

Insights From Simulation of a CIL/CIP Circuit

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ABSTRACT

A model was developed of a CIL/CIP circuit for recovery of gold and silver. The model uses a simple two rate representation of the leaching kinetics. Parameters for the leaching kinetics may be simply obtained by sampling the feed to a circuit and the tailings from each tank and assaying the solids for both gold and silver to obtain the assay profiles of the solids down the tanks. The parameters in such a leaching model would be expected to be a function of the ore type, the grind, whether oxygen was used and the general chemical environment. These parameters are primary inputs to the model for any simulation.

Standard power law relationships are used to express the practical equilibrium between the gold and silver cyanides adsorbed on the active carbon and the tenor of the solutions. The parameters for this may be derived from solution and carbon assays for the same set of samples.

The model is constructed by writing a mass balance for gold and silver in and out of each tank in the train and for the circuit overall. Other inputs to a simulation are the feed rate of ore, per cent solids of the slurry in the tanks, gold and silver loadings of the input 'barren' carbon and the rate of movement of carbon through the circuit counter-current to the slurry. With appropriate practical parameters the simulation is able to reproduce typical tank profiles.

The model was used in a series of simulations to explore the influence of:

- leaching parameters,
- variation in ore feed rate and per cent solids of slurry,
- variation in rate of counter-current movement of active carbon,
- the number of tanks,
- tank size,
- tankage configuration,
- using some tanks for leaching only, and
- the impact of a thickener recycling a percentage of tailings solution to the head of the circuit.

The results of these simulations highlight factors which need to be considered in the design and optimisation of a CIL/CIP circuit and provide an insight into deficiencies in performance and the best way to counteract them.

INTRODUCTION

The leaching of gold and silver by cyanide and counter-current adsorption of the cyanide complexes on activated carbon is a well established technology. In essence the process is simple. Values are leached from finely ground ore by cyanide under alkaline conditions in a series of well stirred tanks. Freshly activated carbon, with a much larger particle size than the ore, is introduced into the final tank. Typically the carbon is retained in any tank by suitable screens and is intermittently progressed up the bank by some type of gentle pumping system and is finally removed from the slurry for desorption of the values. Not all tanks are necessarily used for both

leaching and adsorption. The carbon may be removed for stripping prior to reaching the tank at the head of the bank so that one or more of the initial tanks may be leach only.

Despite the intrinsic simplicity of the process, interaction of the design and operating variables may be very complex so that it is not easy to intuitively predict the results of an operational or configuration change. The model was developed to assist in making decisions about circuit changes, to help identify anomalous behaviour and to provide insight on the influence of the major variables for guidance in design and operation of a CIL/CIP circuit.

THE MODEL

The model used in this study is an idealised representation of the leaching and adsorption system. The model has been structured so that all required parameters may be derived from readily obtainable plant data and thus easily fitted to any particular plant.

There are three primary elements in the model – the representation of leaching as a two rate system, the power law equation relating equilibrium solution concentration and carbon loading and the mass balance constraints across each tank and the bank as a whole.

Leaching

The representation of ore as consisting of a fast leaching component and a slow leaching component is similar to the way in which flotation has been interpreted and modelled (Cameron *et al*, 1971; Imaizumi and Inoue, 1965). While it is not claimed that this is a rigorous representation of the physical leaching properties of an ore, nonetheless it provides a convenient way of visualising the leaching behaviour and so long as it matches the leach profile over the range of leaching times under consideration then valid simulations may be obtained.

In a batch leaching situation:

$$F_{rem} = f.e^{-\alpha T} + (1-f).e^{-\beta T} \quad (1)$$

where F_{rem} is the fraction of silver or gold remaining unleached, f is the fraction of the gold or silver which is fast leaching, T is the time of leaching and α and β are the leaching rates of the fast and slow components (reciprocal time).

It follows from this that if tanks are fully mixed, then in a continuous flow situation, where f_{in} is the fraction of fast leaching material entering, the fraction of values unleached in any tank is:

$$F_{rem} = \frac{f_{in}}{(1 + \alpha T)} + \frac{1 - f_{in}}{(1 + \beta T)} \quad (2)$$

The leaching rates are primary inputs to the leaching model for simulation. They can be estimated from the leaching profile down the bank. Leaching rates, and hence the parameters for the leaching rate equations, would be expected to be a function of the ore type, the grind, the general chemical environment, particularly available cyanide and whether oxygen is used to enhance the leaching process.

Thus unexpected deviations from a normal leaching profile may be due to lack of cyanide or aeration (or oxygenation) deficiencies. Identification of such deviations may be obscured by short term variations in ore (order of hour to hour) changing the material from which the contents of any tank have been derived.

In order to fit the model to any circuit, data must therefore be obtained by sampling the feed to the circuit and the tailings from each tank. A series of campaigns may be required to obtain sufficient data to accommodate changes in parameters due to variations in head grade and ore

type. The number of such ‘runs’ needed to obtain reasonable averages may be reduced by sampling each tank in succession allowing for the residence time of slurry in each tank, thus removing the dynamic effect of changes in ore properties and feed rate changes. The solids, solution and carbon in all samples are assayed for both gold and silver to give the assay profiles for all components down the tanks on each occasion. If it is desired to extrapolate simulations reliably to a more barren tailing, plant leaching data may be extended by bottle rolling final tails for an additional period or special plant runs may be done at reduced feed rate. Where appropriate, adjustment should be made for the difference between continuously mixed conditions in a tank and batch conditions in a bottle.

Carbon Adsorption

Ideally the sampling of solution and carbon would be taken immediately prior to moving the carbon on to the next tank up the train. In this way the carbon has had maximum opportunity to equilibrate fully to the solution concentration in a tank. In practice because of head grade variations, the solution concentration in a tank does not remain steady and timing sampling to correspond to carbon movement is difficult to coordinate. However if the approach of the carbon to equilibrium loading in any tank occurs at a reasonable rate (eg within a few hours) then there may be quite a big window for satisfactory samples to be obtained.

As concentrations of the gold and silver cyanides in solution (max. order of 40 ppm) and on the carbon (< 3 per cent) are relatively low, the adsorption of gold and silver on active carbon in the model has been expressed by the Freundlich equation (Treybal, 1955):

$$L = A \cdot s^N \quad (3)$$

where L is the carbon loading, s is the solution concentration of gold or silver and A and N are the adsorption constants. The model assumes that gold and silver adsorb independently and that there is no interaction at typical maximum loadings. The model also assumes that adsorption and desorption may both occur in the circuit provided the carbon has adequate residence time in each tank. Given carbon residence times in each tank typically of the order of at least ten hours this is observed in practice. Hence the loading on the carbon removed from the circuit is considered to be in equilibrium with the solution in each tank regardless of the solution profile and especially with the solution in the last tank and the tank from which the final loaded carbon is removed. As will be seen this proves to be important when significant solution losses occur in circuit tails.

Mass Balance

Values enter the circuit primarily contained in the ground ore fed to the circuit. Water being recycled to the circuit may contain values especially water returned from a thickener which is intended to recover unused cyanide and values which would otherwise be lost in solution in the tailings. Regenerated carbon may also contain some residual values. The mass balance for the circuit and for any tank may be written as follows:

$$G(h-t) = C(L_{out} - L_{in}) + G \frac{(1-p)}{p} (s_{out} - s_{in}) \quad (4)$$

where G is the feed rate of ore, h is the head grade of the feed to the entity (circuit or tank), t is the tailings grade, p is the per cent solids in the slurry and s is the solution concentration. L_{out} and s_{out} are related by the adsorption equation. t is calculated progressively down the bank, the tailings from any tank becoming the head to the next tank. For the circuit overall h is the initial head and t the tails from the final tank.

These equations were programmed in Excel, solution outputs from one tank being the input to the next and the carbon output from one tank being the carbon input to the preceding tank. The number and function (leach only or leach and adsorption) of the tanks is set in any version of the program and the size may be input. Normal inputs apart from the leaching and adsorption parameters were the feed rate of ore, the per cent solids in the slurry, the carbon advance rate and the residual loadings of the regenerated feed carbon. If the effect of return of water containing values being recycled from a thickener was being explored then a further input was the fraction of water in final tails which was recycled. A direct analytical solution of the complex equation set is not readily obtained, so the circuit was generally solved by progressive approximation of the solution concentration of the final tails. Alternatively the final solution concentration for either gold or silver could be chosen and one of the other operating variables changed to give the selected concentration.

SOME MASS BALANCE CONSIDERATIONS

The overall mass balance provides some preliminary insights into the primary relationships between the key operational variables. A key design and operating consideration is selecting a carbon advance rate. If a particular loading is targeted for say gold, then a corresponding advance rate would be:

$$C = \frac{G \cdot h_{Au}}{L_{Au}} \quad (5)$$

This simple relationship underlines the fact that the advance rate is the most important aspect of carbon management in circuit performance. The quantity of carbon in each tank is only important in ensuring that the carbon has sufficient residence time to approach equilibrium with the solution closely and that it is in sufficient concentration to be readily moved from tank to tank.

The fractional recovery, R , in the circuit is:

$$R = \frac{C(L_{out} - L_{in})}{G \cdot h} \quad (6)$$

L_{in} may be considered to be negligible. If it is assumed that the ore is very fast leaching so that all the values are leached in the first tank, then the maximum possible solution concentration in the first tank will be:

$$s = \frac{h \cdot p}{1 - p} \quad (7)$$

Substituting from Equations 3 and 7, Equation 6 above then becomes:

$$R = \frac{C \cdot A}{G \cdot h^{(1-N)}} \left(\frac{p}{(1-p)} \right)^N \quad (8)$$

This assumes that there is sufficient tankage for the values to be adsorbed from solution so that carbon entering the first tank is already well loaded. If this ratio, R , is significantly greater than 1, then the solution concentration in the first tank will be less than the maximum possible. If the ratio is significantly less than 1 this implies that values are being fed at a rate greater than they can be recovered and a solution loss will result. This could mean that the carbon advance rate, C , is too low so that leached values cannot be taken out of the circuit at a

high enough rate. In the same way if A (the primary constant in the adsorption relationship) is low, a low carbon loading will demand an unduly high value of C in order to avoid solution loss. A low A may be due to inadequate regeneration, degradation or fouling of the carbon.

As N is typically in the range 0.65 - 0.85, the value of R proves to be relatively insensitive to the head grade h , a doubling producing only a change of about 20 per cent in R . In contrast the recovery from solution may be very sensitive to pulp density, an increase in p of six per cent from say 40 to 46 per cent increases R by about 20 per cent.

These considerations form a background for the simulation studies which embody a more complete representation of the leaching and adsorption circuit.

SIMULATION STUDIES

A standard set of parameters were chosen for the simulation. The leaching parameters are shown in Table 1 below. These were derived from a range of plant and laboratory data and represent an ore in which the gold leached readily but the silver in contrast was more resistant to extraction.

TABLE 1

Leaching parameters.

	No supplementary oxygen		With oxygen	
	Gold	Silver	Gold	Silver
Fraction fast leaching, f	0.85	0.45	0.88	0.55
Leaching rate fast fraction, α	0.75	0.80	4.00	2.00
Leaching rate slow fraction, β	0.035	0.020	0.030	0.012

The effect of oxygen addition on the parameters is also typical of the data examined. There is a relatively small increase in the fraction which is fast leaching, a major increase in rate at which this fraction leaches and a decrease in the leaching rate of the slow leaching fraction. The leaching profiles generated by these parameters are illustrated in Figure 1. The profiles are for an ore feed rate of 209 t/h at 38 per cent solids to a series of tanks of 840 m³ active volume. The residence time in each tank is very close to two hours.

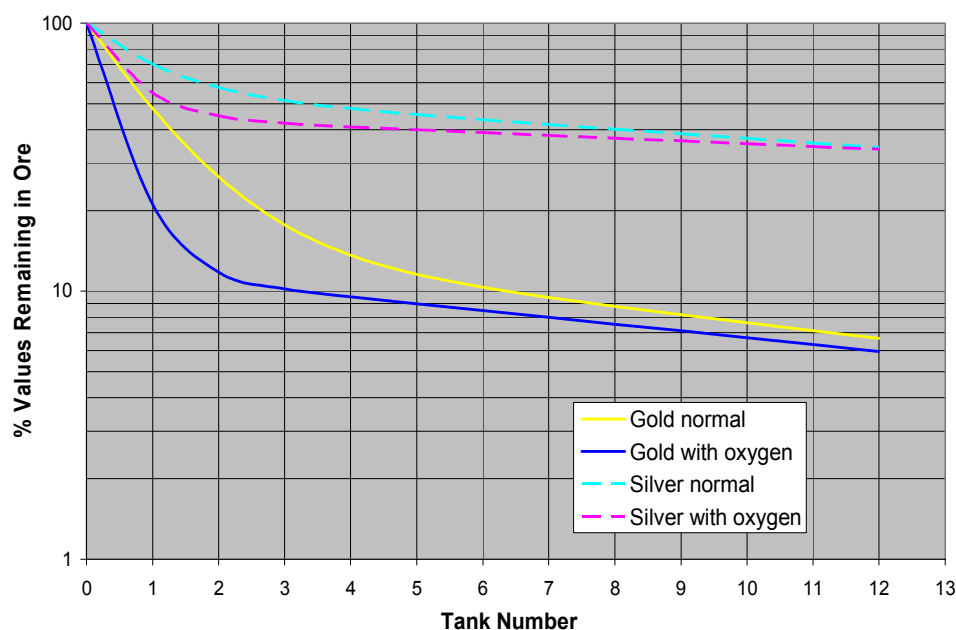


FIG 1 - Leach profiles for models used.

Whilst the initial leaching is much quicker with oxygen – in the first tank with oxygen almost as much leaching occurs as with three tanks without oxygen – surprisingly after 12 tanks (24 hours) the final amount leached is little more with oxygen (0.7 per cent additional recovery for gold and 0.5 per cent more for silver). As the number of tanks is reduced the differences with the use of oxygen increase.

Adsorption parameters used in the standard simulation are given in Table 2 below. The parameters are scaled for expressing the carbon loading and solution concentrations in ppm. They represent the adsorption of active carbon which is a seasoned inventory, ie carbon in a plant which has been in operation for a sufficient period for the carbon to have reached the equilibrium blend of carbon which has been regenerated many times, through carbon which has been though progressively fewer regeneration cycles, to fresh carbon.

TABLE 2
Adsorption parameters.

	Gold	Silver
A	2600	1900
N	0.8	0.7

The adsorption relationships are shown in Figure 2 below for a normal range of carbon loadings.

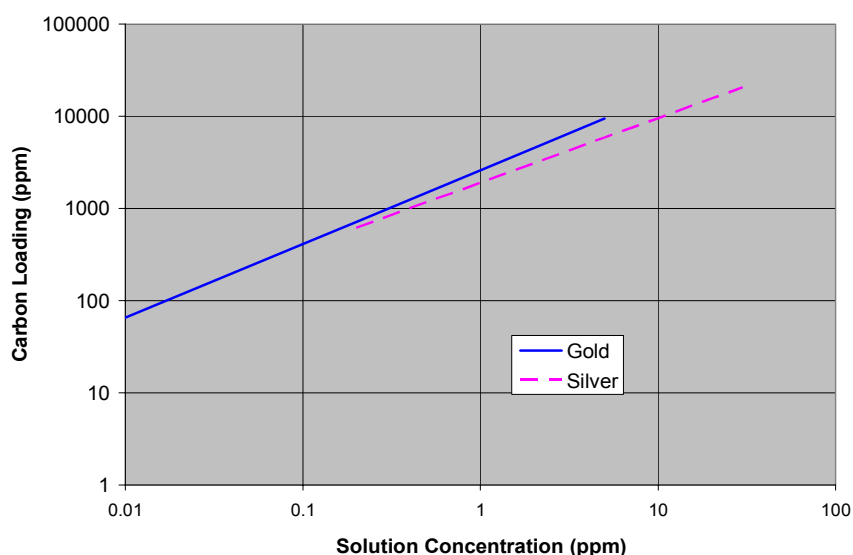


FIG 2 - Adsorption relationships.

These relationships were developed for ores where there were no important quantities of competing species (other metal cyanides) and show that, in general, gold is more readily adsorbed than silver.

The 'standard' ore being processed in the simulations was presumed to have 5 ppm gold and 35 ppm silver. Feed rates and residence times were generally kept in a range such that recoveries of gold would be greater than 85 per cent and silver 30 - 70 per cent. Thus feed rates were typically 150 -300 t/h (250 t/h when not a variable), slurry per cent about 38 per cent and residence time about 14 - 30 hours. Carbon rates were constrained to maintain a carbon loading of at least 2000 ppm gold and were generally in the range of 6 - 10 t/day.

Whilst the model studies illustrate typical behaviour and generate some broad principles of circuit design and management, decisions in relation to any particular plant will be very dependent on head grades, the actual range of leaching rates for any ore, the adsorption parameters pertaining to a seasoned carbon inventory and the cost of any indicated changes.

Economic evaluations have therefore not usually been presented. However in this study a percentage change in silver recovery has approximately only one tenth of the value of the same percentage change in gold recovery.

CIRCUIT DESIGN

Number and size of tanks – constant total residence time

The effect of additional tanks of the same size on recovery was investigated by maintaining a feed rate of 250 t/h at 38 per cent solids with a carbon advance rate of 10 t/day. The number of tanks was varied from 4 to 12 changing tank size so that the total tankage volume remained constant at 8400 m³ and total residence time did not change (16.7 hours). The solids and solution losses for gold and silver for constant tankage volume are shown in Figures 3 to 6 below.

As the number of tanks is increased the flow approaches ‘plug flow’ which would correspond to an infinite number of tanks, the most efficient leaching and the minimum solid loss for the set residence time. With an increasing number of tanks in the simulation the solid loss progressively approaches this value.

The solution loss for gold follows a similar pattern, decreasing with increasing number of tanks whether oxygen is used or not. Four tanks are insufficient for adsorption and seven is clearly enough. However the pattern of solution loss for silver is quite different and appears at first to be anomalous. The loss passes through a minimum at five tanks for no oxygen and seven when oxygen is used. It then increases steadily. This means that from the point of view of silver metallurgy there can be too many tanks. Such behaviour appears to be associated with high solution loss and therefore an inadequate carbon advance rate for full silver recovery. However, if the carbon rate is increased sufficiently to reduce solution loss of silver to low levels and avoid this silver behaviour, the final gold loading drops undesirably low conflicting with a primary constraint.

Some insight to the interactions underlying this behaviour can be gained from the profile of solution loss down the bank – shown in Figures 7 and 8 for the case with no oxygen enhancement. The solution concentration passes through a maximum which depends on the number of tanks in the bank and is accentuated by late leaching. This behaviour has been observed frequently in practice confirming the realism of this aspect of the simulations. Interchanging the adsorption parameters between gold and silver provides further insight – Figures 9 and 10. A maximum emerges in the gold solution profiles for seven or more tanks and the maxima and final solution concentrations for silver are greatly reduced.

This shows that the behaviour is a function of the interaction of leaching rates, adsorption relationships and carbon advance rates. In a situation where the carbon rate is insufficient to carry all leached values from the circuit or where leaching occurs late, fresh carbon entering the circuit at the last tank meets a high solution concentration and is quickly loaded. The carbon thus carries values back up the train. However in the initial tanks insufficient leaching has occurred to provide solution concentrations which will sustain the carbon loadings of the entering carbon. The values therefore desorb until equilibrium is attained, creating a maximum as the progressively more lightly loaded carbon moves further up the bank. The behaviour is less pronounced with fewer tanks as the residence time in each tank is greater producing more leaching in the earlier tanks, higher early solution concentrations, higher final carbon loadings and lower final solution losses.

Number of tanks of constant size – increasing total residence time

The complementary tank change is adding more tanks of the same size, thus increasing residence time. This may be done in order to gain more recovery or to permit a feedrate increase. Figure 1

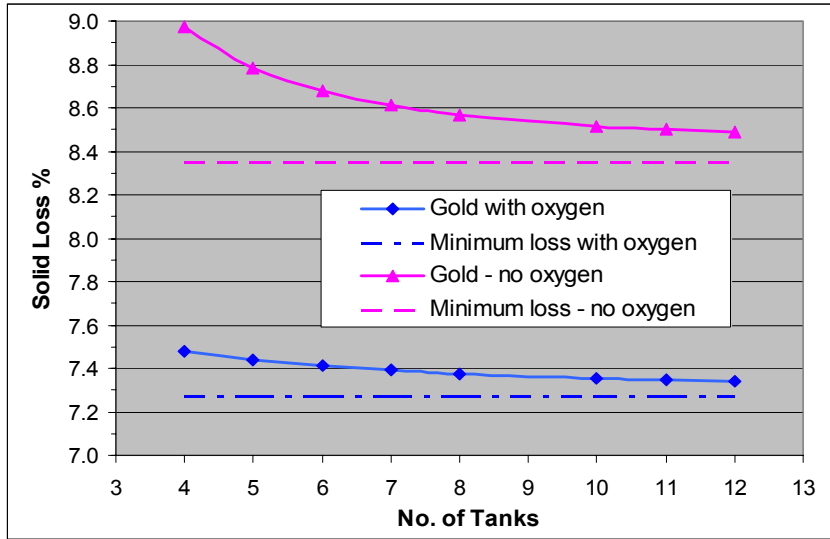


FIG. 3 - Solid loss for gold, constant total tankage volume.

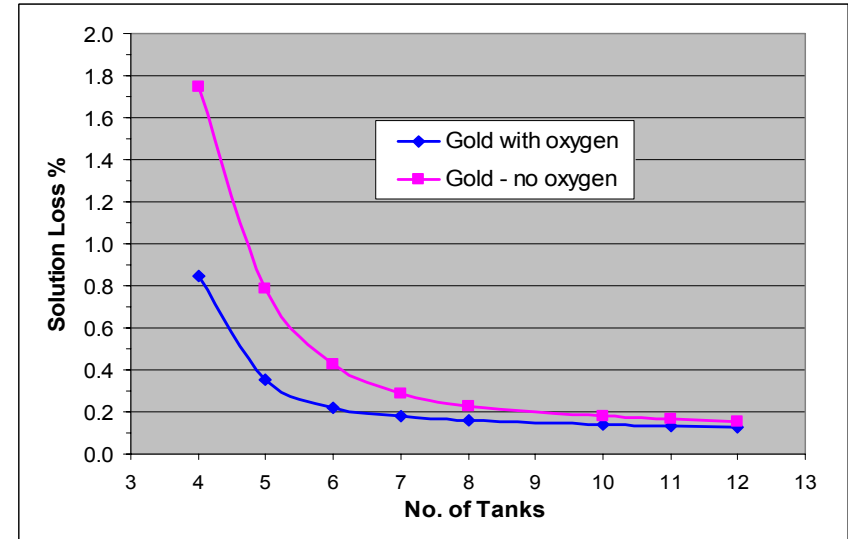


FIG 4 - Solution loss for gold, constant total tankage volume.

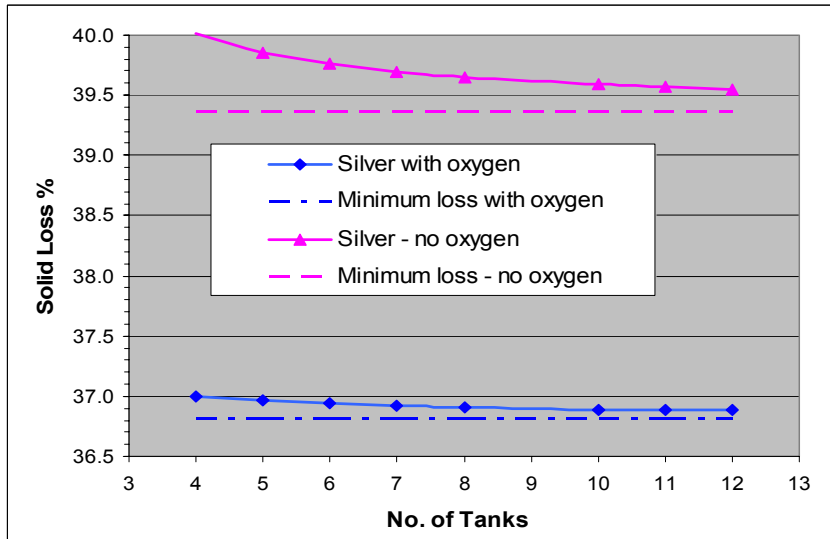


FIG 5 - Solid loss for silver, constant total tankage volume.

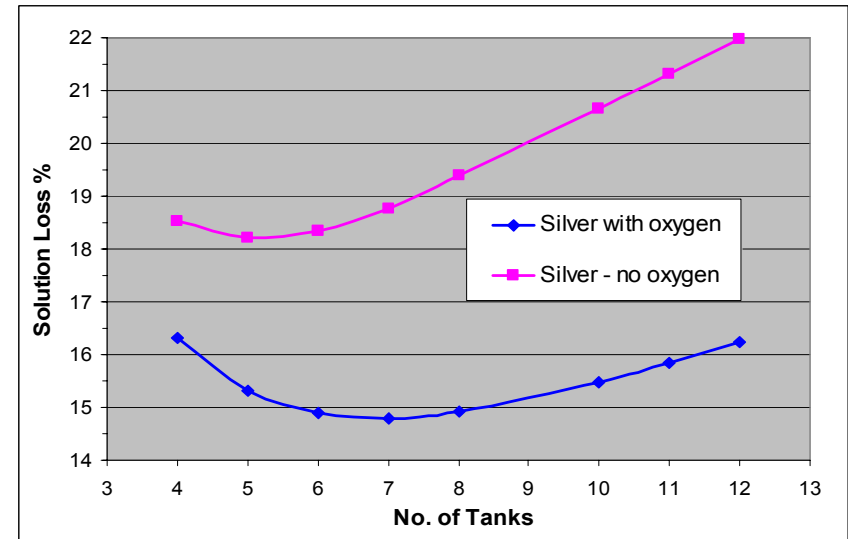


FIG 6 - Solution loss for silver, constant total tankage volume.

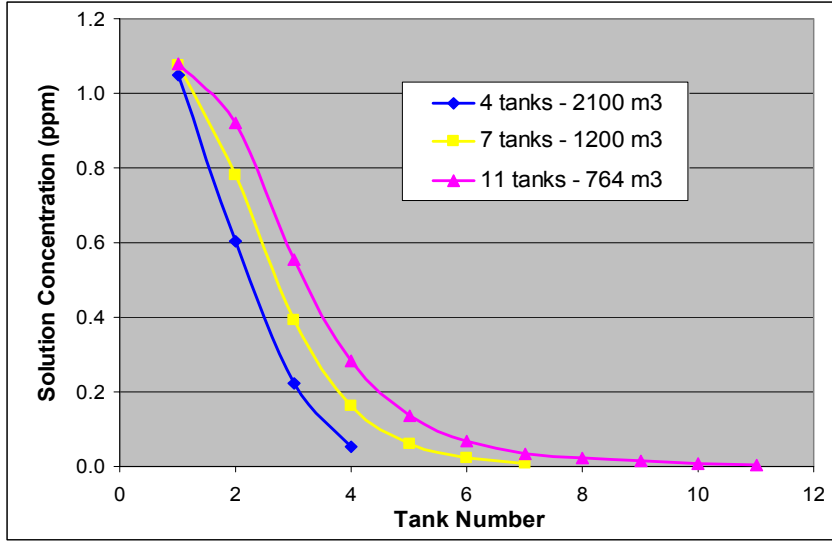


FIG 7 - Solution profiles for gold (no oxygen).

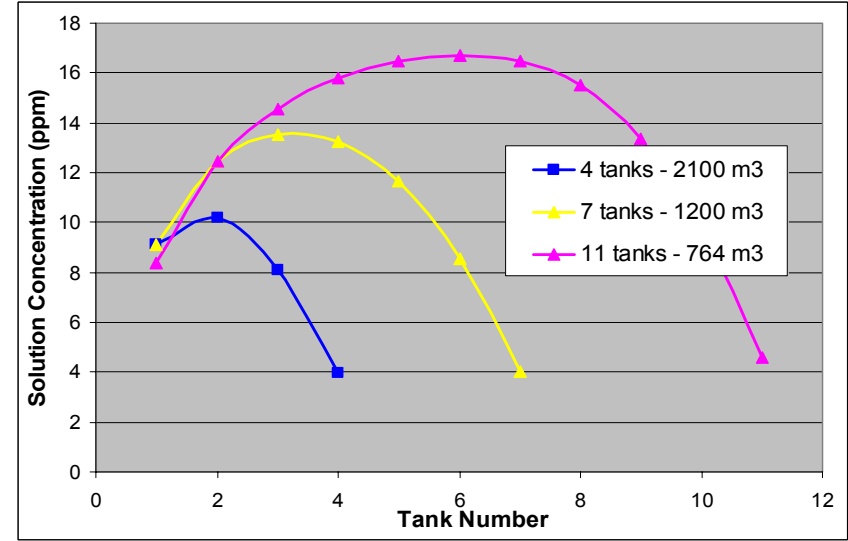


FIG 8 - Solution profiles for silver (no oxygen).

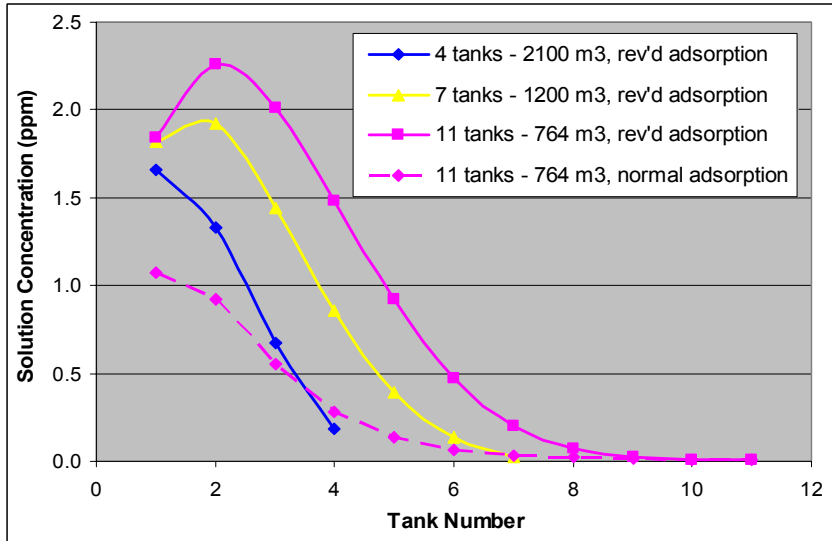


FIG 9 - Solution profiles for gold with interchanged adsorption.

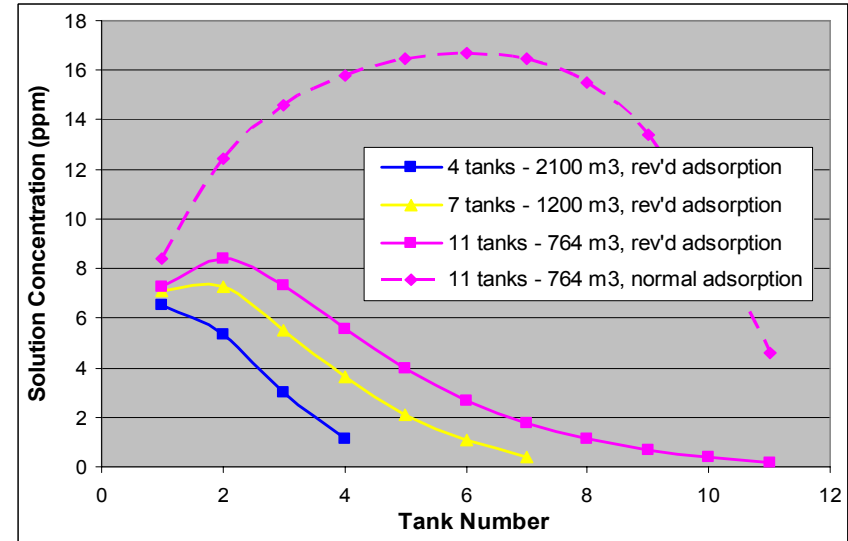


FIG 10 - Solution profiles for silver with interchanged adsorption.

shows the increase in leaching which is typical of adding tanks. The change in solution loss with added 840 m³ tanks is shown in Figures 11 and 12. These show that whilst solution loss of gold decreases steadily with additional tanks and increasing residence time ranging from 6.7 to 20 hours, the behaviour of silver is more complex. Even though solution loss of silver may increase, the total recovery increases with extra tanks. Solution profiles are very similar to those in Figures 7 and 8 showing that in this case the carbon advance rate and adsorption capability are inadequate to recover all the additional silver which has been leached in the additional tanks.

One of the questions which is relevant is whether additional recovery can justify any additional tanks. The annual revenue which can be credited to each additional tank may be estimated in order to illustrate the significance of these enhanced recoveries in this example. The results are shown in Table 3 below for the assumptions of 8000 hours of operation at 250 t/h, 5 ppm gold, 35 ppm silver and \$US400/oz.

TABLE 3
Additional annual revenue resulting from an additional tank.

No. of the additional tank	GOLD		SILVER		Extra tank annual revenue, \$USM	
	Oxygen	No oxygen	Oxygen	No oxygen	Oxygen	No oxygen
5	0.49	2.49	1.89	2.16	1.61	5.62
6	0.45	1.44	1.18	1.34	0.90	2.98
7	0.43	0.95	0.74	0.82	0.68	1.70
8	0.44	0.79	0.47	0.49	0.61	1.14
9	0.73	0.55	0.27	0.29	0.53	0.86
10	0.36	0.50	0.16	0.16	0.49	0.71
11	0.35	0.48	0.10	0.09	0.47	0.65
12	0.34	0.45	0.06	0.05	0.44	0.60

If an additional tank costs \$1.2M and must be paid out in two years, when oxygen is used an eighth tank is justified. When oxygen is not used then up to 12 tanks are indicated. In this latter case the circuit would need to be re-evaluated on the basis of two parallel banks. Clearly a proper analysis would need to take account of operating and maintenance costs of additional tanks and include a pertinent evaluation of the cost of capital. It is of interest in this example that, even though the head assay of silver is seven times that of gold, on average over 90 per cent of the additional revenue comes from additional gold recovery. Conclusions from an analysis such as this are very dependent on head grades, feed rate, carbon rate and model parameters and thus need to be related to a specific plant to have real significance.

Parallel banks

Metallurgically two parallel banks contrasted with the same tanks in series are equivalent to doubling the tank size with half the number of tanks. The configuration was simulated by halving both the feed rate and the carbon advance rate. A summary of results for comparing parallel banks of 4, 5 and 6 tanks each with banks of 8, 10 and 12 tanks are given in Table 4 below. Differences are expressed as the loss in the parallel arrangement minus the loss in the corresponding serial bank

The table shows that losses in solid are greater when the tanks are arranged as two parallel banks rather than in series. The same is true for solution loss of gold and most total losses. However solution loss of silver is generally less for the parallel bank arrangement because the longer residence time in the early tanks gives more leaching and a higher solution concentration in these tanks resulting in higher final carbon loadings and improved solution recovery.

TABLE 4 Differences in losses for parallel and serial banks.

Total Number of Tanks	Total Res.Time (hours)	Gold			Silver		
		Difference in solid loss %	Difference in sol'n loss %	Total loss % difference	Difference in solid loss %	Difference in sol'n loss %	Total loss % difference
No oxygen							
8	13.4	0.55	1.71	2.26	0.37	-0.86	0.48
10	16.7	0.27	0.61	0.88	0.27	-2.44	2.17
12	20.1	0.18	0.24	0.42	0.25	-3.61	3.36
With oxygen							
8	13.4	0.08	0.68	0.75	0.07	1.30	1.37
10	16.7	0.08	0.21	0.30	0.07	-0.17	-0.10
12	20.1	0.09	0.10	0.19	0.08	-1.24	-1.15

When oxygen is used all differences are less. Overall the serial arrangement is preferable metallurgically. However for practical operation a parallel arrangement would probably be preferable. If the differences are expressed in terms of annual revenue then a more clear-cut perception may emerge for decision making as Table 5 below illustrates.

TABLE 5 Revenue differences serial minus parallel tank arrangements.

Total Number of Tanks	Revenue difference \$USM/y		
	Gold	Silver	Total
No oxygen			
8	2.91	-0.06	2.85
10	1.13	-0.28	0.85
12	0.54	-0.43	0.11
With oxygen			
8	0.97	0.18	1.15
10	0.38	-0.01	0.37
12	0.25	-0.15	0.10

In general in this example differences in gold recovery are more financially significant than for silver. Regardless of whether oxygen is used or not, eight tanks are clearly better operated in series and practical advantages would probably outweigh financial imperatives for operating twelve tanks as two parallel banks. For ten tanks parallel banks are probably preferred if oxygen is used but not otherwise.

Thickener

Water fed to a tailings thickener splits between the thickened underflow and the clear overflow which is recycled for reuse in the plant. Values in solution split with the water. Operation with the maximum of dilution water consistent with water balance constraints and at maximum per cent solids in the underflow, maximises return of values. Figures 13 and 14 show the effect of such a thickener on solution concentrations for a bank of ten tanks where 70 per cent of values are recycled to the head of the bank. As the solution loss of gold is low, the recycle has little impact on the solution profiles, virtually all the recycled gold in solution being recovered. Nonetheless even the small solution losses are reduced by 70 per cent and the thickener is a significant safeguard against solution losses.

For silver with a major recycle of values the whole solution profile is raised. Carbon loadings and thus overall recoveries are increased. In this example 85 per cent of the silver in the recycle is effectively recovered and final solution loss is 40 per cent of what it would be in the absence of a thickener (eg with oxygen the solution loss is reduced from 15.5 per cent to 6.4 per cent).

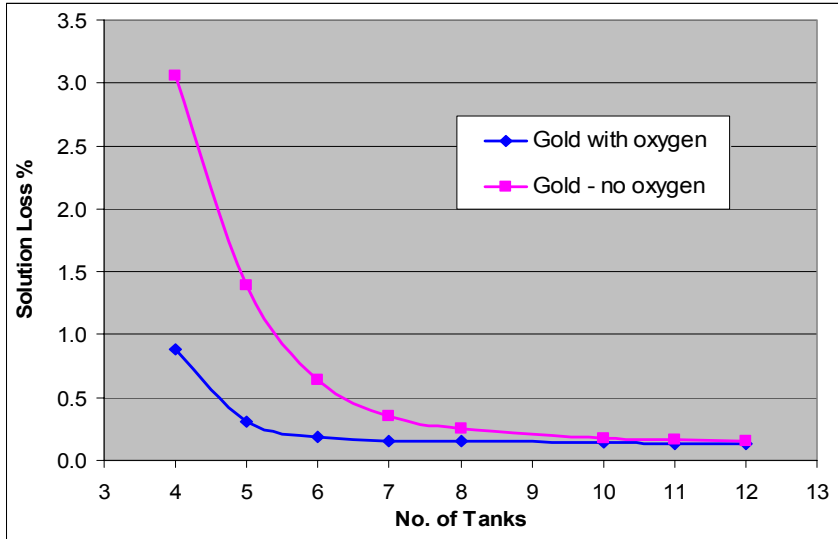


FIG 11 - Solution loss of gold with more, constant size tanks.

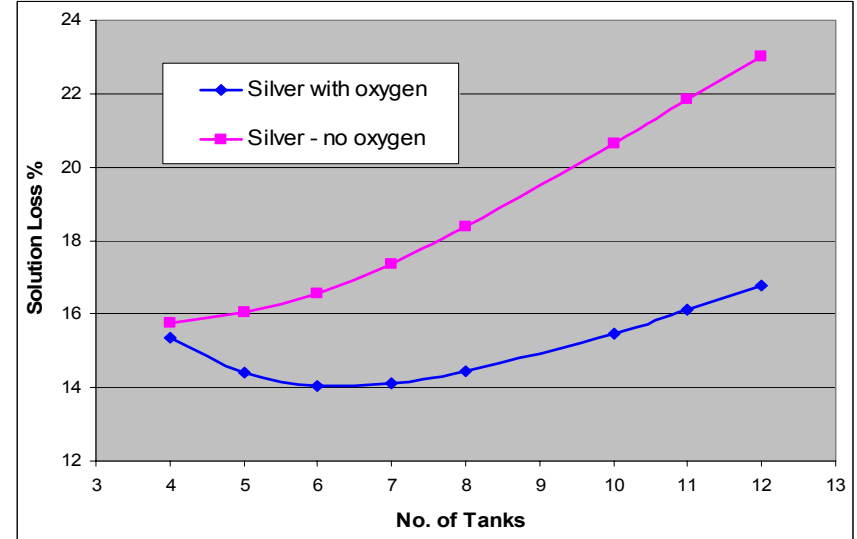


FIG 12 - Solution loss of silver with more, constant size tanks.

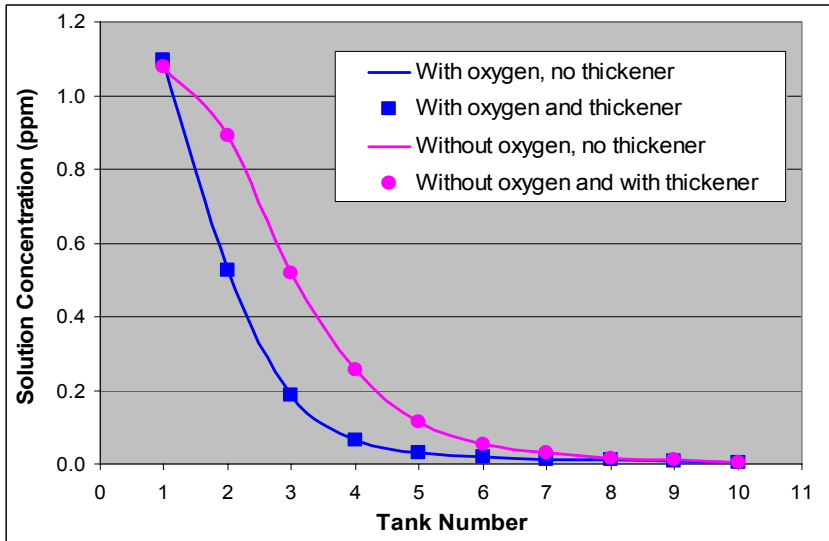


FIG 13 - Effect of thickener on gold solution profiles for ten tanks.

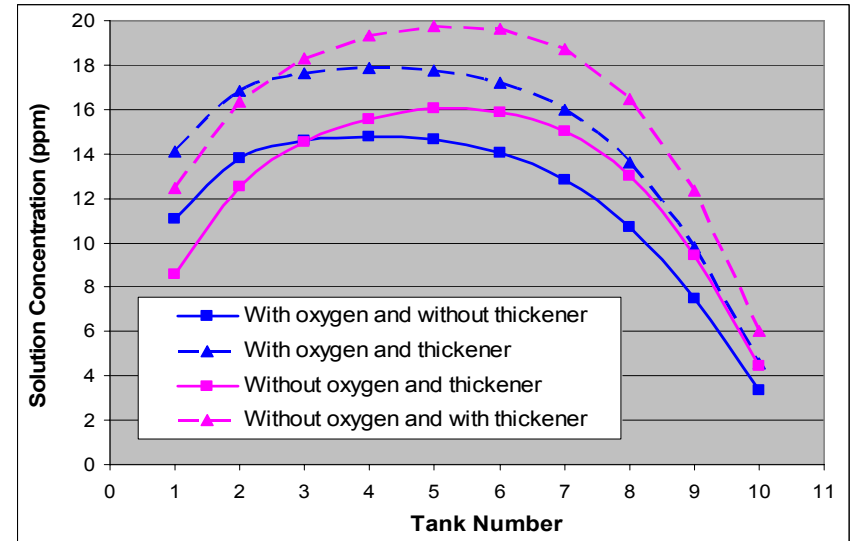


FIG 14 - Effect of thickener on silver solution profiles, ten tanks.

Leach only tanks

It is common practice to use early tanks in a bank purely for leaching as shown in the diagram in Figure 15 below for a bank with two leach only tanks. The impact of this on solution profiles is shown in Figures 16 and 17. For gold the concentration rises steadily in leach only tanks from tank to tank. This is shown as a separate relationship in the figures. Once carbon is introduced, the gold concentration falls steadily to the end of the bank. As there is little solution loss and the carbon rate is constant at 10 t/day, the carbon loading is virtually unchanged and the concentration in the tank from which carbon is extracted is the same regardless of how many tanks are leach only. As the number of tanks used for adsorption drops there is a very small increase in final solution concentration and solution loss rises slightly.

For silver the concentration also rises steadily for the leach only tanks. However even for five 'leach only' tanks the concentration continues to rise for tanks after carbon is introduced by the mechanisms discussed previously. Nonetheless carbon is extracted from tanks with progressively higher solution concentrations as the number of leach only tanks is increased. Carbon loadings therefore increase and solution losses fall. The gain in silver recovery with one leach only tank is over three per cent, for two leach only tanks the additional recovery is 2.4 per cent when oxygen is not used and only 1.0 per cent with oxygen, but in going from two leach only tanks to three the gains are 1.3 per cent and 0.4 per cent respectively. One or possibly two leach only tanks would appear to be usually justified with a three or more leach only tanks unlikely to be of much value.

This is another method of reducing solution losses and the use of one or two leach only tanks generally gives a more robust circuit. However, the use of leach only tanks leaves the circuit vulnerable to solid losses with ores in which preg-robbing may occur.

OPERATIONAL VARIABLES

Ore feed rate

Feedrate is one the primary operational variables and there can be a constant pressure to increase feedrate sometimes without full regard to the consequences. Figures 18 to 21 show the effect of feedrate change on losses in the solids and solution. When solid losses are plotted against the log of the feedrate (as shown), the data give almost a straight line plot. By maintaining the same leaching parameters, these data assume that feedrate change has not affected the size distribution of the feed ore and thus the leaching rates. In general a higher feedrate would be expected to coarsen the size distribution and reduce the fraction of fast leaching values and solid losses would increase as a result.

Solution losses for silver form a sigmoid shaped plot when a wide range of feedrates is used. The data for gold indicate only the start of such a shape as the solution losses for gold are all low.

A proper evaluation of the most desirable feedrate requires an assessment of the NPV of the remaining ore with an economic model for a range of feedrates and associated recoveries.

Percentage of solids in the slurry

As with feedrate selection, the percentage of solids in the feed slurry may be subject to various constraints. If there is no thickener between the grinding and CIL circuits, then the maximum per cent solids for CIL may be determined by grinding considerations. Viscosity and flow considerations may also limit the allowable slurry concentration. High viscosity or inadequate stirring at high per cent solids could hamper both leaching and adsorption rates. This can only be assessed by appropriate testwork and this analysis assumes there are no such limitations. Deviations from model behaviour could signal the presence of such effects. Table 6 below shows the effect of per cent solids on recovery (ten tanks, feedrate 250 t/h, carbon 10 t/day).

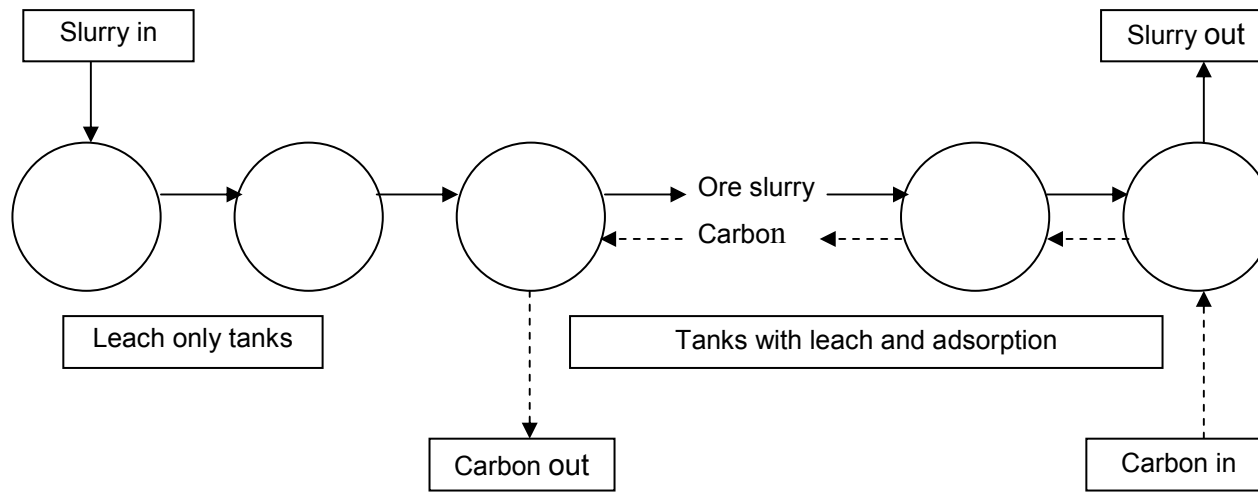


FIG 15 - Hybrid CIP/CIL circuit with two tanks 'leach only'.

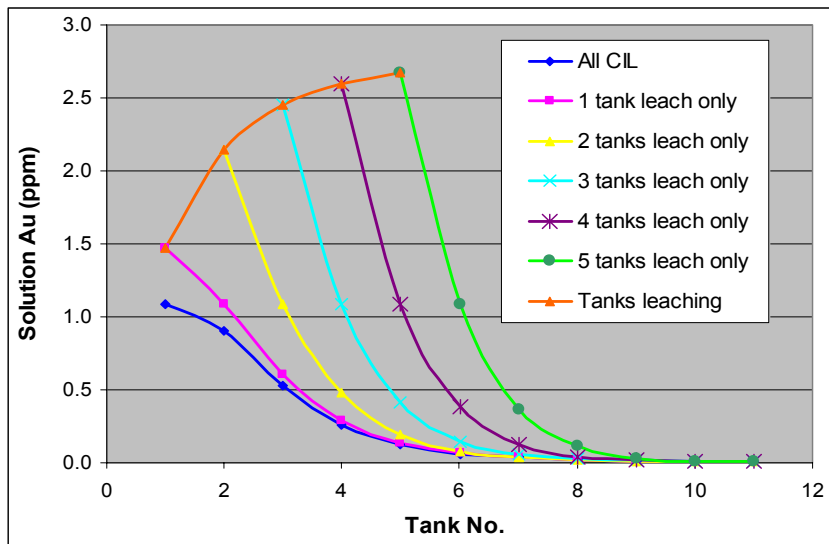


FIG 16 - Solution profiles for gold for hybrid circuits.

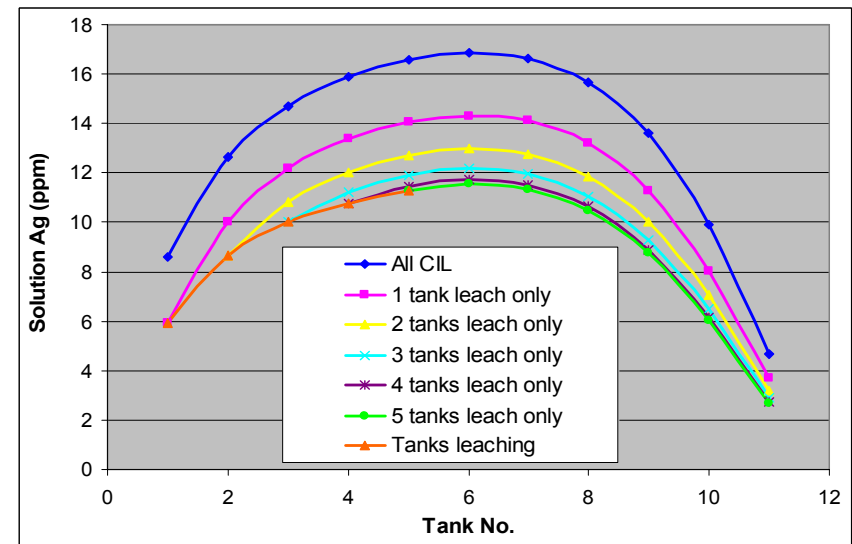


FIG 17 - Solution profiles for silver for hybrid circuits.

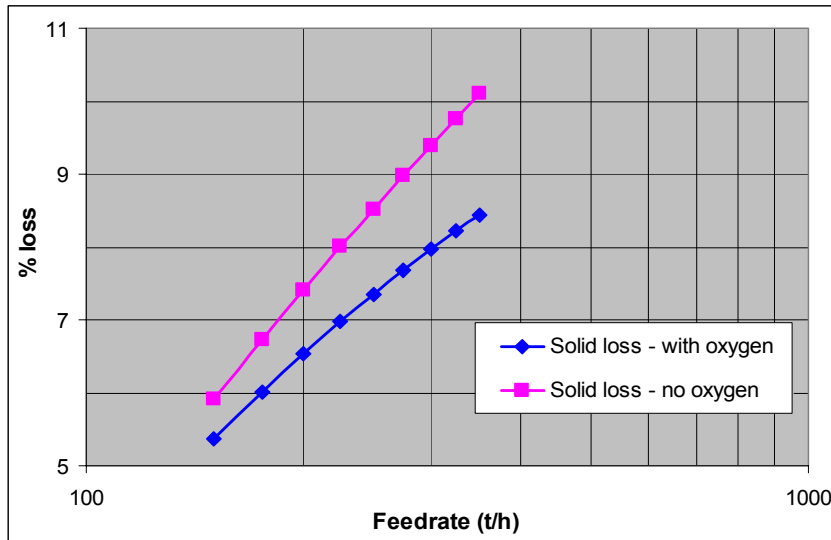


FIG 18 - The effect of feedrate on loss of gold in solid.

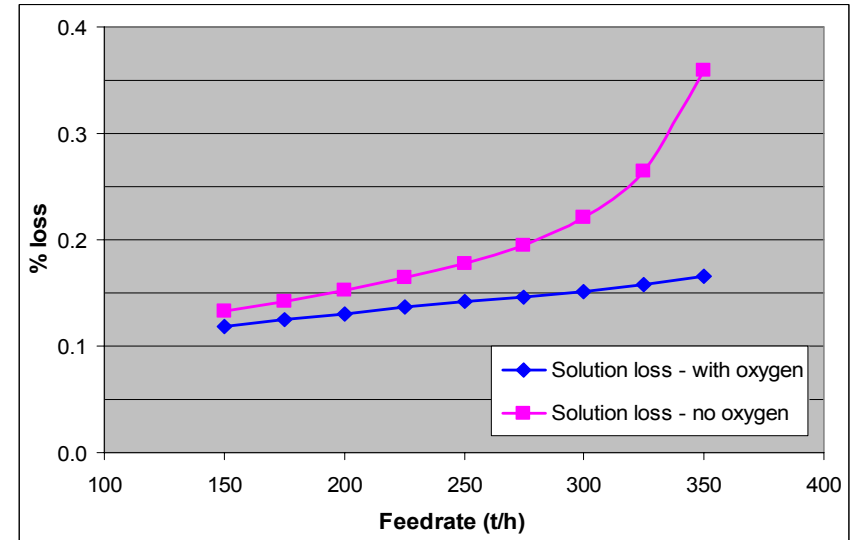


FIG 19 - The effect of feedrate on loss of gold in solution.

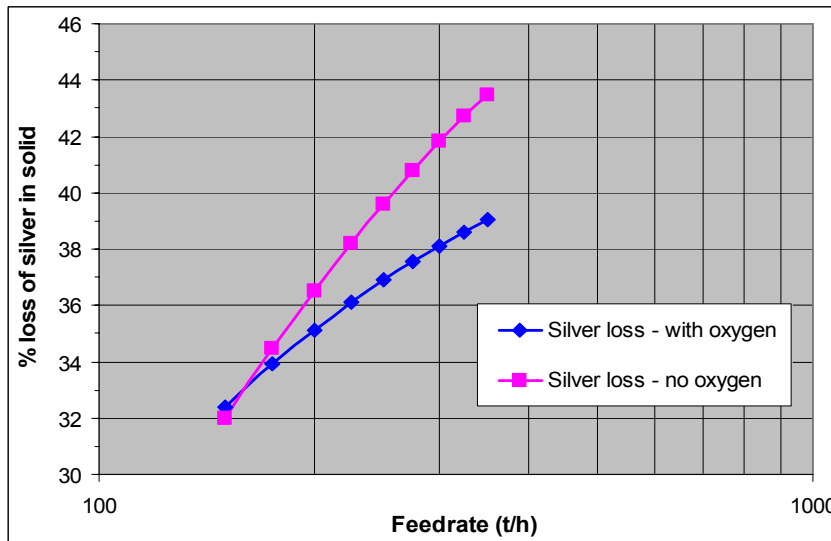


FIG 20 - The effect of feedrate on loss of silver in solid.

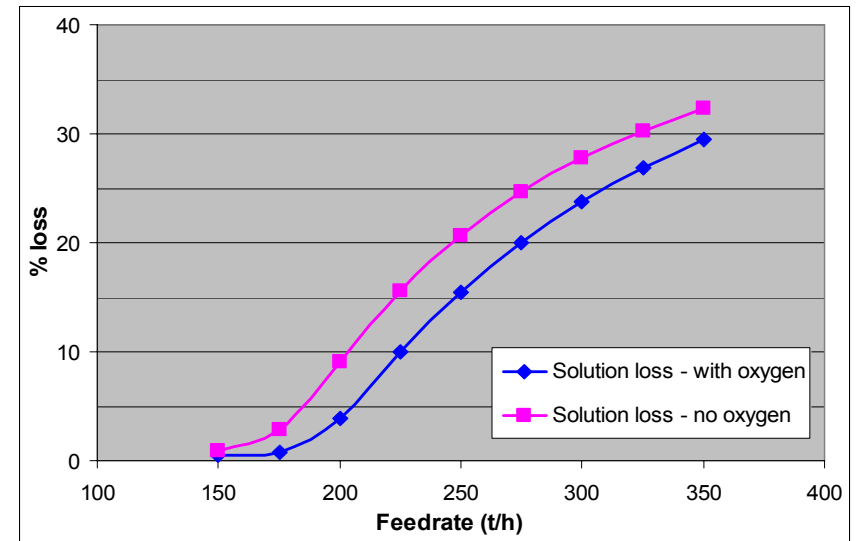


FIG 21 - The effect of feedrate on loss of silver in solution.

TABLE 6
The effect of per cent solids in slurry on recovery.

With oxygen				
% solids in slurry	Gold recovery %	Silver solid loss %	Silver solution loss %	Silver recovery %
32	91.7	38.4	24.0	37.7
38	92.5	36.9	15.5	47.6
44	93.3	35.3	6.3	58.4
50	94.0	33.7	0.7	65.6
No oxygen				
32	90.2	42.2	27.9	29.9
38	91.3	39.6	20.7	39.8
44	92.3	36.9	11.8	51.3
50	93.3	34.2	3.1	62.7

The table shows that increase in per cent solids brings very significant increases in recovery. This can be best illustrated by comparison with the effect of changing feedrate. Thus in order to effect the same increase in recovery in gold as changing per cent solids from 38 to 44, the feedrate must be reduced from 250 t/h to 200 t/h. For silver a reduction to 209 t/h is required. For gold, the solution loss is low for all per cent solids and the gain is in reduced loss in solids. In contrast for silver, although there is a substantial reduction in loss in solids, the major improvement with increasing per cent solids is in reduced solution loss.

There are two mechanisms which are operative: firstly higher per cent solids reduces slurry flow for the same solids feedrate and hence increases residence time and leaching in each tank, and secondly the values are leached into a smaller volume of solution enhancing initial concentrations and carbon loadings. At 50 per cent solids the flow of solution is half that at 33 per cent solids. With improved loadings the overall mass balance is improved and final concentration and solution loss fall. Figure 22 below shows these effects on solution concentration.

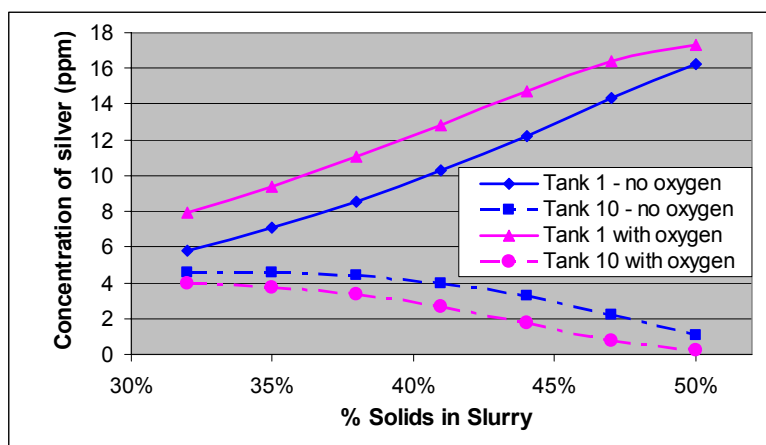


FIG 22 - Solution concentrations of silver for tank 1 and tank 10.

Carbon advance rate

Carbon advance rate impacts directly on the overall mass balance and hence on solution loss as discussed earlier. Figure 23 shows that solution loss for gold remains low until the carbon advance rate is reduced to about 7 t/day. For silver significant solution losses commence as the carbon advance rate drops below 14 t/day (Figure 24). However carbon loading for gold is close to or less than 2000 ppm at 14 t/day. For both metals solution losses are reduced with the use of oxygen.

The onset of significant solution loss can be related to solution profiles. These are shown in Figures 25 and 26. When the solution concentration peaks in tank 2, as for 7 t/day for gold without oxygen, then the circuit is prone to significant solution losses if there is any further reduction in carbon advance rate (or reduction in adsorptive capacity). The same association can be seen for silver with oxygen when the carbon rate is 14 t/day. In this way solution profiles can be a useful diagnostic tool in indicating when the plant may be on the verge of serious solution losses.

Using Equations 3, 5 and 7 the theoretical minimum carbon rate is given by:

$$C = \frac{G.h^{(1-N)}}{A} \left(\frac{1-p}{p} \right)^N \quad (9)$$

The theoretical minimum carbon advance rate for gold calculated from this relationship is 4.73 t/day and is 13.1 t/day for silver. Thus at 10 t/day C is much greater than required for gold but less than needed for full recovery of silver from solution. However in the simulations above the leaching of silver was 63 per cent for which a theoretical minimum rate would be 8.25 t/day. This assumes that the maximum possible concentration is attained in tank 1 giving maximum carbon loading and this is not generally the case. Allowing for the amount leached and using the estimates from Figures 25 and 26 above, it is estimated that a carbon advance rate about 1.7 times the theoretical minimum, allowing for the percentage leached, is desirable.

Ultimately in practice the maximum carbon rate will be dependent on the ability to move carbon through the circuit, the minimum desired loading or the capacity of the stripping and regeneration circuits. Hence it may not be always possible to avoid solution loss when there is slow leaching or poor adsorption without a major, undesirable reduction in fresh feedrate of ore.

The residual loading on barren carbon can have an undesirable effect on final solution concentration. An increase in gold loading on barren carbon from 20 to 55 ppm adds 0.2 per cent to solution loss and an increase in silver loading from 200 to 650 ppm adds two per cent to silver loss. In this study such barren loadings correspond to prudent limits.

Leaching rates and adsorption equilibria

Throughout, this paper has contrasted the behaviour of four different sets of leaching parameters – gold with and without oxygen and silver with and without oxygen. This has shown that when oxygen is used, the larger fast leaching fraction and faster leaching rates of this fraction are always advantageous. The associated slower leaching rates of the slow leaching fraction with oxygen use never had a deleterious impact overall.

To examine further the interaction of leaching and adsorption parameters, a version of the model with a single leaching rate and with both leaching and adsorption in all tanks was used. When a very high rate of 50 was used, 99 per cent of the leaching took place in tank 1. In this situation with a carbon rate of 10 t/day, recovery of gold was very close to 100 per cent, the only loss being 0.07 per cent in solution, corresponding to the equilibrium concentration with the barren carbon loading of 20 ppm. Recovery from solution was over 98 per cent complete in the first three tanks.

In contrast, loss of silver in solution was 24 per cent. The silver concentration in tank 1 was 21.3 ppm which is very close to the maximum possible for all the silver going straight into solution (21.5 ppm) as calculated from Equation 7 above. Hence carbon loadings were close to maximum. However as discussed earlier, the carbon advance rate was inadequate.

The response of gold and silver to changing the leaching rate is shown Figures 27 to 30. The loss in solution exhibits a maximum at a leaching rate of about 0.1 hr^{-1} . Low leaching rates correspond to ‘late leaching’ and reduced total leaching. Consequently the quantity leached also goes through a maximum in final tanks at about $k = 0.1 \text{ hr}^{-1}$. With the carbon rate over double the

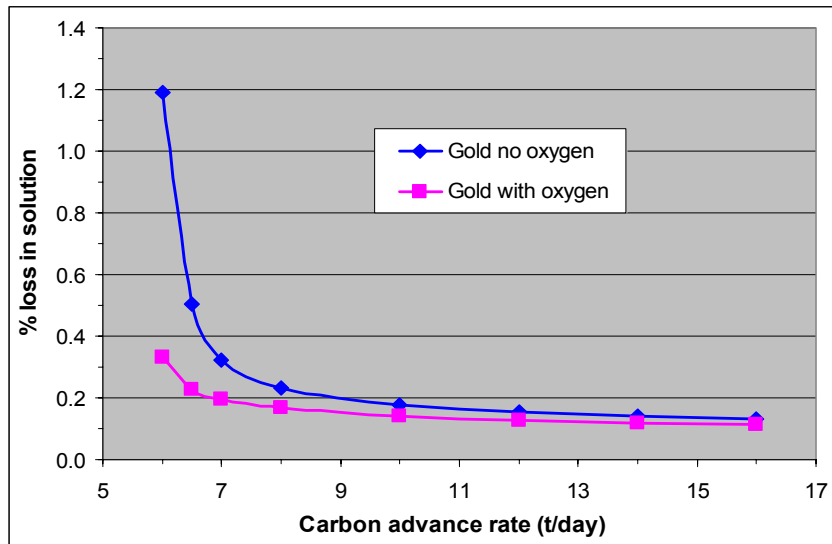


FIG 23 - Effect of carbon advance rate on loss of gold in solution.

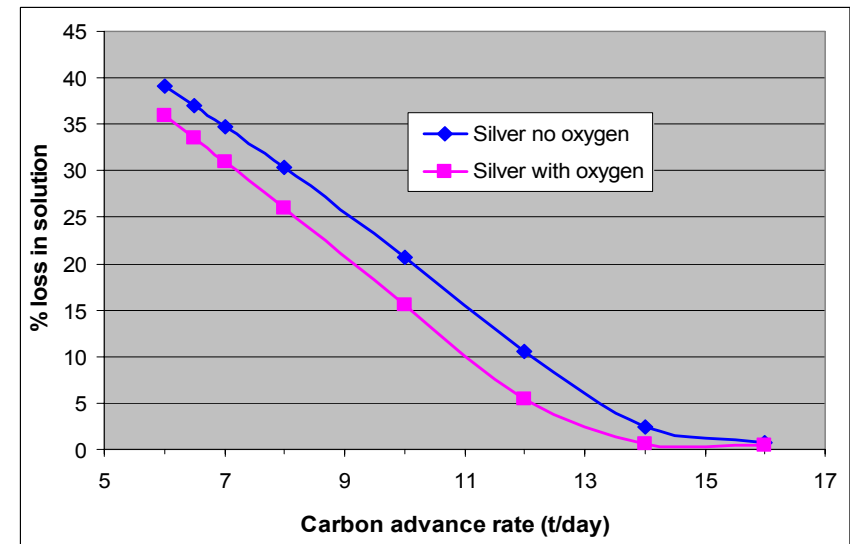


FIG 24 - Effect of carbon advance rate on loss of silver in solution.

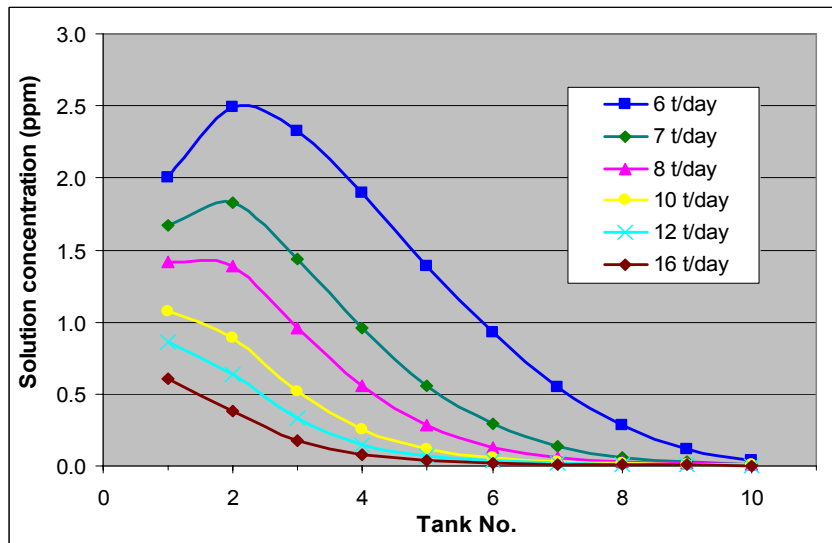


FIG 25 - Solution profiles of gold (no oxygen, various carbon rates).

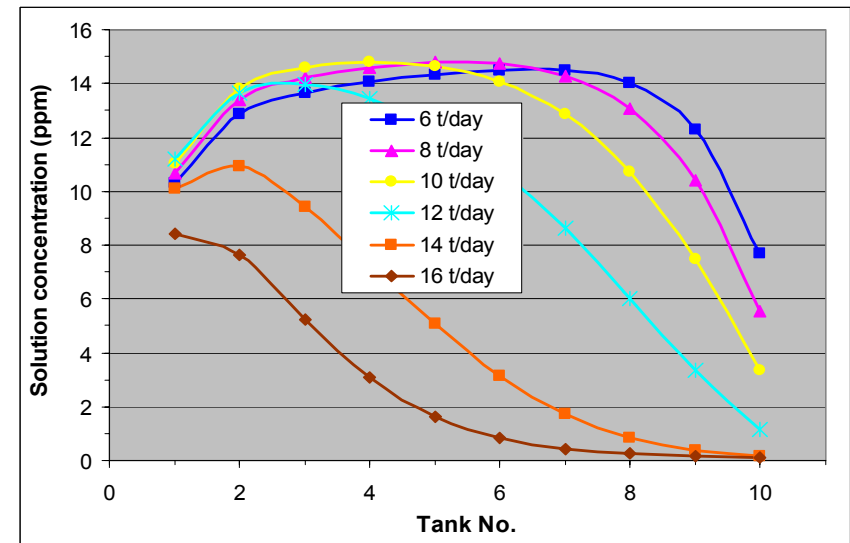


FIG 26 - Solution profiles of silver (with oxygen, various carbon rates).

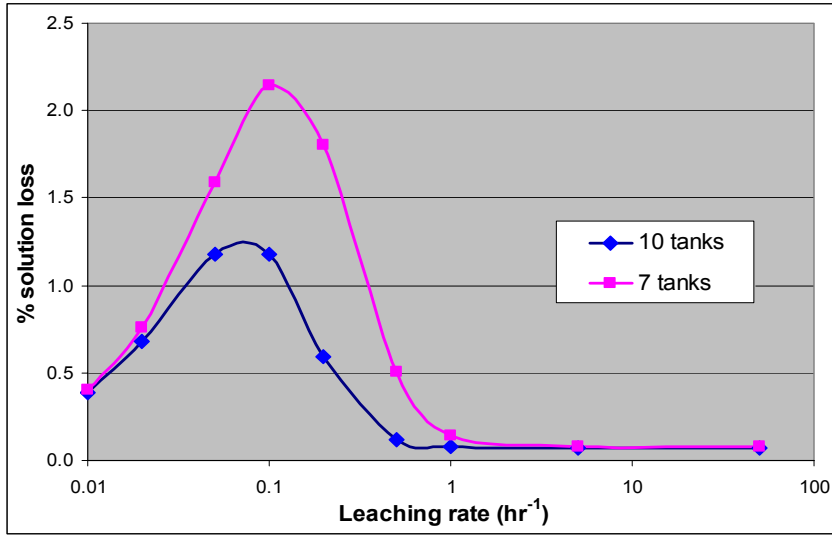


FIG 27 - Dependence of solution loss of gold on leaching rate.

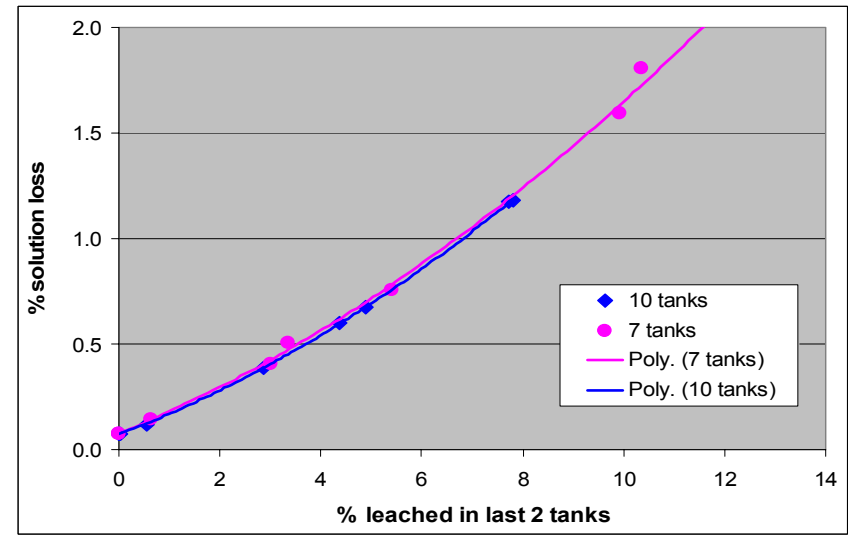


FIG 28 - Solution loss of gold vs leaching in final two tanks.

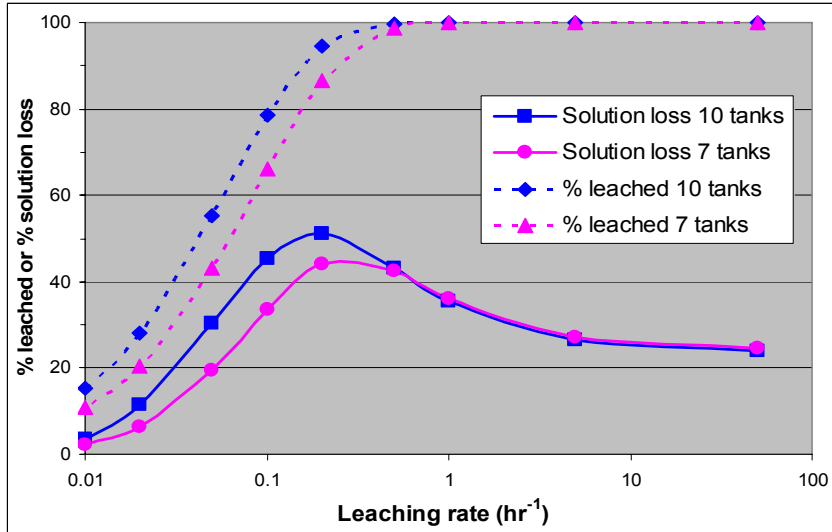


FIG 29 - Dependence of solution loss of silver on leaching rate.

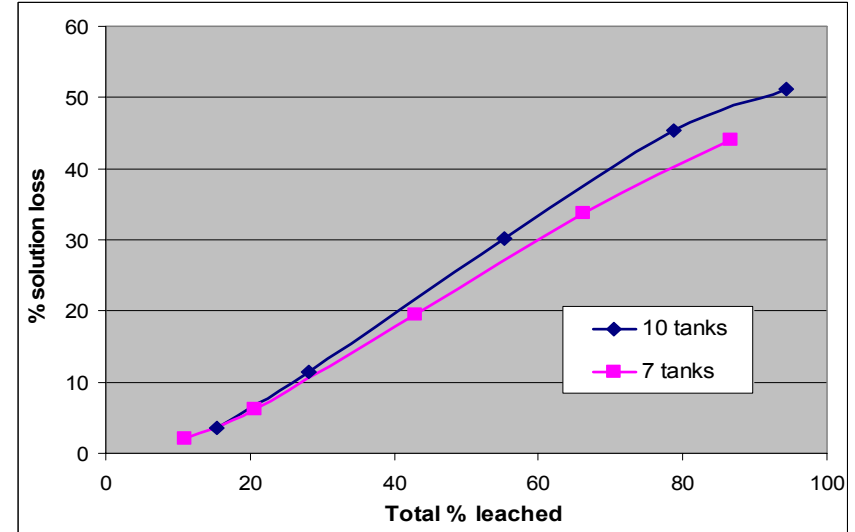


FIG 30 - Solution loss of silver vs per cent leached for $\alpha = 0.01-0.2 \text{ hr}^{-1}$.

theoretical minimum, any gold leached is quickly adsorbed and it is the leaching in the final two or three tanks which control solution loss as shown in Figure 28. In a seven-tank plant more leaching is done in final tanks than in a ten-tank plant.

An examination of the solution profiles for silver shows that for the lower leaching rates because of the inadequate carbon rate combined with late leaching, a major amount of the adsorption is done in the final tanks but solution loss tends to be related to the total leached (Figure 30). The total leached is greater for a ten-tank plant than for a seven-tank plant for any leaching rate. At a rate of 0.2 hr^{-1} leaching is about 90 per cent and solution loss is at a maximum. As the rate is increased above 0.2 hr^{-1} progressively more leaching is done in the early tanks, the concentration in tank 1 approaches the theoretical maximum and solution loss asymptotes to the value corresponding to the residual silver which is not adsorbed by the carbon.

CONCLUSIONS

A simple model of a CIP/CIL circuit has proved valuable in providing insights into the complex interactions in a leaching and adsorption plant. An examination of the influence of a range of design and operating factors has indicated preferred configurations and conditions.

For the parameters used in this study (which are believed to be typical of many ores), those for use of oxygen gave a circuit which was more robust to variations in operating conditions.

At least five tanks are normally required to ensure good adsorption and low solution loss. For six to nine tanks a series arrangement performs best. Ten tanks may be arranged in series or as two parallel banks but parallel banks are preferable for 12 tanks or more.

If solution losses are significant, a tailings thickener recycling a high percentage of tailings solution back to the circuit will bring a major reduction in these losses.

The use of one or two tanks as leach only, when there is no threat of preg-robbing, provides a degree of insurance against solution losses if solids are either late leaching or adsorption parameters are unfavourable.

The highest per cent solids in the slurry, consistent with maintaining a satisfactory size distribution of the ground ore and mixing and flow conditions in the circuit, is advantageous in limiting loss in both solids and solution. Viscosity modifiers to assist in raising per cent solids could be economically justifiable.

To ensure efficient adsorption of values the carbon advance rate should be at least 1.7 times the theoretical rate calculated from the feedrate, head grade, per cent solids and adsorption parameters.

ACKNOWLEDGEMENTS

The author would like to acknowledge the support and assistance of colleagues, who collected much of the data on which the concepts embodied in the model were based and who were always happy to discuss any of the applications or practical aspects in early development of the model. In particular the author expresses his gratitude to Natsima Phunmuang, Kritsana Kritsanaphak and David Morrison.

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