THE RE-EMERGENCE OF RESIN IN PULP WITH STRONG BASE RESINS AS A LOW COST, TECHNICALLY VIABLE PROCESS FOR THE RECOVERY OF URANIUM

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ABSTRACT

Data will be produced from two uranium feasibility studies that were conducted in 2008, both involving uranium recovery from leached slurry by resin-in-pulp (RIP) with strong base resins. Excellent metallurgical results were produced in RIP pilot plants in both studies, and the loaded resins were eluted with strong sulphuric acid and recycled to RIP in an integrated process. Resin losses by either abrasion or mechanical breakage were immeasurably low.

Metallurgical results from these two campaigns will be presented, along with an economic analysis of RIP versus the alternative process of CCD or filtration and solvent extraction.

INTRODUCTION

The recovery of uranium directly from pulp with strong base resins in RIP contactors was practiced commercially in the USA in the 1960s, but the process never established itself as a viable alternative to solid/liquid separation and solvent extraction. This was due mainly to the necessity to do a sand-slime separation partially resulting from the unavailability of coarse resin beads, inadequacies in the design of the RIP contactors and screens to separate resin from pulp, as well as problems with the strength and durability of resins in agitated pulps.

There have been significant advances in the last 30 years in each of these areas. All the commercial resin manufacturers are now producing very tough, durable, strong base resins, and are manufacturing them in a large bead size that is better suited to RIP applications. In addition, with the emergence and wide spread implementation of the carbon in pulp process in the gold industry in the 1980s, many innovative techniques for preparing leached slurries and separating carbon from pulp have become firmly established. These techniques have proven to be equally applicable to resin separation from pulps in RIP contactors.

Uranium recovery in the Western world today is generally accomplished by solvent extraction of the dissolved uranium from clarified liquors, which are produced by counter-current decantation and/or filtration, and resin in pulp is only just beginning to re-emerge. The use of resin-in-pulp is however quite widespread in the both the uranium and gold industries in the former Soviet Union countries and in China.

STRONG BASE ION-EXCHANGE RESINS

When leached in an oxidizing, acidic environment, uranium forms the anionic complex, $UO_2(SO_4)_3^{4-}$. In a strong base resin-in-pulp circuit, this dissolved uranyl sulphate anion in the leach slurry is adsorbed onto the anion exchange resin in an ion exchange reaction described by the following equation:

$$4RX + [UO_2(SO_4)_3]^{-4} \rightarrow R_4UO_2(SO_4)_3 + 4X^-$$
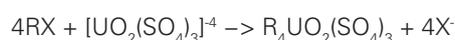
where R is the fixed quaternary amine cationic group and X is the mobile anion, which in the case of sulphuric acid elution systems is the sulphate anion.

Purolite A660/4759 resin has been used for several uranium projects at the SGS Minerals laboratory in Lakefield over the last couple of years. The resin is a gel-type, strong base resin with quaternary ammonium functional groups, which are fixed to a polystyrene backbone that has been cross-linked with divinylbenzene. The beads have the appearance of transparent, amber-coloured spheres with a mean particle size of 900 μm. This resin was chosen because if its strong affinity for the uranyl sulphate anion and its resistance to silica fouling relative to other strong base resins that were evaluated.

RIP CONTACTORS AND RELATED EQUIPMENT

Carousel type carbon-in-pulp (CIP) systems (such as the Kemix Pumpcell) are gaining favour in the gold industry. One of the benefits of Pumpcells compared to conventional CIP adsorption tanks and counter-current operating systems is that the carbon remains in its Pumpcell tank throughout its loading cycle. Carbon transfer in a countercurrent direction to the flow of pulp is achieved by moving the pulp feed and discharge points in the circuit rather than by physically pumping the carbon out of the tank. This has several claimed advantages:

- There is no back mixing of pulp as occurs in a traditional CIP plant when carbon and pulp are pumped up stream countercurrent to the flow of slurry. This increases the steady state concentration of gold in solution in each of the tanks, which improves the gold adsorption rate and the overall

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metallurgical efficiency (Dippenaar and Proudfoot). The net flow of pulp through the carbon screens is also reduced, which extends their operating life.

- Carbon breakage and make up requirements are significantly reduced by eliminating carbon/slurry pumping between stages.

- The carbon in each stage remains as a discrete batch throughout its loading cycle, with all the carbon loaded to approximately the same gold concentration. This also results in improved metallurgical efficiency compared to conventional CIP operations where carbon is transferred semi-continuously (Fleming and Nicol, 1984). In the latter case, there is a wide distribution of residence times and gold loading on the carbon in each stage.

- The Pumpcell screens are able to separate pulp and carbon at much higher carbon concentrations (up to 100 g/L carbon) than conventional CIP screens (up to 25 g/L carbon). This reduces the size of the adsorption tanks for a given pulp throughput, lowering capital costs and potentially decreasing the plant foot print.

Plant trials in South Africa have shown that the Pumpcell technology produces considerable cost savings and improved plant performance for gold operations (McArthur and Rogans, 2002). All the above advantages will also hold for resin in pulp applications, particularly for a uranium recovery operation in which it will be necessary to advance resin at a relatively high rate (compared to a gold plant).

RIP AND RESIN ELUTION TESTWORK

Two successful uranium pilot plants involving resin-in-pulp have recently been completed by SGS at their pilot plant facilities in Lakefield; the first for the Aurora Energy Inc. Michelin Uranium Project and the second for the Gold Fields Limited Driefontein Tailings Project.

AURORA ENERGY – MICHELIN PROJECT IN LABRADOR

Leach and RIP Pilot Plant

Michelin ore was pumped to the pilot plant at a pulp density of 56% and a solids flowrate of 2 kg/hr, through a circuit that consisted of 6 stages of sulphuric acid leaching, followed by 6 stages of RIP. Pulp residence time in the leach circuit was 36 hours and in the RIP circuit, 2.3 hours.

The flowsheet is shown schematically in Figure 1 and a photograph of the pilot plant is shown in Figure 2. The RIP contactors used were about 1 L in size, and each overflowed by gravity to the next stage. The RIP contactors were gently agitated with large diameter paddle-type impellers to avoid damaging the resin. Air spargers were used to apply a small flow of air behind the resin retention screens, to keep the screens from plugging and blocking the flow of slurry.
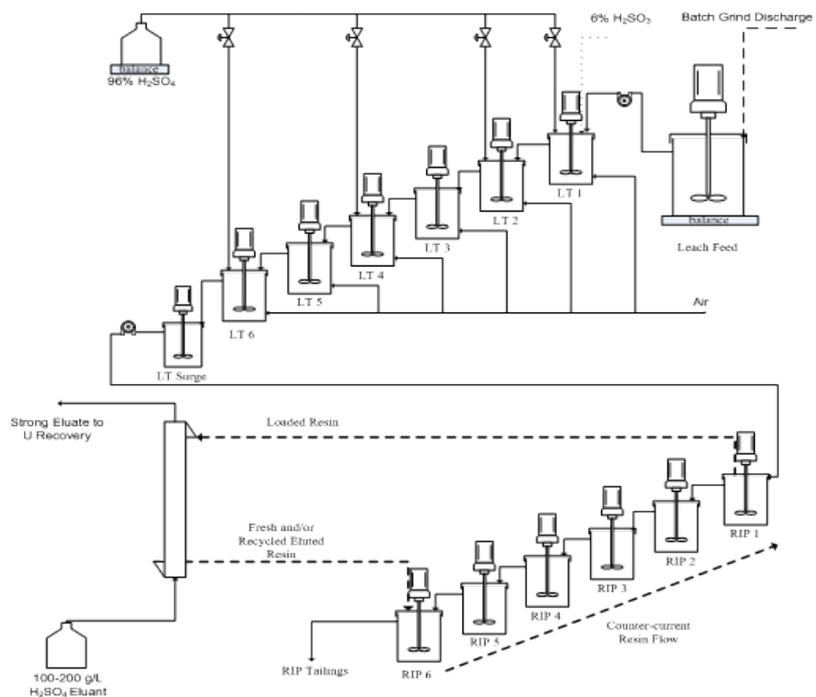
Figure 1: Leach / RIP Pilot Plant Flowsheet for Aurora Energy – Michelin Project

Figure 2: Michelin Pilot Plant – RIP Circuit

Uranium leaching[1] with sulphuric acid requires an oxidant as per the following reaction:

$$2UO_2 + 6H_2SO_4 + 3O_2 = 2UO_2(SO_4)_3 + 6H_2O$$
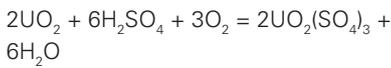
While oxidation in the above reaction is shown to be provided by oxygen, this is actually achieved with indirect oxidation with ferric sulphate. The resulting ferrous sulphate is re-oxidized to ferric using a chemical oxidant. Historically sodium chlorate, hydrogen peroxide, manganese dioxide, oxygen or Caro's acid have been used as oxidants in acid leach circuits. Auto-oxidation by O_2/SO_2 is well proven in the INCO cyanidation destruction process and could be used in uranium leach circuits to provide inexpensive oxidation. This was extensively tested for the Michelin project.

Although sulphur dioxide by itself is a reductant, coupled with oxygen (or air) it has strong oxidizing potential, the overall reaction with ferrous iron is:

$$SO_2 + O_2 + 2Fe^{2+} = 2Fe^{3+} + SO_4^{2-}$$
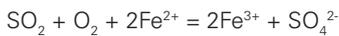
The main advantage of SO_2/air is that it is relatively cheap compared to other oxidations if a source of SO_2 (i.e. an acid plant) is readily available – as is the case for the proposed Michelin Project. The rate of oxidation however is relatively slow by this method since it is dependant on the low solubility of oxygen, though due to the long residence times common in uranium leach circuits SO_2/air could still be feasible. (Ho and Quan, 2007)

In the Michelin pilot plant leach circuit air was continuously sparged through each of the 6 leach tanks, and an aqueous solution of sulphur dioxide (sulphurous acid, H_2SO_3 containing 6% SO_2) was added to multiple tanks using ORP control loops to maintain an ORP (vs. Ag|AgCl sat'd KCl electrode) of 480-500 mV. This approach worked very well during the pilot plant and the oxidation was easily controlled to the setpoint to achieve maximum uranium dissolution.

Highlights from the longest of the four pilot plant campaigns (3 weeks) that were run for Aurora Energy are listed in Table 1 (all data are averages for the entire campaign).

The feed to RIP contained about 840 mg/L uranium in solution, and this

Ore Throughput	1.96	kg/h
	872	kg processed
Leach Feed	58%	solids
Flowrate	35	mL/min
Pulp Density	1585	g/L
Leach Temperature	20	°C
H_2SO_4 Addition	29	kg/t
SO_2 Addition	12.5	g/t (as H_2SO_3 w 1.2 vvm air)
Feed Grade	0.071%	U
Leach Extraction	86.8	
PLS Grade	840	mg/L U
RIP Recovery	99.70%	
RIP Barren	3	mg/L U
Total Operations	444	h
Resin Cycles	8	
Resin Losses	<5%	

Table 1: Michelin Leach / RIP Operating Highlights
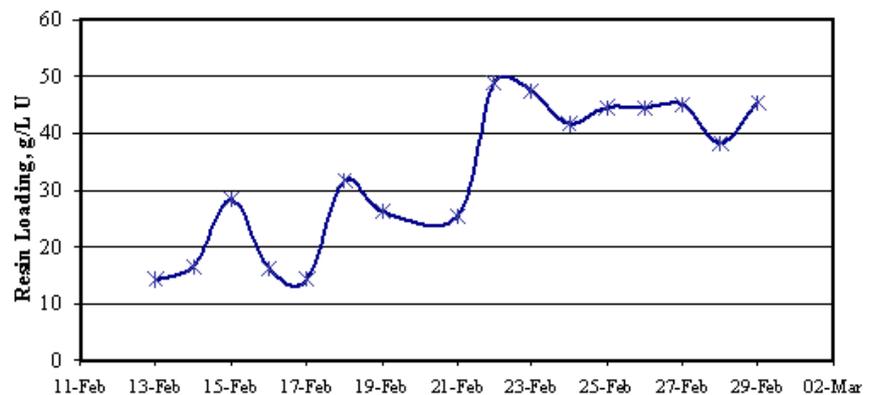
Figure 3: Uranium Loading Trend
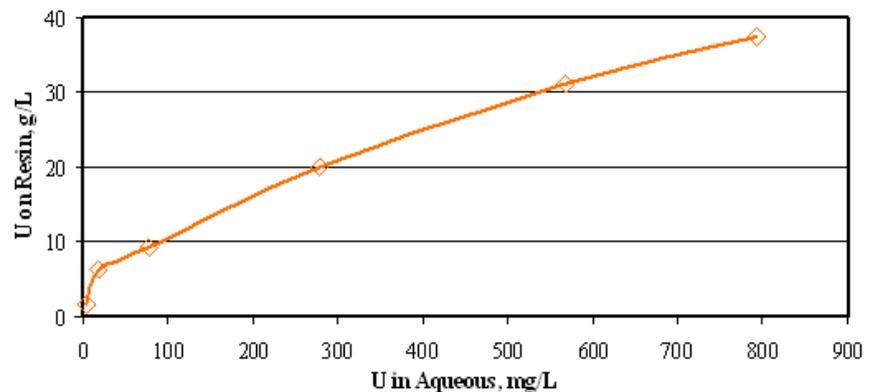
Figure 4: Resin Loading Isotherm

[1] *Uranium ores typically contain uranium in both the hexavalent and tetravalent forms, hexavalent uranium is readily soluble in acid while the tetravalent need first be oxidized to become soluble.*

was consistently reduced to ~ 3 mg/L uranium in solution in the RIP tailings slurry (99.7% extraction efficiency).

The loading of uranium on the resin that was advanced from the first RIP tank over the 3 week campaign is shown in Figure 3. The plant was started up conservatively, with a fairly rapid resin advance rate to demonstrate good uranium extraction efficiency, and was then slowed (while keeping the pulp flowrate constant) after the first week of operation. As a result, uranium on the resin increased from 20-30 g/L the first week start to the design target of 40-50 g/L at the end of the run. This was achieved with no loss in uranium extraction efficiency, which remained >99% throughout the campaign.

The average steady state concentrations of uranium on the resin and in solution in each of the 6 stages towards the end of the campaign are plotted in Figure 4. The highest loading (and highest concentration of uranium in solution) corresponds to the lead stage, which contains the most loaded resin, while the tail stage (from which comes the barren pulp) contains essentially fresh or eluted resin, with minimal uranium loading. As the resin is advanced (batch-wise) up the circuit, counter current to the pulp flow it sees gradually higher concentrates of uranium in solution and as a result adsorbs more uranium.

The kinetics of uranyl sulphate loading on a strong base resin are fast (attaining equilibrium in 2 to 4 hours), and with a 6 hour contact time per stage, the loading is expected to be very close to equilibrium in each stage of this RIP circuit. Therefore, the curve in Figure 4 will be very close to an equilibrium isotherm for uranium in the Michelin ore on A660 resin.

Resin Elution

Sulphuric acid was used for the elution of uranium from the loaded resin during the RIP campaigns. Uranium loads onto a strong base resin (R^+X^-) as the uranyl sulphate anion, according to the following stoichiometry:

$$4RX + [UO_2(SO_4)_3]^{-4} \longrightarrow R_4UO_2(SO_4)_3 + 4X^-$$
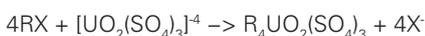
The uranium is then displaced in the elution reaction by a strong concentration of the counter-ion X^-, which reverses the above reaction. Sulphuric acid in the form

of the bisulphate anion (HSO_4^-) has only a modest affinity for anion exchange resins, so a strong solution of sulphuric acid (> 100 g/L) is needed to reverse the above reaction.

$$R_4UO_2(SO_4)_3 + 4HSO_4^- \longrightarrow 4RHSO_4 + [UO_2(SO_4)_3]^{4-}$$
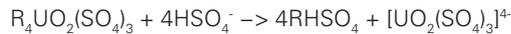
The loaded resin from the Michelin RIP pilot plant was eluted in a column at 45°C by pumping 16 bed volumes of 150 g/L sulphuric acid solution upflow through the column at a flowrate of one bed volumes per hour. The elution was performed as a split elution, with weak eluate from the previous elution making up the initial 6 bed volumes of eluant. The elution efficiency under these conditions was > 97%, and the eluted resin averaged 1.2 g/L residual uranium. An average elution curve, for the final 3 elutions performed during the campaign is shown in Figure 5.[2]
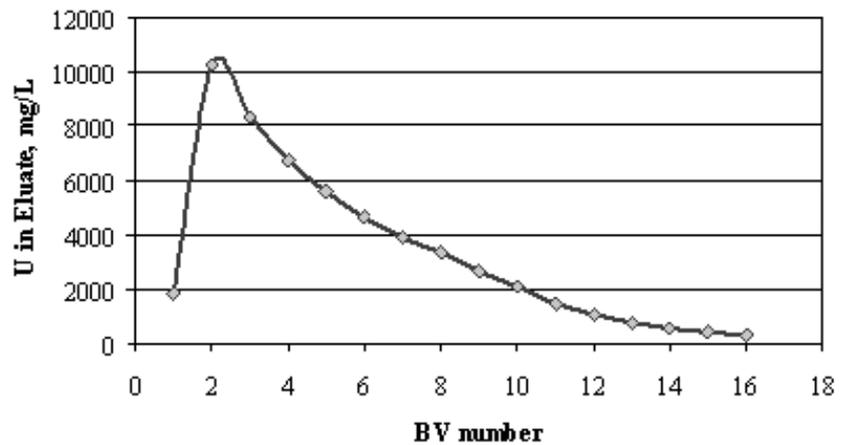
Figure 5: Michelin PP Elution Curve – Average of Final 3 Elutions

Eluted resin was recycled to the RIP circuit for most of the Michelin campaign and the number of loading/elution cycles was estimated at about 8 based on circuit volumes and resin advance rates.
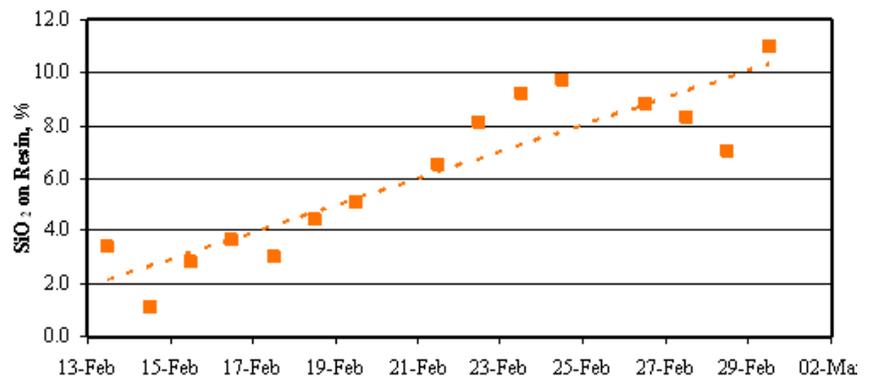
Figure 6: SiO$_2$ Fouling

Silica was found to build up on the resin during the pilot plant operation, as illustrated in Figure 6. The buildup of silica over this range of 2 to 10% on the resin had no detrimental impact on uranium loading during the pilot plant operation (as seen in Figure 3) or on the overall uranium extraction efficiency, with uranium in the barren

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[2] *The data in Figure 5 was generated from assays obtained with a portable XRF analyzer, not by accredited assay techniques in the SGS analytical lab. The portable XRF unit was used extensively during the Michelin pilot plant, particularly the elution circuit and several calibration curves were generated using actual ICP uranium assays and the data in Figure 5 is believed to be a good representation.*

solution remaining < 3 mg/L until the end of the campaign. However, if silica continued to accumulate over more extended periods, its presence on the resin would almost certainly impact either the kinetics of uranium loading or the loading capacity (or both) sooner or later. Traditionally, this would be handled by treating a bleed stream of eluted resin with a regenerant solution of sodium hydroxide, but A660/4759 is reported by the manufacturers to be unstable in alkaline solution, and a strongly alkaline treatment could degrade the resin's functional groups. Purolite is working on an alternative regeneration treatment to deal with silica. It should be noted though that the A660/4759 has very good selectivity against silica and loads considerably less silica compared to previous generations of strong base resins.

While resin losses were difficult to quantify, the volumes of resin recovered during each advance cycle and on a fine tramp screen used to separate any broken resin beads from the RIP tailings slurry revealed little to no broken resin. In addition to this, observing samples of the resin from the circuit under an optical microscope revealed no significant breakage of the resin.

GOLD FIELDS LIMITED – DRIEFONTEIN TAILINGS

Background

The second project involved flotation treatment of historic gold plant tailings from South Africa for Gold Fields Ltd., to produce a rougher float concentrate. This concentrate was then cleaned to produce two products, a fairly low grade pyrite concentrate and cleaner scavenger tailings. The low grade cleaner concentrate was subjected to pressure oxidation at 200°C under an oxygen partial pressure of 100 psi for 1 to 2 hours. This oxidized the sulphide minerals to sulphuric acid and ferric sulphate, and dissolution of uranium was essentially 100% complete in the autoclave. In addition, the refractory gold within the sulphide concentrate was liberated for recovery by subsequent cyanidation.

The discharge slurry from the autoclave was then combined with the cleaner scavenger tailings in a hot (50 to 90°C),
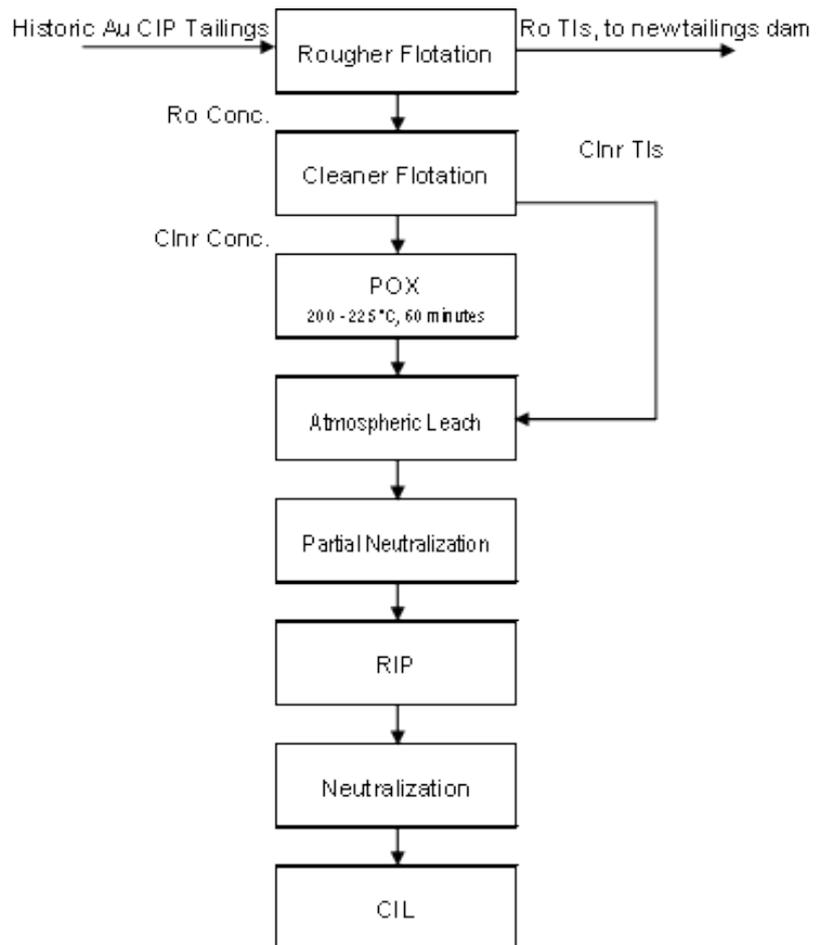
Figure 7: Block Flow Diagram – Gold Fields Ltd HTO Driefontein Project

atmospheric leach, which was designed to dissolve any acid-soluble uranium in the cleaner tails by utilizing the ferric iron and acid generated in the autoclave. The combined leached slurry was then partially neutralized with limestone to ~ pH 3 to precipitate gypsum and ferric hydroxide, and then treated in a 6 stage RIP circuit to recover uranium. The block flow diagram of the conceptual process is illustrated in Figure 7. Details of the performance of the autoclave and atmospheric acid leach circuits are presented elsewhere in the proceedings of this conference (Fleming et al, 2009).

RIP Pilot Plant

The partial neutralization ahead of RIP was done to reduce the amount of iron (and also silica) in the leach solution proceeding to RIP. Anionic ferric sulphate complexes load on strong base resins, and in the case of the Driefontein leach pulp iron was found to load up to 55 – 70 kg/t. It would also be highly desirable to reduce iron loading on the resin, as the large amount of iron on the

resin can not only occupy more than half the functional groups on the resin (at the expense of uranium), but will also create difficulties in the downstream processing of the resin eluate to recover a pure uranium product. Therefore, the effect of raising the loading pH from 2 to 3 and precipitating most of the ferric from solution before resin loading was examined. The dramatic impact of this change is illustrated in Figure 8. Most of this improvement probably stemmed from the reduced competition from ferric ions, but part would have been due to the fact that uranium loads better at pH 3 than at pH 2.

The uranium to ferric iron ratio in the PLS for the Gold Fields project was about 1:50, compared to 1:4 for the Michelin PLS. Since strong base resins are quite selective for uranium over ferric iron, the need for ferric iron removal ahead of the RIP circuit in the Gold Fields project was much greater; though some benefits would be realized if a similar approach was applied to the Michelin project. In

fact, a different approach was tested during the Michelin pilot plant whereby the leach discharge slurry was sparged with sulphur dioxide ahead of the RIP circuit to reduce ferric iron to ferrous in an attempt to increase uranium loading; however that operation was only run for a short period due to issues related to ventilation in the pilot plant.

The kinetics of uranium loading observed during the Gold Fields testwork are illustrated in Figure 9, from which it can be seen that the rate of uranium loading is fast, with a $t_{1/2}$ for the reaction of about 1 hour. On this basis, it can be estimated that the loading of uranium on the resin in each stage of a multi-stage RIP plant will be close to equilibrium (>90%) if the resin spends at least 4 to 6 hours in each stage. In fact, the pilot plant was designed with a residence time of 12 hours per stage for the resin, more for operator convenience than anything else.

The combined leach slurry was pumped to the neutralization and RIP circuits at an initial target flowrate of 7.5 L/hr. Uranium was extracted from the solution phase of the leach slurry in 6 RIP tanks in series, each with a volume of about 7.5 L. The resin was advanced counter-current to the direction of slurry flow every 12 hours, by advancing all the cells in the RIP train. The loaded resin in the lead cell was recovered by discharging the contents of the tank over a 35 mesh screen, which retained the resin and allowed the slurry to flow through and back to the circuit. The washed, loaded resin was transferred to the elution section, while the slurry was pumped slowly back into the RIP circuit by including it in the feed to the first RIP tank.

The circuit started up with 250 mL of wet settled resin in each contactor, and resin was advanced every 12 hours with a pulp flow of ~ 7.5 L/h. This resulted in a pulp to resin flow ratio of ~ 350, which would have loaded the resin to ~ 50 g/L uranium. Uranium extraction efficiency was excellent throughout this period, with the RIP pilot plant producing barren solution losses of < 1 mg/L U. Once the iron and silica levels in the feed to RIP were lowered after the first week of operation (by making improvements to the partial neutralization circuit), it was
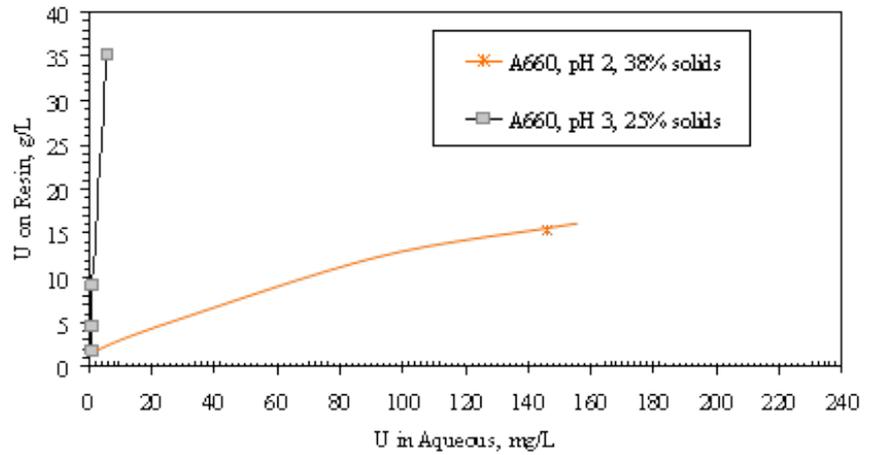
Figure 8: Effect of RIP Feed Partial Neutralization
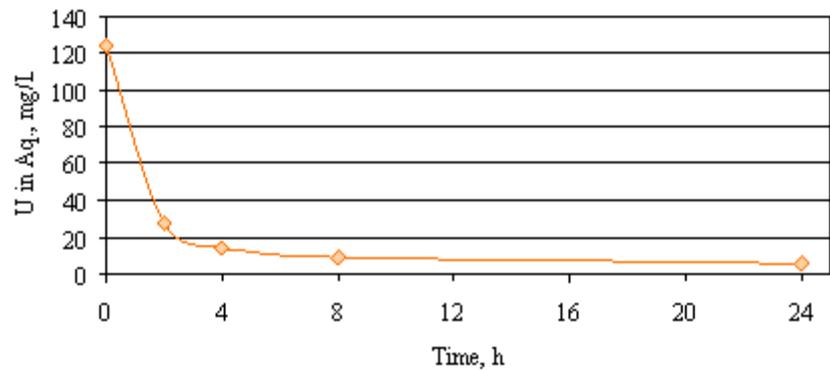
Figure 9: Uranium Loading Kinetics on A660 SBA resin (3 mL/L resin, pH 3, 25% solids)
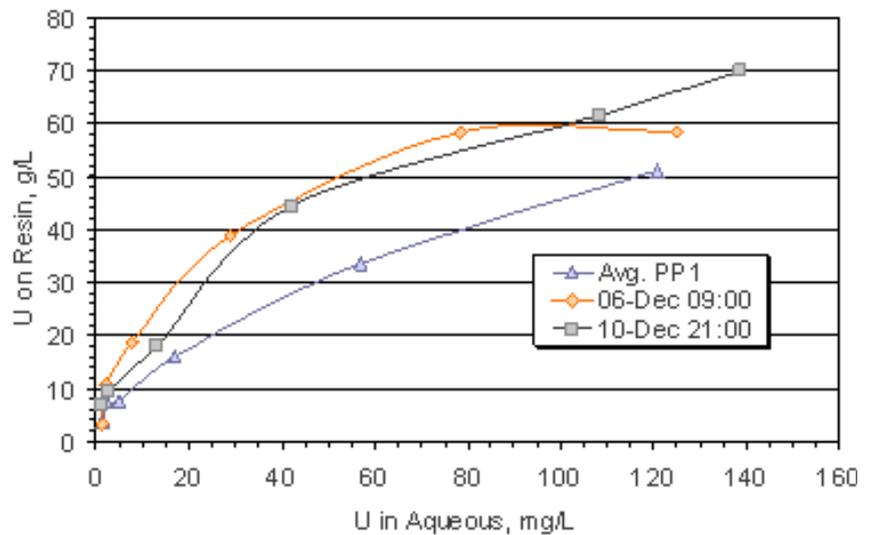
Figure 10: Resin Loading Isotherm

realized that higher uranium loading on the resin could be achieved. Therefore, the amount of resin in each stage was reduced to 200 mL, which increased the pulp to resin flow ratio ~ 450, and

increased the target uranium loading on the resin in this phase to ~ 60 g/L U.

These conditions remained unchanged for another week at which stage the amount of resin in each stage was

further reduced to 180 mL, the pulp flowrate was increased to 10 L/hr, and the frequency of resin advance was increased from 12 hourly to 9 hourly. These changes increased the pulp to resin flow ratio to 500 and the target uranium loading on the resin to 70 g/L. This loading was achieved quite consistently during the last week of piloting.

Steady state concentrations of uranium on the resin and in solution at the start of the campaign (PP1) as well as at the mid point (Dec 6th) and towards the end of the campaign (Dec 10th) are shown in Figure 10. The uranium loading isotherm was depressed somewhat at the start of the campaign by competition from ferric sulphate anions on the resin, but once the pre-oxidation and neutralization section had been optimized for maximum iron removal, the uranium loading characteristics improved quite significantly.

The uranium concentration in solution for each RIP stage is plotted in Figure 11, based on the average solution profiles (taken twice daily) for the two pulp flowrates used in the RIP campaign (~7.2 L/h during PP1 and first half of PP2, ~10.2 L/h during second half of PP2). The data show low uranium barrens by the 5th loading stage under both conditions. However, it can be seen that the uranium on the resin in Stage 1 would have been close to the maximum (saturation) loading when the pulp flowrate was 10.2 L/hr during PP2, as evidenced by the fact that the concentration of U in solution didn't change much in stage 1 (~5%). These conditions were therefore close to optimum, with maximum U loading on the resin in Stage 1 (~70 g/L) and minimum U in solution in Stage 6 (<1 mg/L).

Resin Elution

Sulphuric acid was used for the elution of uranium from the loaded resin during the RIP campaigns. The effects of variables such as eluate flowrate, direction of eluate flow (up or downflow through the column) and temperature were examined.

The most important operating variable was found to be eluate flowrate, and the effect of flowrate (upflow) is illustrated
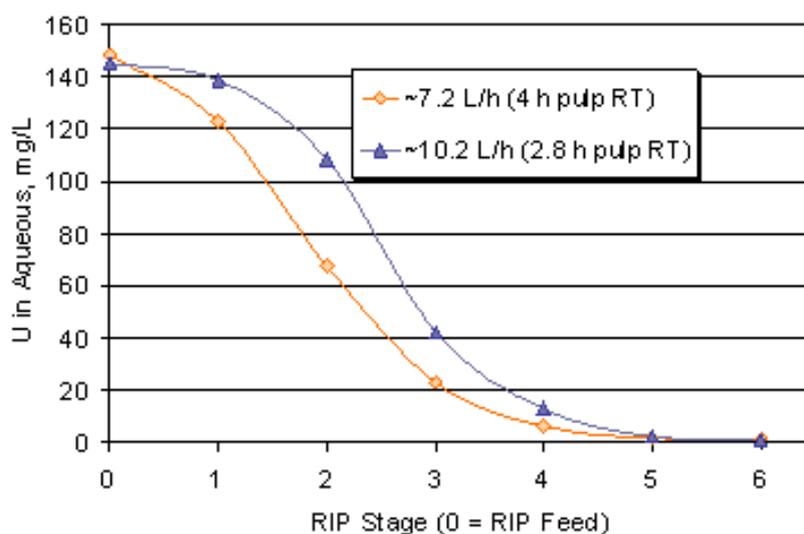
Figure 11: Uranium Loading by Stage

in Figure 12. Good elution kinetics and elution efficiency are characterized by an elution curve with a sharp, high peak, dropping rapidly to zero. Poor elution is characterized by a flat rounded peak, with uranium persisting in the eluate for many bed volumes.

It is noteworthy that the elution efficiency improved when the flowrate was decreased from 2 BV/h to 1.5 BV/h to 1 BV/h (BV/h being solution flow in equivalent bed volumes of resin per hour), but then deteriorated sharply when the flowrate was reduced further to 0.5 BV/h and 0.25 BV/h. To account for the fact that elution efficiency improved with decreasing flowrate, the rate of elution must have been controlled by diffusion of the large uranyl sulphate anion in the resin pores. If this was the case, the elution efficiency should have continued improving with decreasing flowrate until constant elution efficiency was achieved. This would be the optimum flowrate. But the elution efficiency should not have deteriorated at very low flowrates, as happened here at 0.5 BV/h and 0.25 BV/h. It is believed that this occurred because of gravitational back-mixing of the concentrated uranyl sulphate eluate when solution was pumped upflow through the column at very slow flowrates. This back-mixing of the concentrated, heavy uranyl sulphate anion resulted in the flat uranium concentration profiles seen for the two slower flowrates in Figure 12.

It was therefore reasoned that the only way to take advantage of the superior elution efficiency that should be achieved at very low flowrates (less than 1 BV/h), would be to pump the eluate solution downflow through the elution column, so that gravitational forces and solution flow would be in the same direction, thereby preventing back-mixing. This became the standard elution methodology for the duration of the pilot plant, and the best elution profiles were achieved when a 200 g/L sulphuric acid solution (at 60°C) was dripped slowly down through a bed of resin at a flowrate of 0.25 BV/hr.

The ferric sulphate anion [$Fe(SO_4)_2^{2-}$] does not load onto anion exchange resins as strongly as uranyl sulphate, and would therefore be expected to strip off the resin more rapidly. This was confirmed in the test work, where the ferric was all stripped in the first ~25% of the eluate volume. But, sulphuric acid was unable to produce a "clean" separation of uranium and iron in the strip solution, and all increments of eluate contained both metals. Good solution purification and iron removal ahead of uranium precipitation is difficult to achieve and iron inevitably ends up in final uranium product. A method of selectively eluting iron from the strong base resin was therefore sought during the Driefontein pilot plant campaign at SGS in Lakefield.

The fact that ferrous iron does not form anionic complexes (and thus would not load on strong base anionic resins)

was exploited in a reductive first stage elution of the loaded resin. An aqueous solution of sulphur dioxide (sulphurous acid, 6% SO_2) was passed through the resin bed, reducing ferric on the resin to ferrous and very effectively stripping iron from the resin. Since sulphurous acid is a weak acid, the amount of uranium stripped from the resin during this step was minimal. This methodology was applied in the later stages of the pilot plant campaign.

The overall resin elution scheme consisted of first eluting with 6 bed volumes of a 6% aqueous sulphur dioxide solution (sulphurous acid) at a flowrate of 1 BV/h, to strip the iron selectively from the resin. This was followed by uranium elution, which comprised dripping 6 bed volumes of 200 g/L sulphuric acid downflow through the elution column at 0.25 BV/h. The two stages of elution took a combined 30 hours. The uranium elution was operated as a split elution, with only the most concentrated first half of the eluate (bed volumes 0.5 to 3.5) advancing to the uranium recovery process, whilst the second half of the eluate (the first 0.5 bed volumes plus bed volumes 3.5 to 6) was recycled to the next elution. On a once-through basis, 75 to 80% of the uranium on the loaded resin reported to the most concentrated 3 bed volumes of eluate, at an average uranium concentration of ~ 22 g/L. About 15% reported to the 3 bed volumes of eluate that were recycled to the next elution, 2 to 3% reported to the iron rich sulphurous acid solution and 1 to 2 reported to the final 4 bed volumes of de-ionized water wash. Both of these solutions would be recycled in the process, somewhere upstream of RIP, so the uranium would not be lost. Roughly 5% of the loaded uranium remained on the eluted resin that was recycled to the last stage of RIP. This distribution is illustrated in Figure 13.

The theoretical capacity of the A660/4759 strong base resin for anions is reported as ~1.3 equivalents per litre of resin. Therefore, the stoichiometric amount of sulphuric acid that would be needed to replace all the complexed metal anions on the loaded resin with bisulphate anions is about 130 kg/m3. The amount of acid actually required to
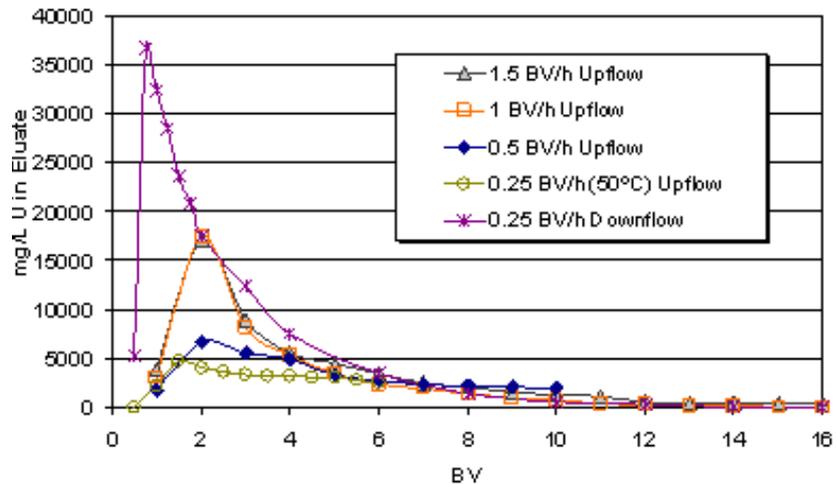
Figure 12: Pilot Plant Elution – Variable Flowrate (200 g/L H_2SO_4, 60°C)
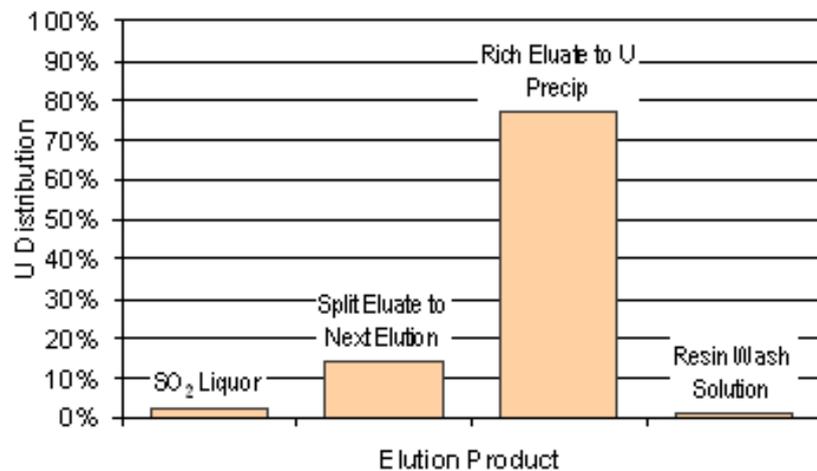
Figure 13: Uranium Distribution for Optimized Elution Procedure

strip the uranium from the resin under the optimum conditions established during the pilot plant was 600 kg/m3 if the split elution procedure is adopted (3 bed volumes of 200 g/L acid), which is 4 times the stoichiometric amount. This large excess is needed to drive the elution reaction, which is quite unfavourable because of the great affinity of the resin for the uranyl sulphate anion. Acid consumption could be reduced further by treating the strip liquor (~22 g/L U and ~200 g/L sulphuric acid) by a process such as membrane separation or acid retardation technology, both of which have the potential to increase the uranium to acid ratio by a factor of at least 2.

CAPEX & OPEX COMPARISON

In 2007, Aurora Energy Inc. conducted an internal trade-off study to determine

the optimum metallurgical process for treating ore from the Michelin and Jacques Lake uranium deposits in Labrador, Canada. The trade-off study only considered processing costs and was done strictly to focus the metallurgical testwork that was continuing at that time. Costs were generated for belt filtration – solvent extraction (SX), counter-current-decantation (CCD) – SX and RIP flowsheets.

On September 8, 2009, Aurora Energy issued a press release discussing a Preliminary Economic Assessment (PEA) of the Michelin project covering development of the Michelin project. The PEA contemplated a 10,000 t/d treatment rate and RIP processing of the ore. Highlights from the press release are provided in Table 2.

The earlier Scoping Study costs have been applied to the PEA data to arrive at the cost comparisons presented below.

Total U$_3$O$_8$ produced	96.9 million pounds
Average grade (U$_3$O$_8$)	0.09%
Average annual production of U$_3$O$_8$	5.7 million pounds
Life of mine	17 years
IRR (pre-tax)	19.40%
NPV (pre-tax)	US$914 million
Discount rate	8%
Pay-back period	4.7 years
Net cumulative cash flow	US$3.038 billion
Direct cash costs	US$28.57/lb U$_3O_8$
Initial capital expenditures including mine, mill, infrastructure (port, road, grid power, etc.) tailings management, environmental, owners costs, decommissioning and engineering, procurement, construction management (including $132 million contingency)	US$984 million
Metallurgical recovery rate	87.50%
Long term exchange rate	C$1 = US$0.88
Long term uranium price	US$75/lb U$_3O_8$

Table 2: Basic data for Michelin project from September 8, 2009 press release

Note: Rainbow, Nash, Gear, and Inda resources not included in the study. This PEA is preliminary in nature as it includes inferred mineral resources that are considered too speculative geologically to have the economic considerations applied to them that would enable them to be categorized as mineral reserves at this time and as such there is no certainty that the preliminary assessment and economics set forth in the PEA will be realized. Mineral resources that are not mineral reserves do not have demonstrated economic viability.

PLANT DESIGNS

The Michelin uranium project is located in a remote part of Labrador with only seasonal access. Because of the remote location, the project will be provided with on-site accommodation and employees will be transported into and out of the site on a rotational basis. A port on the seacoast and extensive reagent and supply storage facilities are needed.

Testwork completed prior to the 2007 process trade-off study established that an acid leach plant would be appropriate and established the major design parameters for the three process options.

All process options include a semi-autogenous grinding (SAG) mill and ball mill to prepare a leach feed at 80% passing 90 µm. The ground ore is thickened then leached at about 250°C for 48 h in agitated tanks with sulphuric acid produced in an on-site, sulphur-burning plant. Oxidation is obtained using air and SO$_2$.

In the belt filtration plant alternative, leached pulp is filtered using five installed 130 m² filters. Filtrate is clarified then processed in a normal SX circuit for uranium extraction. Both pulsed columns and more conventional mixer-settlers were considered and the conventional units carried through to the process trade-off study. In the proposed plant, loaded solvent is stripped with strong sulphuric acid, the excess acid neutralized with limestone followed by yellowcake precipitation, drying, and packing.

In the proposed CCD option, pregnant solution is separated from tailings in a series of six 36 m diameter thickeners. Pregnant solution is clarified and then treated by solvent extraction as in the belt filtration step.

In the RIP option, the pregnant solution is partially neutralized to reduce the possibility of silica fouling and reduce the free acid concentration ahead of ion exchange resin loading. A Kemix carousel was included in the design for loading resin with Delkor-type linear screens ahead and after the RIP loading circuit to control trash and catch any fugitive resin beads. Loaded resin was eluted with hot sulphuric acid in a multi-batch system designed to minimize the acid to uranium ratio. Eluate was processed in a similar way to the SX strip solution using limestone for excess acid removal followed by product precipitation.

In all options, tailings are neutralized with finely ground limestone then lime before sending to either a backfill preparation plant or a secure tailings storage facility. Lime is obtained from an on-site, high-efficiency lime kiln.

URANIUM RECOVERY

The belt filter – SX and the CCD – SX plants will lose uranium as soluble losses from the liquid-solid separation step and in the SX raffinate – albeit some of the raffinate could be recycled. Soluble losses from the liquid-solid separation step might amount to between 0.5 and 1% of the solubilized uranium depending on the wash ratio, number of separation stages, etc.

In contrast to the belt filter and CCD cases, the RIP option does not include a liquid-solid separation step and uranium losses to barren slurry are all that need be considered. Consequently the overall recovery from the RIP circuit could be in the order of 0.5% higher than that from the other circuits. In the base case cash flow considered below, the RIP option is not given any credit for higher recovery but this is examined in an alternative case.

CONCEPTUAL PLANT CAPITAL COSTS

Equipment vendors were approached for budget costs for major equipment packages and individual pieces of process equipment. The total direct capital cost for the project was then obtained by multiplying the costs for the major equipment by a factor of between 2 and 3 to account for site preparation, foundations, installation, piping,

electrical, and buildings. A factor of 1.3 was applied to obtain installed costs, including site preparation, foundations, electrical supply, etc. for the acid plant, lime kiln, and solvent extraction plants which were quoted as lump sum costs.

Engineering, procurement, and construction management (EPCM) costs were factored from the total direct capital cost on the basis of 15%. Indirect construction costs have been taken as 15% of the direct costs while contingency has been added at 20%. The resulting capital cost is summarized in Table 3.

PROCESS PLANT OPERATING COSTS

Process plant operating costs for ore from the Michelin deposit are summarized below in Table 4 and discussed later in this section. It should be noted that Jacques Lake ore processing costs are slightly higher than Michelin because it consumes more acid but, otherwise, the plant operating costs ores are very similar.

Various aspects of the operating costs are discussed below.

The labour complements for the belt filter, CCD, and RIP plants are estimated as 104, 98, and 91 respectively. Belt filter plants pose relatively intensive operating and maintenance labour requirements. In contrast, the RIP plant is simpler to operate and maintain and labour requirements are lower.

Michelin ore is very soft so comminution energy requirements are very modest at about 15 kWh/t. The belt filter plant includes vacuum pumps and other high energy demand equipment resulting in a higher overall electrical power demand. The total electrical energy demand of the belt filter, CCD, and RIP plants are estimated as 31, 28, and 28 kWh/t respectively.

The costs for reagents and supplies are similar for all process options. The RIP option requires slightly more strong acid to elute uranium from the resin compared to the acid demand for stripping loaded solvent. The resin replacement rate in the RIP option was taken as a very conservative 50% of inventory per year. This equates to $0.45/t ore or 6% of the total process operating cost for the RIP option. Actual

resin replacement rates are expected to be lower – perhaps 20 to 30% of inventory per year.

DISCOUNTED CASH FLOW COMPARISON

The above data show that the RIP plant offers the lowest capital cost, lowest operating cost, and the clear probability of a higher uranium recovery. Under these circumstances, RIP will always show an economic advantage over the alternative processes.

The discounted cash flow (DCF) calculations used in the PEA were adjusted to measure the effects on

economic parameters of the alternative flowsheets. The impact of an increase in uranium recovery was also evaluated. The economic parameters evaluated were the Net Present Value (NPV) at an 8% discount rate, the payback period, and the Internal Rate of Return without taxes (IRR). The results of these studies are summarized in Table 5.

The RIP circuit offers an increase in the NPV of the project of US$56 million and a full 1% increase in the IRR over the filter-SX option. The NPV and IRR are increased to US$74 million and 1.2% respectively if recovery is improved by

PROJECT AREA	FILTER+SX	CCD+SX	RIP
Site preparation	Inc.	Inc	Inc.
Crushing	8,739	8,739	8,739
Grinding	28,380	28,380	28,380
Neutral and tailings thickeners	3,809	3,809	7,618
Acid leaching and surge tanks	18,773	18,773	18,773
Belt filtration	35,075	-	-
CCD circuit	-	34,845	-
Solution clarification	4,945	4,945	-
SX columns and strip MS	27,025	27,025	-
Initial SX solvent (500m3 solvent)	750	750	-
RIP trash, safety, and wash screens	-	-	4,805
RIP load cells and strip columns	-	-	19,717
Initial IX resin (450m3 resin)	-	-	3,604
Slurry neutralization	15,493	15,493	15,493
Tailings decant treatment plant	2,508	2,508	2,508
Tailings paste preparation	13,434	13,434	13,434
Uranium precip., drying and packing	9,993	9,993	9,993
Limestone grinding	4,782	4,782	4,782
Acid plant (100,000 t/a capacity)	51,573	51,573	51,573
Reagent storage and preparation	11,286	11,286	11,286
Lime kiln, lime handling and slaking	10,463	10,463	10,463
Plant mobile equipment (inc. crane)	1,500	1,500	1,500
Assay and metallurgy laboratories	1,500	1,500	1,500
Total direct process costs – Can$'000's	250,028	249,798	214,168
Total direct process costs – US$'000's	218,778	218,577	187,400
INDIRECT COSTS			
EPCM (15% direct)	37,504	37,470	32,125
Construction indirects (15% direct)	37,504	37,470	32,125
Contingency (20% direct)	50,006	49,960	42,834
Total indirect process costs	125,014	124,899	107,084
Grand total process capital cost	375,042	374,697	321,252
Percentage of Filter+SX case	100	100	86

Table 3: Detailed capital cost estimates, Can$'000's

0.5%. The CCD-SX case also offers improvements over the filter-SX case but also falls well short of the RIP option.

The economic advantages to RIP are significant and help explain the widespread interest in this technology for uranium recovery, as well as the recovery of other metals such as gold from cyanide leached pulps and base metals from laterites.

CONCLUSIONS

Resin-in-pulp using strong base resins has been proven to be a technically viable and robust process for recovery of uranium during two different pilot plant test programs at SGS in Lakefield. RIP offers important economic advantages over more traditional process options such as filtration-SX and CCD-SX. While the technology is still considered unproven in the Western world, it has been and is used in FSU countries as well China. With continued improvement and confidence in commercial resins and RIP equipment within the industry, uranium processing by RIP is now becoming a significant option for today's uranium development projects.

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AREA	FLOWSHEET OPTION – COSTS $/T		
	FILTER+SX	**CCD+SX**	**RIP**
Labour	3.78	3.57	3.32
Electrical power	1.44	1.29	1.27
Process reagents	8.17	7.94	8.20
Maintenance parts	1.58	1.59	1.35
Mill overhead	0.16	0.16	0.16
Tailings allowance	1.00	1.00	1.00
Grand total – Can $/t	16.13	15.55	15.30
Percentage of filter-SX cost	100	96	95

Table 4: Summary of Michelin ore process operating costs, Can$/t

CASE	NPV AT 8% US$'000'S	PAYBACK PERIOD YEARS	IRR %
Filter-SX flowsheet	858	4.85	18.4%
CCD-SX flowsheet	871	4.82	18.6%
RIP flowsheet	914	4.66	19.4%
RIP flowsheet, recovery increased to 88%	931	4.62	19.6%
RIP flowsheet, recovery increased to 88.5%	949	4.59	19.7%

Table 5: DCF parameters for process and recovery options

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