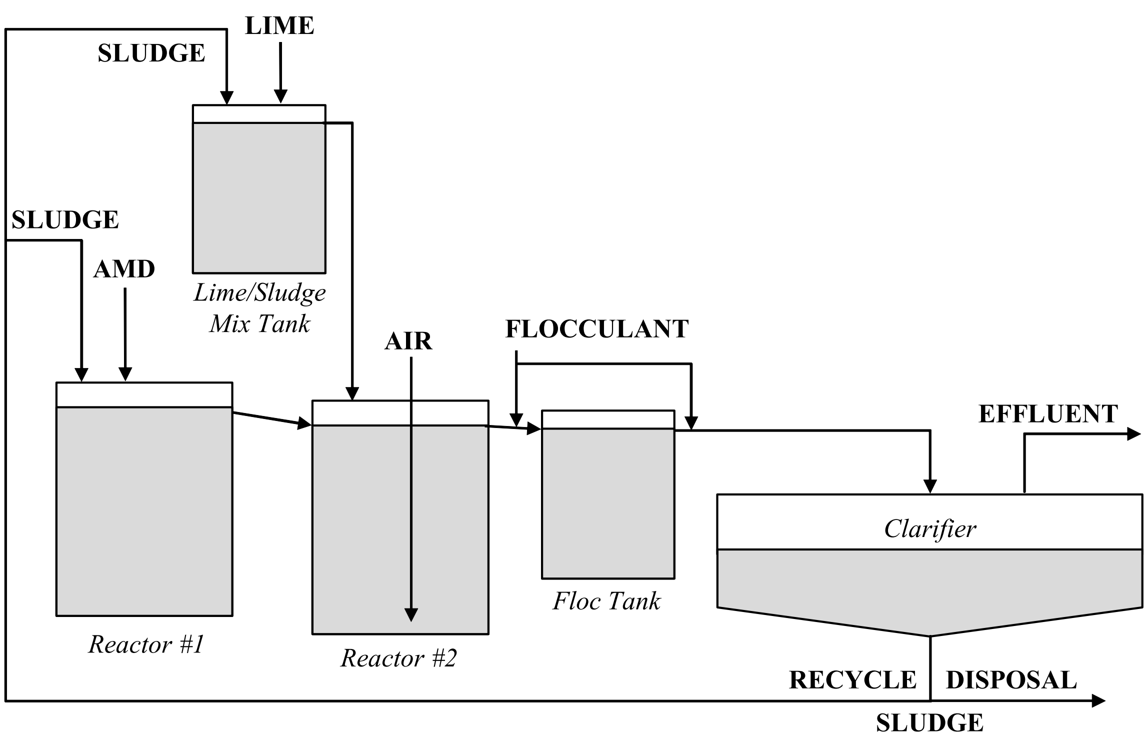

Figure 1 - NTC Process

The flocculated slurry is then fed to a clarifier for solid/liquid separation. The clarifier overflow is released as a treated effluent meeting all discharge criteria. The clarifier underflow is recycled to the process as previously explained. A fraction of the sludge is either bled continuously or in batches to control the solids inventory.

This process had been tried in a number of pilot experiments and resulted in effluent and sludge qualities at least as good as those of other HDS processes. Geco took this process design and submitted it to contractors for construction.

Geco Process

During the detailed engineering phase of the project, it was acknowledged that it would be much easier and cheaper to construct the four vessels (tanks in Figure 1) in-line. Through a communication error with a treatment specialist at NTC at the time, it appeared that this modification would make little difference to the process. The result, in fact, was an entirely different process, as shown in Figure 2.

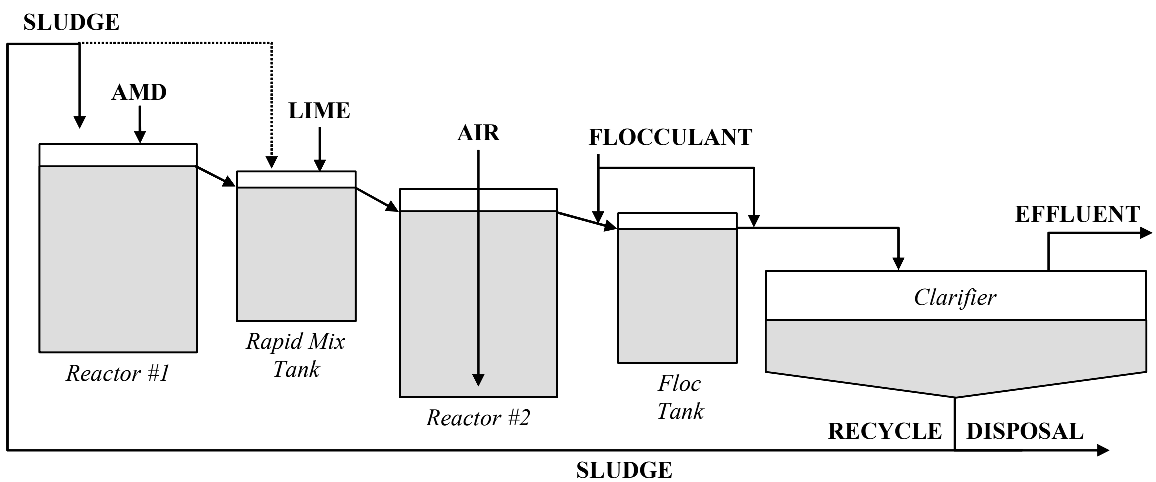

Figure 2 - Geco Process

By moving the L/S Mix Tank in line with the other vessels, it simply became a Rapid Mix Tank. In fact, other than slightly improved pH control, this tank could simply be removed from the process without affecting results (Aubé, 1999 [2]). The process now starts with the AMD and sludge fed into the first reactor, overflowing to the Rapid Mix Tank (RMT) where lime is added for pH control. The process also had the option of feeding additional sludge to the RMT.

Reactor #2 received a premixed slurry and only aeration and retention time is added. The Floc Tank and clarifier serve the same purposes as for the NTC process.

Particularly to the untrained eye, these two processes look very similar, but mixing sludge with lime was a key part of the original process and was removed in the modification. All existing HDS processes prior to this plant had a L/S Mix Tank. In theory, the L/S Mix tank serves to form a coating of lime on the existing precipitates. When these lime-coated precipitates are mixed with the AMD, this forces additional precipitation reactions to occur on the surface of the existing solids. This is why the solids become heavier and therefore the sludge holds a higher solid content.

PLANT START-UP

Once construction was completed and treatment started, some of the typical problems occurred with piping and electronics. These issues added to costs but were not unusual or completely unexpected. The more important issues, and the measures taken to resolve them, are discussed in the following sections.

Start-up Problems

On the process control side, the sludge recycle ratio was initially set very low, to be increased gradually. By this time it was clear that this was not the originally intended process and risks of creating an unmanageable material in the clarifier demanded prudence. The recycle rate was therefore started at a volumetric ratio of approximately 2% (sludge flowrate/feed flowrate*100%). At this ratio, the clarifier effluent was clear and the sludge density of 4 to 7% solids.

The recycle ratio was then increased to 4% volumetrically. At this ratio, the sludge density increased to a range of 7.5 to 10% solids and viscosity problems began. These results caused serious concerns as the design target for sludge density was 20% solids. Sludge viscosity caused problems for both the clarifier and the sludge pumps.

For the clarifier, the sludge viscosity would cause it to stick to the rakes and turn with them as opposed to being pushed to the center cone. As a result, the slurry fed into the center well flowed directly into the clarifier cone with very little densification. This caused a “donut” to form in the clarifier. The rake torque would also increase and cause the rake mechanism rise. The rake mechanism was designed for a 1.5-ft (46 cm) automatic lift on high torque. These problems were time-consuming as the rakes needed to be manually lowered gradually to slowly cut through the gelatinous mass of sludge.

The sludge recycle and purge pumps simply could not convey the sludge when it became very viscous. SRL pumps used for conveying sludge do not create much suction and could not pull the sludge from the clarifier cone to the pump intake. Water had to be injected in the clarifier cone to facilitate transfer to the sludge pumps. This was part of the procedure when gradually lowering the rakes to cut through the high-viscosity sludge.

Pilot Project

A pilot plant was brought from the Noranda Technology Centre to try different operating scenarios and solve these problems. The pilot campaign and results are described in detail in Aubé and Payant, 1997 (3). The pilot plant was designed to operate at 1 L/min and contained a series of modular vessels arranged to represent the process. The retention times, pH control, polymer addition, and clarification could all be arranged to adequately duplicate the full-scale WWTP.

Pilot Results

The first pilot trial was a control test, to try and duplicate the existing conditions of the WWTP. After a few days operation, the peristaltic pumps used for recycling sludge could no longer convey the sludge and the test was abandoned. The results were quite similar to those obtained in the WWTP.

The second pilot test included pumping sludge at a starting recycle ratio of 12%. Although this was the design recycle rate, it would never have been tried in the WWTP due to risk of creating an unmanageable mass of sludge in the clarifier. In the pilot plant, the results were excellent. Instead of creating an amorphous mass of gelatinous sludge, a high recycle rate created a high-density, low viscosity slurry.

These impressive results were attained without modifying the process, only the recycle rate. Detailed tests showed three distinct modes of operation:

1 – Low-density sludge (LDS) operation:
Recycle rate in the 1 to 3% volumetric ratio, sludge densities below 7% solid content, pumpable sludge.

2 – Medium-density sludge operation:
Recycle rate in the 4 to 8% range, sludge densities of 7.5 to 15% solids, high-viscosity, gelatinous sludge mass causing clarifier and pump problems.

3- High-density sludge (HDS) operation:
Recycle rate in the 10 to 12% volumetric range, sludge densities of 25 to 35% solids, low-viscosity, easily pumpable sludge.

Sludge Analysis

Microscopic analysis showed that the low to medium recycle rates resulted in an amorphous mass of sludge particles interlaced almost as strands. Scanning electron microscope (SEM) photos could distinguish no specific particles at any resolution. The medium-density operation apparently didn’t affect the chemistry or resulting precipitates, only ended up increasing the population density of similar precipitates to the point where the sludge viscosity became a problem.

The high-density particles under microscope were apparently spherical nodules of 1 to 5 μm. They seemed to be compact, individual particles. The result was a slurry much like that expected in a mill flotation circuit. It appeared that by recycling a high solids mass, the precipitation reactions occurred on the surface of existing particles. This caused individual particles to grow and increase in specific density. The result is similar to that of the Conventional HDS process, but the Geco Process uses the neutralising potential of the sludge as a primary treatment.

The products of the two recycle scenarios could be compared to cotton balls for low-density and ball bearings for high-density operation. The large mass of cotton balls, though light, does not flow easily and is difficult to handle. The ball bearings, though heavy, flow easily and are readily handled by clarifier rakes or SRL pumps.

Pilot Results Scale-Up

Following the exceptional results from the pilot project, a full-scale trial was attempted. The results were similar: high-density sludge with low viscosity was created.

Scale-Up Issues

The clarifier overflow and final effluent seemed dirtier than usual, but the Total Suspended Solids (TSS) limit of 15 mg/L was not approached. TSS was used as the guideline to determine whether or not the effluent should be released and it was maintained typically at or below 5 mg/L. It was only after a monthly average iron limit of 1 mg/L was surpassed that a problem was recognized. This triggered a study of the clarifier suspended solids.

The Geco TSS contained about 30% iron. With a regulated monthly limit of 1 mg/L total iron, this means the clarifier overflow must contain less than 3 mg/L TSS. This could be achieved, but only in low-density operation. In HDS treatment, the solids loading to the clarifier were simply too high. While a typical HDS clarifier operates with a solid loading of 5 to 15 kg/(m^2•hr), the Geco clarifier could be treating over 30 kg/(m^2•hr) in HDS mode. Table 1 shows actual measurements of solids in the clarifier feed and how this impacts the solids loading.

The data in Table 1 shows that the clarifier could handle the solid loadings resulting from low-density operation, but would have problems with the loadings at the high recycle rate, no matter what the process. To bring the loadings down to an acceptable range and still produce HDS, the flowrate would have had to be cut by half.

The plant was designed for high-density sludge to save on long-term disposal costs, by cutting the flowrate, no savings would have been realised. Since mine closure, the plant operates on a seasonal basis and can run for a shorter period by treating at full capacity. By reducing the flowrate, labour and electricity costs will be increased unnecessarily, thus eliminating any advantage to creating HDS.

Table 1 - Geco Clarifier Solids Loading.

Geco WTP Data		
Design Flowrate	2000	usgpm
	7571	L/min
Clarifier Diameter	75	ft
	22.9	m
Surface Area	410	m2

	Low Density Operation		High-Density Operation	
Recycle rate	2	% vol	10	% vol
Recycle flowrate	40	usgpm	200	usgpm
Solids Content in R#2	5	g/L	25	g/L
Clarifier Loading	6	kg/(m2•hr)	28	kg/(m2•hr)

Return to Pilot Scale

In order to solve the problems with TSS, additional pilot runs were completed. Different types of flocculants were tested, a polishing pond was simulated, and sand filtration was attempted. As before, the sludge density quickly increased to above 20% solids. A change of flocculant improved effluent quality, but improvement was insufficient to meet discharge criteria. The polishing pond made little or no difference. Sand filtration, on the other hand, showed excellent final effluent quality. Although the pilot plant ran only a few days, scaling was also monitored and found to be acceptable.

SAND FILTRATION

In light of these pilot results and the facts shown in Table 1, it was clear that either a larger clarifier or sand filtration was necessary to operate in HDS mode. A larger clarifier would likely have resolved the problems, but would have been very expensive. Sand filtration was feasible as the current clarifier could easily produce an effluent with suitably low suspended solids to feed sand filter banks.

An additional advantage to sand filtration is that it could allow for continuous operation even when minor upsets to the plant or clarifier occurred. For example, a bad batch of flocculant could have resulted in a loss of solid/liquid separation efficiency in the clarifier, even a larger one. With sand filtration, as long as the pH control was adequate, the effluent would meet discharge criteria and release could continue. Continuous operation was considered important due to the impending added volume from the nearby Willroy property.

Commissioning Sand Filter Banks

The arguments above led to the decision to design and construct sand filtration banks for the Geco WWTP. These were commissioned in 1997, two years after completing the pilot tests.

Operations in the first few months were excellent, although the backwash frequency was above the desired rate. The plant could produce sludge with up to 30% solids and low viscosity. The recycle pumps and clarifier rakes had no problem handling this sludge. The clarifier effluent appeared dirty but normally contained less than 10 mg/L of TSS. Sand filter effluent (final effluent) contained the lowest concentrations of iron and zinc ever produced in the plant history. Essentially, as long as the pH was well controlled, the plant was releasing an excellent effluent.

The sludge produced during operation of the sand filter banks was characterised and compared to sludges sampled from other treatment plants. As discussed in detail in Aubé and Zinck, 1999 (4), the Geco Process was the only one found to create crystalline metal precipitates. Crystalline sludges are an advantage as they offer greater density and long-term stability. These have been formed in laboratory experiments and pilot projects, but no other full-scale plant has shown crystal formation (apart from gypsum, a non-metal component of most sludges).

As operations continued, the backwash frequency, triggered by the head across the filtration media, increased beyond the acceptable range. Following shutdown of the filtration banks, the media, consisting of a layer of anthracite over a bed of sand, was found to be agglomerated. Samples were sent to NTC where the media was dried, broken up, and passed through a fine sieve. This allowed for collection of the material causing the agglomeration. Analysis showed that the agglomeration was caused by gypsum.

Gypsum Scaling

At the time when the pilot tests were run, the iron concentrations were in the order of 300 mg/L and sulphate about 2000 mg/L. By the time the sand filters were commissioned, the iron and sulphate concentrations often exceeded 500 mg/L and 3000 mg/L, respectively. High concentrations of iron required more lime for neutralisation and resulted in an increased dissolved calcium concentration. Combined with the high sulphate concentrations, this put gypsum well over saturation and caused precipitation in the form of scaling.

It was finally concluded that the sand filters should not be used when the raw water sulphate concentration exceeds 2500 mg/L. At those high concentrations, the scaling rate is too fast to be economical, as the filter media would need twice yearly replacement. It is therefore cheaper to operate in LDS mode and not replace the media on a yearly basis.

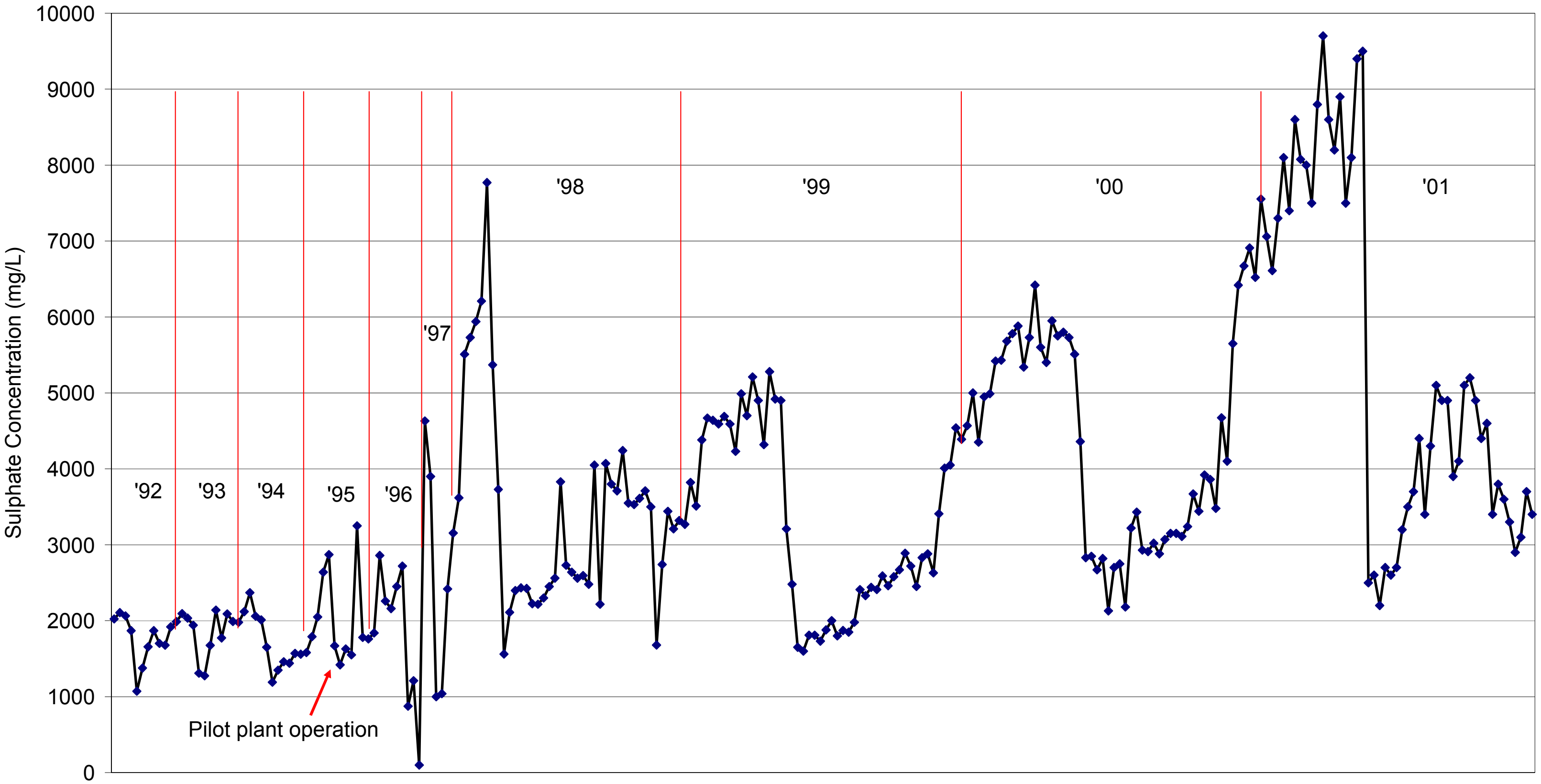

Figure 3 - Sulphate Concentration in Geco Raw Water (1992 to 2001)

Presented and published at the 41st Conference of Metallurgists, in Montréal on August 11-14, 2002.

As shown by Figure 3, the sulphate concentrations have been increasing every year. This is due to a number of reasons, starting from continued oxidation of the high pyrite content in the tailings. Other reasons include the fact that there is no longer any added alkalinity from the mill (as when it was operating) and very acidic toe seepages are pumped back into the pond. Note that the yearly cycling is due to concentration of the sulphates in winter under the ice, then dilution following spring thaw. The pilot tests were conducted during the summer of 1995, when the sulphate concentration rarely exceeded 2000 mg/L. The graph clearly shows that sulphate concentrations are mostly above 2500 mg/L since 1997.

Current Operating Strategy

The current tailings closure plan includes covering and revegetation. Tailings banks have all been re-sloped and graded, but the interior sections not covered by water need a natural cover for aesthetics and dust control. This is why the sludge from the WWTP is presently used to cover the tailings in cells. This sludge can then be revegetated more easily than acidic tailings. For a faster cover, low density sludge is actually preferred at this stage.

Once the tailings are sufficiently covered, the additional sludge produced will be discharged to the underground workings through an abandoned shaft. The WWTP will preferably be operated in HDS mode at this time. To allow for use of the sand filters when operating in HDS mode, Geco is currently attempting to control the sulphate concentrations. This is done by liming the seepage waters prior to pumping them back to the pond. This reduces the iron concentration, increases the pH, and adds calcium which can form gypsum in the pond prior to treatment. The goal is to reduce the sulphate concentration to about 2,500 mg/L, and then it will be feasible to run the sand filters.

LESSONS LEARNED

There were a number of challenges throughout this project and each offered a valuable lesson:

Challenge 1 – Change in process during detailed engineering

Lesson: The process expert should be involved throughout the project. In this case, the people responsible knew only low density sludge treatment and made some decisions with just a quick consultation to a specialist. A miscommunication without written confirmation resulted in a new (inadvertent) process design.

Challenge 2 – Sludge viscosity problems

Lesson: This issue was resolved using a pilot plant. The key to making this new process work was to put the recycle to full throttle right from the start, but this would never have been tried in the full-scale plant for fear of creating a significant problem in the clarifier. A pilot plant allows trouble-shooters to try different things at little or no risk. As an added lesson, the process results showed that the representativity of the pilot work was excellent – different operating procedures resulted in similar sludge qualities in both plants.

Challenge 3 – Sand filter scaling

Lesson: Sand filtration was tested with low-strength acid mine drainage in the pilot plant and found to cause little scaling. By the time it was installed in full scale, the raw water had changed and scaling was an important issue. The lesson here is the importance of prediction. Expected changes in feed water quality, flowrates, or reagents should be considered prior to any important investment. In this case, it could have been predicted that the raw water concentrations would increase after closure and it is known that sand filtration does not work well with high sulphate concentrations.

CONCLUSIONS

1. When designing and building a new water treatment plant, a process expert should be involved throughout the entire procedure (design, detailed engineering, construction, and start-up).

2. Pilot plants can be used to optimise new processes or try process changes that would not be attempted in the full scale. Pilot plants can also show excellent representativity when properly designed.

3. Before building a treatment plant, significant effort should be spent on predicting future conditions and flowrates. These plants are typically built to operate for 20 years, but often treat a completely different feed within a few years. If prediction is difficult, built-in flexibility and over-sizing will likely save future operating and capital costs.

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