Optimisation of Flotation Circuits With Large Flotation Cells

B K Gorain[1]

ABSTRACT

The recent trend of installing flotation cells with volumes as large as 200 m^3 in concentrators is driven mainly by economic considerations such as lower capital, operating and maintenance costs. Despite the economic advantages, the metallurgical benefits from these large cells have not yet been fully realised by the industry. This is mainly due to the lack of understanding of cell operation, viz cell mixing and hydrodynamics, gas dispersion and froth transportation behaviour in large cells. There is still no rational basis for selecting optimum cell operating conditions and down-the-bank operating profiles. The relationship between cell operating conditions and metallurgy is not well established. This has resulted in poor utilisation of cell capacity in many operations. The present trend of installation of fewer cells in a bank requires each cell to be operated very efficiently. Optimisation of cell operation provides an opportunity for significant metallurgical improvements in a flotation circuit with minimal capital expenditure.

Extensive work on optimising of cell performance has been carried out at various Teck Cominco operations in Canada, Peru and USA. Three different levels were involved in this work. The first level was aimed at identification of cell operating conditions for optimum mixing and hydrodynamics, gas dispersion, entrainment and froth flow behaviour in individual cells. The objective of the second level was to identify the optimum bank operating profile for cells down the bank in a circuit. The third level then focused on controlling these cell operating conditions and bank operating profiles for tighter control of concentrate grade and recovery for optimum bank metallurgy. This paper presents some of the major findings at all three levels and demonstrates the importance of each level in optimising the performance of flotation banks installed with large cells.

INTRODUCTION

Large flotation cells are increasingly being used in flotation circuits by the mineral industry. Flotation cell sizes have increased almost 100 times in the last 50 years or so with cells as large as 200 m^3 being marketed by flotation cell manufacturers. The economic benefits of these large cells have been realised by the industry: fewer cells, lower installed power and consumption, better control and smaller footprints. In the future these machines are expected to be even bigger to reduce capital and operating costs of treating large throughputs. The metallurgical benefits of larger cells, however, have not yet been fully realised by the industry. There have been several instances where new concentrators have required additional capacity to achieve design targets.

The recent trend has been to use fewer cells in a row superseding somewhat the conventional perception of the requirement for a minimum number of cells in a row to reduce short-circuiting and avoid adverse effects on recovery. Arbiter (1999) has pointed out that the short-circuiting phenomenon associated with a limited number of cells in a row is sound in theory but the dependence of flotation results on various other factors, in addition to mixing, precludes a fixed number of cells always being necessary. Irrespective of opinions, tolerance for errors is much lower for a row with fewer large cells than for a row with many smaller cells. Efficient operation of each large cell in a row is critical to achieve optimum metallurgy.

Very limited operating plant data on large cell operation has been published to date. Research efforts over the last decade at the Julius Kruttschnitt Mineral Research Centre (JKMRC), University of Cape Town (UCT) and McGill University under the AMIRA (previously Australian Mineral Industries Research Association, now AMIRA International) P9 project have provided significant insights into the understanding of large cell operation (Gorain *et al*, 1997a, b; Power *et al*, 2000; Deglon *et al*, 2000; Gomez *et al*, 2003). Burgess (1997) reported plant measurements of cell mixing and gas dispersion in a 100 m^3 Outokumpu cell and used these measurements to optimise its performance. Different flotation manufacturers have described the design principles and reported some operating data of their machines (Kallioinen *et al*, 1995; Jonaitis, 1999; Weber *et al*, 1999). These contributions have undoubtedly improved our understanding of large cell operation to a great extent, but still more work is required to obtain a rational basis for selecting cell operating conditions and to understand the reasons for poor froth recovery in large cells, especially of ultrafine (P_{80} <10 µm) and coarse particles (P_{80} >150 µm).

Optimising the operation of a large cell provides an opportunity for significant metallurgical benefits at minimal capital expenditure. These benefits can be realised more effectively by conducting cell optimisation work in a flotation circuit at three different levels:

Level 1: Identifying the optimum range of cell operating conditions in individual cells.

Level 2: Identifying the optimum bank operating profile.

Level 3: Controlling the bank operating profile for optimum metallurgy.

It is important to note that the variability of the plant feed and chemical conditions pose difficulties in obtaining relevant information at all three levels of cell optimisation work. It is almost impossible to eliminate these variations completely but their impact can be reduced to a great extent by carrying out randomised experimental designs with replicates. This work is best planned and carried out with the full cooperation of the operations staff. Batch flotation tests on the feed should be carried out to quantify the variations in feed floatability.

Level 1: Identifying the optimum range of cell operating conditions in individual cells

The selection of cell operating conditions, viz froth depth and air flow rate in different circuits is typically based on experience and pragmatism and not on any scientific rationale. Level 1 work involves identifying the range of cell operating conditions for optimising the four major subprocesses in a flotation cell:

Cell mixing

Residence time distribution (RTD) studies are carried out to understand the mixing and hydrodynamics behaviour in large cells. RTD studies help in determining of the mean cell residence time and in understanding the extent of plug and perfectly mixed flow in a cell. The amount of stagnant volume in cells can also be identified using RTD tests. RTD studies on large cells have been carried out at Codelco's Chiquicamata concentrator to compare the mixing behaviour in 148 m^3 Dorr-Oliver, 160 m^3 Wemco and 160 m^3 Outokumpu cells using both radioactive solid and liquid tracers (Lelinski *et al*, 2002). Yianatos *et al* (2003) reported that the mean residence time of solids is five per cent lower than that for liquid in a large cell. It was also shown in the same work that mineral residence times decreased with increasing particle size.

1. Teck Cominco Metals Ltd, Teck Cominco Research, PO Box 2000, Trail BC V1R 4S4, Canada. Email: barun.gorain@teckcominco.com

Nesset (1988) has used RTD tests to study the effect of baffling to improve flotation bank performance at Brunswick. Interpretation of the RTD test results to identify mixing problems can be found in the classic text by Levenspiel (1989).

Gas dispersion

The gas dispersion behaviour of a flotation cell can be characterised using measurements of superficial gas velocity (J_g), gas hold up (ε_g) and bubble size (d_b). Sophisticated techniques, robust for plant applications, have been developed by McGill University to carry out these measurements (Hernandez *et al*, 2001; Finch *et al*, 1999). These plant measurements can be used to calculate the bubble surface area flux (S_b), an indicator of the collection zone efficiency in flotation cells, which can be directly related to flotation kinetics (Gorain, 1998). Simple measurement of J_g in the large cells is useful for plant operators to diagnose cell operational problems. In the last decade, detailed gas dispersion measurements were carried out in several operations in Australia, South Africa and North America, under the AMIRA P9 project. It is interesting to note that the pioneering work on measurements of gas dispersion properties in plant cells was carried out by Jameson and Allum (1984) as part of the AMIRA P159 project. In recent years, Yianatos *et al* (2001) have carried out gas dispersion measurements in 42.5 m^3 cells in South American operations. All these have provided important insights into the gas dispersion behaviour in flotation cells.

Gangue entrainment

Characterisation of the entrainment behaviour of particles in large flotation cells is very useful for selection of operating conditions for optimum cell performance. The entrainment characterisation involves using size-by-size liberation or modal data of the feed, concentrate and tailings streams. The fully liberated non-sulfide gangue can be used as a tracer for entrainment characterisation, as the recovery of this species is not induced by the true flotation process. The degree of entrainment (ENT) in cells is normally calculated as the ratio between the mass transfer of entrained particles to the concentrate and the mass transfer of water to the concentrate (Lynch *et al*, 1981; Savassi *et al*, 1998). The degree of entrainment of an ideal tracer in the i^{th} size interval of the concentrate is calculated on the basis of the feed, as shown below:

$$ENT_{TR,i}^{feed} = \frac{\omega_{TR,i}^{sus,con}}{\omega_{TR,i}^{feed}} \quad (1)$$

where:

ω represents the concentration of solids (mass of solids per unit mass of water)

subscripts 'TR' and 'i' indicate the tracer and the i^{th} size interval, respectively

Froth flow behaviour

Understanding froth recovery is critical for optimum performance of large flotation cells. Previous plant measurements in large cells have suggested low froth recoveries (Gorain, 2000). The larger the cell, the more difficult it is for efficient transportation of froth. Froth recovery, R_f, is commonly used to quantify the efficiency of recovery of valuable minerals in the froth (Finch and Dobby, 1990; Gorain, 1998). Mathematically, R_f is represented as the ratio between the overall flotation rate constant (k) and collection zone rate constant (k_c).

$$R_f = \frac{k}{k_c} \quad (2)$$

The collection zone rate constants are difficult to obtain and require experiments at different froth depths and air flow rates. For a given air rate, the k_c value is obtained by regressing flotation rate constant against froth depth and then extrapolating this line to intercept the 'Y axis' (Feteris *et al*, 1987). Gorain (2000) has reported R_f values measured in plant cells. The R_f value is an important indicator of the efficiency of the froth phase of a flotation cell. Understanding the effect of cell operating conditions, viz air flow rate and froth depth on R_f, is useful for selection of the optimum range of cell operating conditions in a cell. The effects of J_g and froth depth on R_f can be represented conveniently using froth residence time of air, T_f, which has been explained in a previous publication (Gorain and Stradling, 2002). The modelling of froth transportation in industrial flotation cells has been reported recently (Zheng *et al*, 2004; Zheng and Knopjes, 2004).

Level 2: Identifying the optimum bank operating profiles

Level 1 optimisation work helps in identifying the optimum range of cell operating conditions in individual cells down the bank. Depending on the metallurgical target of a bank, the operating conditions of the individual cells in a bank may need to be adjusted or fine-tuned to obtain the optimum bank operating profile. Level 2 optimisation work involves identification of the optimum combination of cell operating conditions viz air flow rate and froth depth for cells down the bank. There is limited work published in this area. An air distribution profile was successfully utilised at Noranda's Brunswick mine for optimisation of their zinc cleaning stage (Cooper *et al*, 2004). An increasing air distribution profile at Brunswick gave the best metallurgy due to higher selectivity of sphalerite against non-sulfide gangue in the first few cells. This success was replicated for all four stages in the zinc cleaner circuit at Brunswick.

Optimisation of bank operating profiles has the potential for significant metallurgical improvements using the existing cell capacity by proper adjustment of the operating conditions for individual cells in a bank. This simple yet powerful strategy has immense potential for any flotation concentrator with or without large cell operation. It is important to note that the variability of circuit performance with time (autocorrelation) is an important consideration in comparing various bank operating profiles. This is because the relatively small metallurgical improvements being sought are easily obscured by large natural variation in the data due to changes in ore mineralogy, operating conditions and other factors. Details of the significance of autocorrelation on plant trials are described in Napier-Munn and Meyer (1999).

Level 3: Controlling the bank operating profile for optimum metallurgy

Once the optimum bank operating profiles are identified using Level 1 and Level 2 optimisation work, it is important to have a proper control of the optimum bank operating profile. This is critical to ensure a tighter control of bank concentrate grade and recovery. Different strategies can be applied to control bank operation. In the recent years, froth vision based on-line automatic control has shown promise in improving the performance of banks with large cells. JKFrothCam, marketed by JKTech in Australia, was successfully used at Minera Escondida in Chile in 100 m^3 Outokumpu tank cells which increased copper recovery by 1.2 per cent when compared with manual operation (Kittel *et al*, 2001). FrothMaster™, marketed by Outokumpu, showed an increase in copper and gold recoveries of about 2.45 per cent and 5.65 per cent, respectively, at Cadia Hill Gold Mine, in rougher banks using 150 m^3 Outokumpu cells (Brown *et al*, 2001). The Metso VisioFroth system has increased copper and gold recoveries by more than three per cent at PT Freeport Indonesia's Grasberg concentrator (Mular and Vien, 2002).

All these systems utilise video cameras to capture froth images that are analysed by specialised software to produce a number of froth vision parameters. The froth texture, bubble size and velocity are the major parameters used in the vision systems, but the relative emphasis on the parameters differs from one system to another. The froth parameters are associated with metallurgical data such as grade, recovery obtained from model estimates or recovery obtained from OSA assays. The data obtained from a froth vision system are sent to an expert system based on the same logic that an operator would use to perform a task. Based on deviation from control set point, the controller in the PLC or DCS calculates new set points for froth level, air flow rate and/or reagent dosages, depending on the control logic used by the expert system. The plant reacts to the new set points and the loop continues till the desired set point is achieved.

Teck Cominco Research has taken a major initiative to apply the present understanding on large cell operation to its various operations using the three levels of optimisation explained above. Extensive work was carried out at various operations in which Teck Cominco has interest, viz Red Dog mine in Alaska, USA, Highland Valley Copper in BC, Canada and Compania Minera Antamina in Peru. This paper aims to demonstrate the importance of the three levels of optimisation in improving the performance of flotation circuits.

PLANT TESTWORK

Teck Cominco Research has conducted detailed measurements in various flotation circuits at Highland Valley Copper, Compania Minera Antamina and Red Dog operations. Most of the work was carried out on single rows in rougher, scavenger and cleaner circuits. The main cell operating variables used in this study were froth depth, air flow rate and in some cases, frother dosage. Experimental design techniques were used to select various combinations of these variables for these studies. Typically two to four level factorial designs were used. A summary of the measurements and plant surveys carried out are as follows:

- Gas dispersion measurements viz bubble size, superficial gas velocity and gas hold up were carried out using sensors developed at McGill University and under the AMIRA P9 project. These measurements were carried out at various cell operating conditions and in various locations in each cell. These locations were selected based on several initial measurements carried out at different cell depths and horizontal distances from the centre of the cell.
- Froth vision measurements were made using a mobile Metso VisioFroth system. This system provides real time measurement of froth velocity, bubble size, colour, stability, etc. This system was operated continuously during the gas dispersion measurements to relate these results to the froth vision parameters.
- Metallurgical samples of individual cell feed, concentrate and tailings were taken for each cell operating condition. In addition, cell-by-cell surveys were also carried out for each bank to obtain its metallurgical profile. Importantly, many repeats were carried out to obtain statistical significance of the results.
- Residence time distribution (RTD) tests were carried out with lithium chloride in individual cells and banks to understand the mixing and hydrodynamic behaviour of the cells and banks.
- Various cell operating bank profiles were tested for each circuit. Cell-by-cell metallurgical sampling of the banks was carried out for each of the profiles tested. Comparison of the performance of the various bank profiles is difficult due to bank autocorrelation effects and feed variability. In some cases, the profile tests were carried out for a few days and sometimes months to differentiate the metallurgical performance of the various bank profiles.
- In some cases, specialised batch flotation tests were carried out to assess the variability of the plant feed floatability during the period of plant trials.

RESULTS AND DISCUSSION

Some important findings from the three levels of optimisation will be presented in this paper with emphasis on the significance of each level in optimising the performance of flotation cells and banks. The results and discussion presented here demonstrate the value of conducting cell optimisation work in operations with significant potential for improvements in the plant.

Level 1: Identifying the optimum range of cell operating conditions in individual cells

Cell mixing

Figures 1a and b show the residence time distribution profiles for the copper and zinc rougher banks at Antamina and the lead rougher bank at Red Dog, respectively.

Table 1 shows the cell mixing parameters for the copper and rougher banks at Antamina and for the lead rougher bank at Red Dog.

Both Figures 1a and b and Table 1 show that RTD studies provide useful information such as mean residence time, quantification of plug flow, vessel dispersion number and stagnant volumes in cells and banks. In operations where slurry flow metres are not installed, RTD studies are necessary for estimation of cell residence time. The estimated residence times shown in Table 1 were used for calculation of flotation rate constants of minerals. The RTD profiles for the Antamina work suggested higher cell dead volumes of 13 to 15 per cent in the first three cells of M1 copper rougher bank. The vessel dispersion number values, indicating the extent of axial dispersion in the banks, were remarkably higher for the M1 copper rougher bank. For the Red Dog ore, very high dead volume in the lead rougher banks was observed. This was mainly due to accumulation of sand or gravel in the cell and build-up of solids around launders. The total dead volume in cells 3, 4 and 5 in the Red Dog lead rougher bank was estimated to be at least 20 per cent. The dead zones result in improper cell mixing and poor utilisation of existing cell capacity. These results demonstrate the importance of RTD studies in identification of mixing problems in cells and banks.

Gas dispersion

Table 2 shows the operating range of J_g, d_{32}, S_b and T_f values for cells in different circuits at Antamina, Red Dog and HVC concentrators. It is interesting to note that larger cells are operated at higher J_g values than the smaller cells. This could be one of the factors for larger bubbles observed in the larger cells. Figures 2a and b show the bubble volumetric and area distribution, respectively, for cells of different sizes.

Figures 2a and b indicate that more than 70 per cent of the total volumetric air fed to the OK130 cells at Antamina was utilised in making bubbles greater than 2 mm as compared to about 25 per cent for OK50 cells at Red Dog and DR100 cells at HVC. Figure 2b shows that about ten per cent of the total bubble surface area was contributed by bubbles greater than 2 mm for the OK50 and DR100 cells at Red Dog and HVC, respectively, whereas for the OK130 cells at Antamina, about 55 per cent of the total bubble surface area was contributed by bubbles greater than 2 mm. This study showed that there is significant

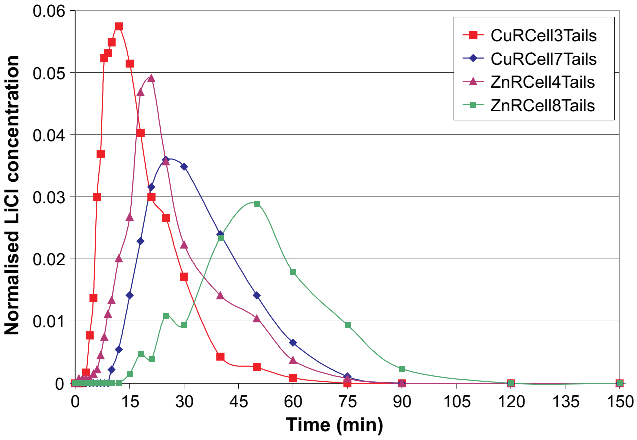

FIG 1A - Residence time distribution profiles for the copper and zinc rougher banks at Antamina.

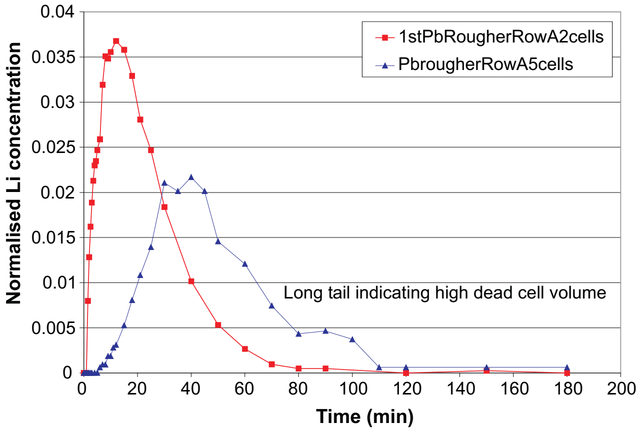

FIG 1B -Residence time distribution profiles for the Red Dog lead rougher bank.

TABLE 1

Hydrodynamics and mixing behaviour of Antamina OK130 and Red Dog OK50 cells.

Row A	Res time (min)	Plug flow %	Vessel dispersion	Dead volume%
Red Dog OK50 1st lead roughers (2 cells)	22.8	12	0.30	10
Red Dog OK50 lead rougher cells (5 cells)	53.4	19	0.22	15
Antamina OK130 Cu rougher (3 cells) M1 ore	17.5	14.8	0.30	15
Antamina OK130 Cu rougher (7 cells) M1 ore	36.0	25.3	0.30	10
Antamina OK130 Zn rougher (3 cells) M4a ore	31.6	16.0	0.14	10
Antamina OK130 Zn rougher (7 cells) M4a ore	51.1	37.3	0.05	8

TABLE 2

Results of the gas dispersion measurements at various Teck Cominco concentrators.

Bank	Cell	J_g (cm/s)	d_{32} (mm)	S_b (m^2/m^2 s)	T_f (s)
Antamina					
Cu Rougher	OK130	0.7 to 1.7	1.9 to 3.3	14 to 29	10 to 42
Zn Rougher	OK130	0.6 to 1.5	1.8 to 3.1	15 to 30	22 to 56
Red Dog					
Pb Rougher	OK50	0.3 to 0.7	1.1 to 1.6	15 to 30	40 to 80
Zn Rougher	OK50	0.3 to 1.0	1.2 to 1.7	15 to 40	15 to 75
Zn Cleaner/Retreat	OK50/MX14	0.3 to 0.9	1.0 to 1.4	17 to 46	40 to 130
HVC					
Moly Rougher	DR100	0.2 to 0.3	0.9 to 1.4	8 to 22	40 to 70
Moly Scavs	DR100	0.2 to 0.4	1.1 to 1.8	7 to 19	25 to 50
Moly 1st Cleaner	DR100	0.3 to 0.4	0.9 to 1.2	15 to 18	35 to 44

opportunity for optimisation of the bubble size distribution in the OK130 cells at Antamina for better utilisation of cell capacity. Making the bubbles smaller for Antamina has the potential for increasing the S_b values and improving the flotation rate constants of the minerals in the OK130 cells.

Entrainment

Figure 3a shows a typical relationship between the degree of entrainment (ENT) of tracer and particle size for cells down-the-bank in the zinc rougher circuit at Red Dog.

Based on this study, it was estimated that about 69 per cent of the non-sulfide gangue, about 57 per cent of the pyrite in the concentrate and about one per cent of the sphalerite were recovered by entrainment in the rougher concentrate. The entrainment of non-sulfide gangue in cells down-the-bank of a rougher circuit is shown in Figure 3b. Figure 3b shows that cells 1 - 4 in the zinc rougher circuit entrained a more significant proportion of non-sulfide gangue than in cells 5 - 8. This demonstrates that entrainment studies provide useful information to diagnose problematic cells in a bank. Detailed studies in this example have shown that the high recoveries of non-sulfide gangue and pyrite are due to low froth residence times in the cells which in turn points to inadequate flotation capacity. Since this study was completed, Red Dog has increased the flotation capacity in the rougher circuit (Lacouture and Hope, 2002).

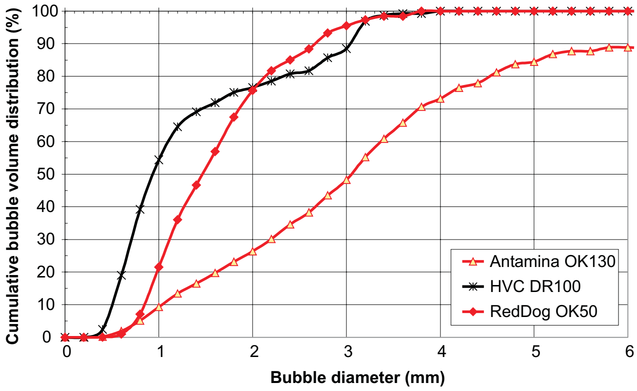

FIG 2A -Comparison of the bubble volume distribution of bubbles for cells of different sizes at Antamina, HVC and Red Dog.

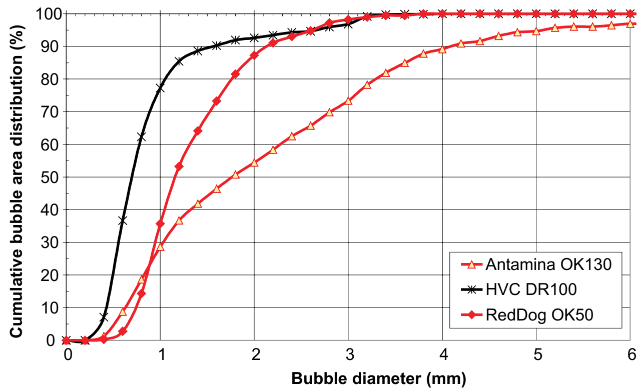

FIG 2B - Comparison of the bubble surface area distribution for cells of different sizes at Antamina, HVC and Red Dog.

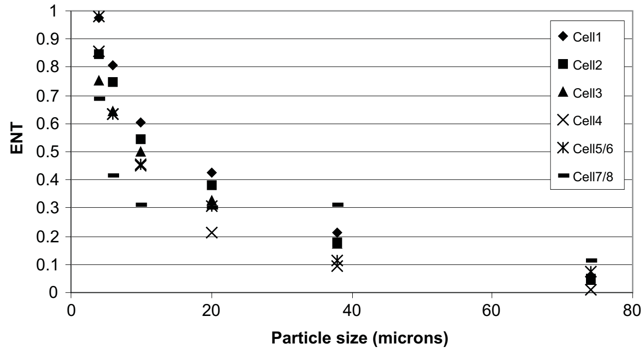

FIG 3a - Relationship between degree of entrainment of tracer and particle size for cells down-the-bank in a zinc rougher circuit.

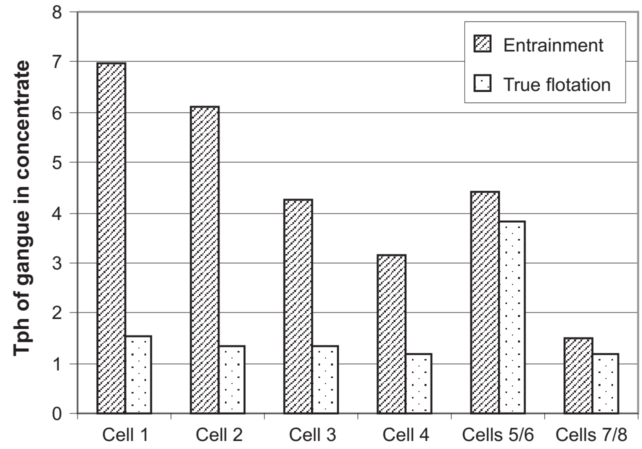

FIG 3b - Tonnage of non-sulfide gangue recovered in a zinc rougher circuit by entrainment and by true flotation.

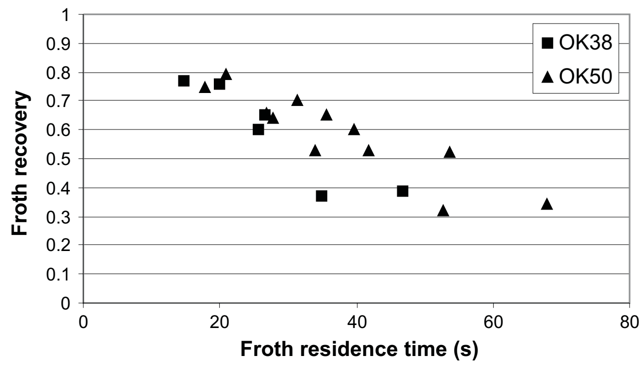

FIG 4 - The effect of froth residence time on froth recovery.

Froth flow behaviour

Gorain (2000) has previously reported that air specific froth tonnage T_{as} is an important criterion to characterise froth transportation behaviour. Though a very useful concept, air specific froth tonnage, T_{as}, is difficult to implement for optimising cell performance in plants. This is because of the difficulty in measuring cell concentrate flow rates. In addition, T_{as} is not an independent entity and is dependent on cell operation. A strong relationship was reported between air specific froth tonnage and froth residence time, T_f, for the lead and zinc rougher circuits at Red Dog. Froth residence time, T_f, is dependent only on froth depth and superficial gas velocity, which are easy to measure and vary in plants. The importance of T_f has previously been emphasised by Heiskanen and Kallioinen (1993) in optimising cell performance.

Figure 4 suggests a strong correlation between R_f and T_f, which is independent of cell type. The relationship appears to be linear but an exponential relationship is possible at lower T_f values as suggested elsewhere (Mathe *et al*, 1998). In this work, lower T_f values have not been used because of difficulties in operating plant cells at shallow froth depths.

The effect of cell operating conditions on metallurgy

Table 3 shows the plant cell froth residence times that result in the best selectivity, recovery and optimum metallurgy for different cells, as well as the froth residence time values for the present cell operation. It is to be emphasised that the optimum metallurgy shown in Table 3 is based on the requirements of the individual cells, in terms of both selectivity and recovery, to meet the metallurgical target of the circuit in which the cell is installed.

At Red Dog, the best recovery values were obtained at froth residence times ranging from 30 to 40 seconds in the zinc rougher and 60 to 70 seconds in the zinc cleaner cells. The best *selectivity* values were obtained at froth residence times around 60 and 70 seconds for the same rougher and cleaner cells respectively. This is not unexpected as greater froth residence time values are associated with deeper froth and lower air rate, both of which increase selectivity. The optimum froth residence time would seem to be from 40 to 50 seconds for roughing and

TABLE 3

Froth residence times for present cell operation and for different levels of metallurgical performance.

Cell type	Circuit	Presently operating	Best selectivity	Best recovery	Optimum metallurgy
Cell 2 OK50	Zn Rougher	29	60	30	50
Cell 3 OK38	Zn Rougher	31	60	40	45
Cell 6 OK38	Zn Rougher	31	60	30	45
Cell 7 MX14	Zn Rougher	55	60	40	45
Cell 2 OK38	2nd Zn Cleaner	45	75	60	65
Cell 3 MX14	2nd Zn Cleaner	54	75	70	70
Cell 3 OK50	Pb Rougher	42	50	35	45
Cell 4 OK38	Pb Rougher	53	60	30	40

65 to 70 seconds for cleaning. The operating cell froth residence times were much lower than those required for optimum performance of both zinc roughers and cleaners, but were close for the lead roughers. This study showed that froth residence time provides a rational basis for selecting froth depth and air rate in a flotation cell for optimum flotation performance. The importance of froth residence in selection of cell operating conditions was also confirmed at Antamina and HVC operations.

Level 2: Identifying the optimum bank operating profile

Grade-recovery plots for cells down the bank

Grade-recovery plots for the copper rougher bank at Antamina and the lead rougher circuit at Red Dog are shown in Figures 5a and 5b, respectively. It is interesting to note that cells down the bank have their unique grade-recovery profiles. These plots provide important information on the realistic range of grades and recoveries that are possible in cells down the bank. This in turn can be used to set metallurgical targets for each cell down the bank. This is critical for large cell operation as there are only a few cells in a bank to achieve the target.

Cell operating profiles in a bank

The importance of cell operating profiles is illustrated in Figures 6a and b. The operating profiles tested at Red Dog and Antamina were identified from various tests carried out by changing the cell operating variables like air flow rate and froth depth. Figure 6a shows that profile four gave the best metallurgy with an estimated two to three per cent increase in lead concentrate grade at Red Dog. Figure 6b shows the effect of four different cell operating profiles in Bank C of the copper rougher circuit at Antamina, compared with the normal cell operating profile in Bank B, operating in parallel with Bank C. As evident from Figure 6b, profile two gave the best metallurgical results for the copper rougher bank whereas profile four gave the worst performance. Details of the cell operating profiles are not discussed in this paper due to confidential reasons.

Level 3: Controlling cell operating conditions and bank gas dispersion profile

Extensive tests at various Teck Cominco operations showed that froth velocity is a simple yet effective froth vision parameter. Other froth vision parameters like bubble size, coalescence, stability and colour could not be related to cell operating conditions. Figures 7a and b show the relationship between superficial gas velocity and froth velocity for the lead and zinc rougher cells at Red Dog and Antamina, respectively.

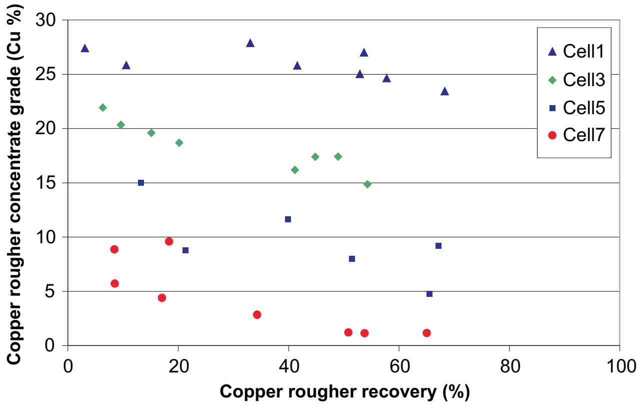

FIG 5A - Grade recovery curve for the cells in the copper rougher bank at Antamina for M4a ore.

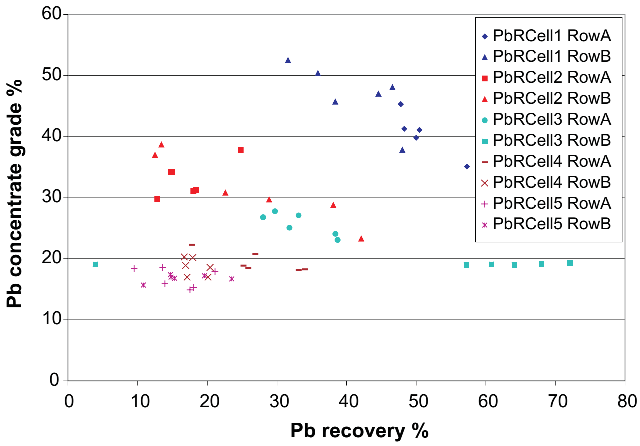

FIG 5B - Relationship between lead grade and recovery for cells in the first and second lead rougher banks at Red Dog.

The relationships between froth velocity and cell metallurgy down-the-bank were investigated for different flotation circuits at various operations. Figures 8a and b show the relationship between froth velocity and concentrate grade for the lead rougher circuit at Red Dog and for the zinc rougher circuit at Antamina, respectively. As may be seen, a strong correlation between froth velocity and concentrate grade in indicated. Repeated studies have confirmed the importance of froth velocity as a parameter to control concentrate grade. It is important to remember that these relationships will vary with feed. Developing a knowledge base is critical for development of a robust control strategy.

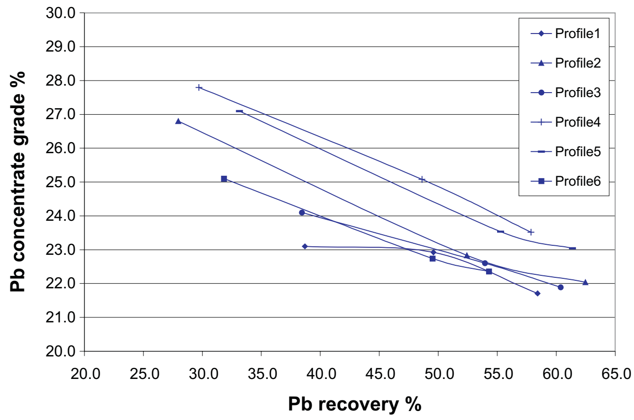

FIG 6A - Metallurgical comparison of the different profiles tested in lead rougher Circuit at Red Dog.

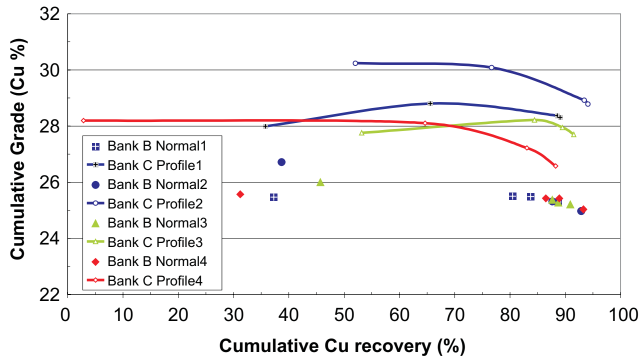

FIG 6B - Metallurgical comparison of the different profiles tested in line C with that of the normal operating profile in line B for the M4A copper rougher bank at Antamina.

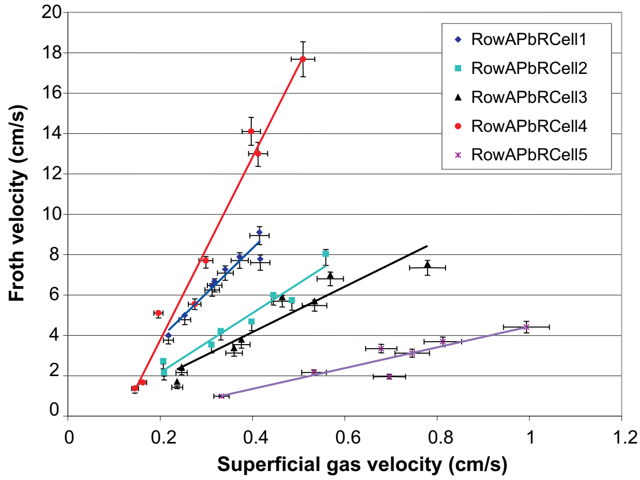

FIG 7A - The relationship between superficial gas velocity and froth velocity in the lead rougher circuit at Red Dog.

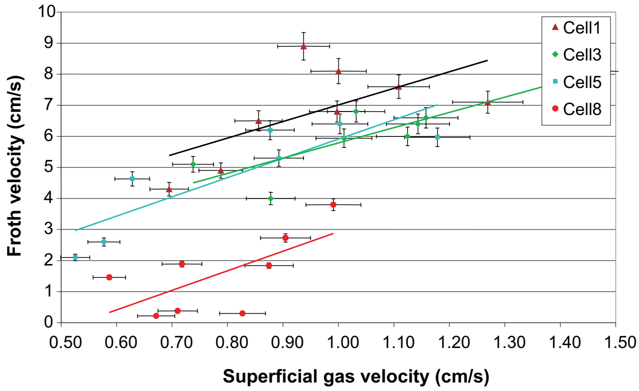

FIG 7B - The relationship between superficial gas velocity and froth velocity in the zinc rougher circuit at Antamina.

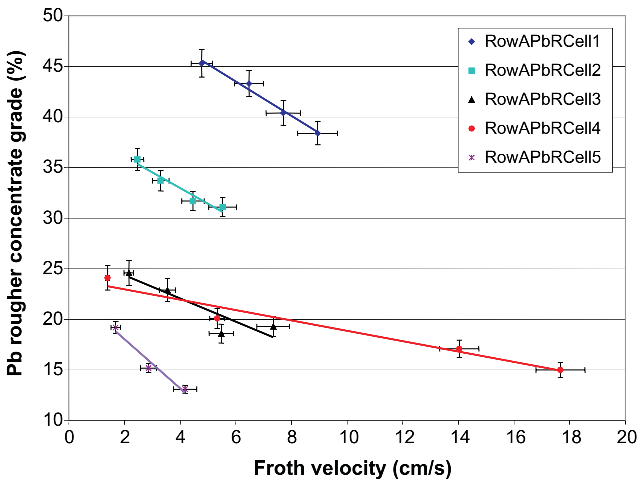

FIG 8A - The relationship between froth velocity and lead rougher concentrate grade in the lead rougher bank at Red Dog.

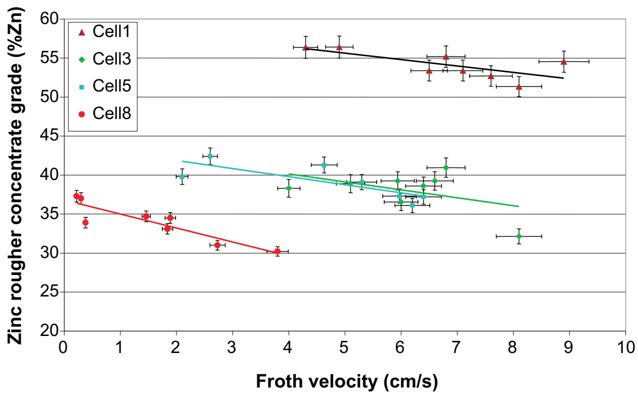

FIG 8B - The relationship between froth velocity and zinc rougher concentrate grade in the zinc rougher bank at Antamina.

CONCLUDING REMARKS

Significant improvements in the metallurgical performance of flotation banks with large cells are possible through a three-level approach to cell optimisation. The first level is aimed at identifying the range of cell operating conditions for optimum mixing, gas dispersion, entrainment, froth flow behaviour and metallurgical performance in individual cells. The second level is aimed at identifying the optimum bank operating profile for cells down the bank in a circuit. The third level is focussed at controlling these cell operating conditions and bank operating profiles for tighter control of concentrate grade and recovery. It is important to note that higher metallurgical benefits are obtained once the findings of the three-level studies are integrated to develop an effective strategy for control of the optimum bank operating profile. Even though this work is greatly leveraged on the high level of technical expertise, transfer of technology is critical for realising the full metallurgical benefits.

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