Selection of Cell Operating Conditions to Optimise Performance of Flotation Circuits with Large Cells

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ABSTRACT

An extensive study was carried out to evaluate the performance of Outokumpu OK50, OK38 and Maxwell MX14 cells at Cominco's Red Dog flotation concentrator in Alaska. Tests were conducted by varying operating conditions of the cells installed in different circuits in the plant. The hydrodynamic and gas dispersion properties were measured and the metallurgical performance of the cells was evaluated for each condition used in the tests. This paper highlights the importance of superficial gas velocity, bubble surface area flux, gas hold-up, power consumption, residence time distribution and froth flow behaviour in optimising cell performance. The findings of these tests provide valuable insights into selection of the best cell operating conditions to optimise circuit performance.

INTRODUCTION

The recent trend in the mineral processing industry has been to install large flotation cells. Cells as large as 150 m^3 are currently being installed to reduce the capital and operating costs of the flotation operation. Flotation manufacturers claim that the large cells are economical without having to compromise metallurgy. However, the full benefits of large cells are still being ascertained by the industry. Installation of large cells reduces the number of cells in a bank. This means that there is a greater probability of short-circuiting and therefore reduced effective residence time. In addition, effective froth transportation in larger cells is a major challenge. Further studies are required to understand the effect of operating conditions on collection zone kinetics and froth recovery in large cells. This lack of understanding may offset some of the benefits of larger cells. Further studies on large cells are therefore essential.

An extensive study on cell hydrodynamics and gas dispersion properties was carried out at Cominco's Red Dog Mine in Alaska, which is the largest zinc concentrate producer in the world. The aim of this investigation was to evaluate the operating and metallurgical characteristics of flotation cells to optimise the performance of the flotation circuit. This paper presents some of the important findings of this investigation.

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TEST CELLS AND MEASUREMENT

A flotation test campaign was carried out at the Red Dog concentrator in June 1999. Most of the mechanical cells in the flotation circuit are of tank cell design. There are three different types of cells installed in different circuits: Maxwell MX14, Outokumpu OK38 and OK50 cells. Details of the test cells selected for the testwork are shown in Table 1.

All the test cells were operated at different conditions. Four different air flow rates were selected for each cell[†]. The values of air flow rates selected in each cell were different, depending on the cell type, cell duty and the froth depth setting. It was only possible to vary froth depth in two cells, viz cell numbers 2 and 3 in the zinc rougher circuit. Three different froth depths were selected in these two cells: one shallow, one intermediate and one deep. The shallow froth depth was selected by adjusting the froth level such that there was no pulping of the concentrate at the different air flow rates selected. The deep froth depth selected was the maximum limit the froth weir could be moved downwards. The intermediate froth depth was selected depending on the size of the weir bars available to achieve a level between the shallow and the deep froth depths. For each of the froth depths selected for cells 2 and 3, four different air flow rates were used, with a total of 12 different combinations of cell operating conditions for each cell.

Physical measurements

The following physical measurements were carried out in each cell at different cell operating conditions shown in Table 1:

1. Superficial gas velocity (J_g) was measured at different locations in the cells using a long perspex tube closed with a pneumatically-operated pinch valve at the bottom of the *tube* (Gorain *et al*, 1996). These J_g values were used to estimate the air flow rates in each cell.

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† The only exception was cell 2 for which six different air rates were selected.

TABLE 1
Details of the operating conditions for the test cells.

Circuit	Cell no	Cell Type	D	N	FD	Air flow meter reading	Q
Zn Rough	2	OK50	90	138	10, 20, 35	30, 50, 70, 90, 140, 240	0.8, 1.4, 2.2, 3.7, 4.9, 6.7
	3	OK38	90	138	25, 38, 48	130, 180, 240, 300	3.1, 4.5, 4.9, 5.5
	6	OK38	90	138	45	90, 150, 210, 270	2.3, 3.9, 4.2, 6.2
	7	MX14	105	120	38	80, 130, 180, 230	2.4, 3.6, 4.4, 4.9
2nd Zn Clean	2	OK38	90	138	54	120, 180, 240, 300	3.4, 4.2, 4.7, 5.5
	3	MX14	105	120	59	80, 160, 240, 300	2.9, 3.6, 4.4, 5.8
Pb Rough	3	OK50	90	138	35	60, 100, 120, 140	3.0, 3.9, 4.9, 5.9
	4	OK38	90	138	43	60, 100, 140, 180	2.7, 3.6, 4.4, 5.3

D = impeller diameter (cm), N = impeller speed (rpm), FD = froth depth (cm), Q = air flow rate (m^3/min)

2. Gas hold-up (ε_g) was measured at different locations using a gas hold-up device, similar to the one in Gorain *et al* (1995).

3. Froth depth was measured using a conductivity meter, similar to the one in Finch and Dobby (1990).

4. Residence time distribution (RTD) tests were conducted using LiCl solution to estimate the effective cell residence time and mixing characteristics.

5. Power consumption was measured using a power meter.

Metallurgical sampling

Samples of the feed, concentrate and the tailings were collected for each of the test conditions shown in Table 1. The procedure adopted for sampling involved setting a froth depth by adjusting the weir height. Then, with the impeller speed fixed, the air flow rate was set to the desired value, and conditions in the cell were allowed to stabilise for three - four times the average cell residence time. Once steady state was achieved, the samples of the feed and tailings streams were collected using a dip (thief) sampler and the concentrate samples were collected with a lip sampler. Each type of sampler was fitted with a long handle to enable sampling of the streams. The air flow rate was then set to a new value and the process was repeated. After measurements were taken at all air flow rates, the froth depth was changed to the new desired value and the whole procedure was repeated from the beginning until tests at all the operating conditions were completed. Some of the tests at different cell operating conditions were repeated to check the reproducibility of the results. Extra care was taken to ensure that the flotation plant feed was consistent and the flotation circuits were steady during the metallurgical testwork.

All samples of feed, concentrate and tailings were weighed, filtered, dried and weighed again to calculate per cent solids. Subsamples were sent to the Red Dog laboratory for the analysis of zinc, lead, iron and silica. Subsamples were stored in case further analysis was required.

Data analysis

The 'JKSimMet' software was used to mass balance the feed, concentrate and tailings stream for each operating condition. Zn, Pb, Fe, SiO_2 and per cent solids were used as components for mass balancing. Excellent agreement was found between the calculated and observed grade for the feed, concentrate and tailings. The overall recovery values of Zn, Pb and FeS_2 were calculated for each test condition. The selectivity values were calculated as the ratio of the recovery of valuables[‡] and the recovery of pyrite.

RESULTS AND DISCUSSION

The hydrodynamic and gas dispersion properties studied in different cells are:

1. superficial gas velocity, J_g;

2. bubble surface area flux, S_b;

3. gas hold-up, ε_g;

4. power consumption, P_o;

5. residence time distribution, RTD; and

6. froth flow behaviour.

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‡ Recovery of lead and zinc values were used in the lead and zinc circuits, respectively.

The operating ranges of these properties in the cells and their implications on cell and circuit performance will be discussed in this section.

Superficial gas velocity (J_g)

The superficial gas velocity J_g values, in general, were found to increase with increase in air flow rate. It was also found that there is a maximum air rate, or J_g, above which the cell froth had a boiling appearance commonly known as cell flooding. This is due to the inability of the impeller to disperse air at high air rates and large slugs of undispersed air bubbles appeared in the froth close to impeller shaft resulting in cell flooding (Gorain *et al*, 1996). Measurements of J_g values showed that the OK50 cells could be operated up to a J_g of about 2.2 cm/s without flooding compared to the J_g values of 1.5 cm/s and 1 cm/s for the OK38 and MX14 cells, respectively. This suggests that the operating range of air rates for the OK50 cells is greater than for the OK38 and MX14 cells.

Bubble surface area flux (S_b)

The operating values of bubble surface area flux were calculated for different cells using a model developed by Gorain *et al* (1999). The S_b values, in general, were found to increase with increase in air rate. The maximum S_b values obtained for the OK50, OK38 and MX14, at their highest operating air flow rate, were about 80, 64 and 49 m^2/m^2 s, respectively. This confirms the finding that the OK50 cells could be operated at a much greater range of cell operating conditions than the OK38 and MX14 cells. Nonetheless, the S_b values were similar at any particular value of air rate, as long as the selected air rate is within the normal operating range of individual cells. The S_b values were found to be around 48 m^2/m^2 s for all cells at the J_g value of 1 cm/s. This suggests that all cells are equally efficient in generating bubble size and therefore S_b at low rates, when the air is dispersed uniformly by the impellers. At high air rates, however, the dispersion ability of the impellers is different, resulting in a different range of S_b values for different cells. Large slugs of air with large bubbles were found on the froth surface of the MX14 cells at higher air rates.

Gas hold-up (ε_g)

Gas hold-up was measured for the zinc rougher cells at their normal operating conditions and the measured values are shown in Table 2. The values were higher for the OK cells ranging from 6.5 to 8.5 per cent whereas the values for the MX14 cells were around five per cent. This difference can be attributed to the high gas dispersion ability of the OK cells compared to the MX14 cells.

Power consumption (P_o)

The total power consumed by each cell was measured using a hand-held power meter. These values are shown in Table 2. The actual impeller power input, after allowing for power losses such as bearing friction, pulley transmission, etc could not be estimated as 'no-load' power data were not measured. Table 2 shows that the power consumption of OK50 and OK38 cells were greater than the MX14 cells. This may partly account for the high gas dispersion ability of the OK50 and OK38 cells as compared to the MX14 cells. A decrease in air rate resulted in an increase in power consumption and at low air rates the power consumed was close to the capacity of the motor, especially for the MX14 cells, and to a lesser extent for the OK cells. This means that the cell motor will likely trip when the cell air is cut off. This could cause solids to settle on the cell bottom, making it difficult to re-start the impeller.

TABLE 2

Measured and calculated values of different physical operating variables for normal plant operation (indicates calculated values).*

Circuit	Cell no	Cell Type	Jg	Q*	FD	S_b*	FRT*	ε_g	P_o
Zn Rough	2	OK50	1.22	6.0	35	55	29	6.5	40
	3	OK38	1.24	5.5	38	66	31	8.5	42
	6	OK38	1.40	6.2	45	63	31	7.2	37
	7	MX14	0.69	3.1	38	37	55	4.8	26
2nd Zn Clean	2	OK38	1.20	5.3	54	55	45	-	40
	3	MX14	1.10	4.9	59	49	54	-	20
Pb Rough	3	OK50	0.83	4.1	35	42	42	-	45
	4	OK38	0.81	3.6	43	41	53	-	40

J_g = superficial gas velocity, Q = air flow rate (m^3/min), FD = froth depth (cm), S_b = bubble surface area flux (m^2/m^2 s), FRT = froth residence time (s), ε_g = gas hold-up (%), P_o = power consumption (kW)

TABLE 3

Mixing parameters for the cells in the zinc circuit.

Cell No	Cell type	Experimental τ (min)	Model fitted τ (min)	P %	Cell Dead Volume (%)
Zn Rough Cell2	OK50	3.87	3.62	3.0	12.8
Zn Rough Cell3	OK38	4.53/4.38	5.07/4.16	5.2/4.9	13.5/12.9
Zn Rough Cell6	OK38	4.90	4.48	2.4	13.2
Zn Rough Cell7	MX14	6.04	5.88	2.9	13.1
2nd Zn Clean Cell2	OK38	1.70/2.41	1.5/2.22	5.9/6.1	12.7/12.7
2nd Zn clean Cell 3	MX14	3.12	2.83	1.2	13.4

Residence time distribution (RTD)

The RTD tests were conducted for each of the test cells in the zinc circuit using LiCl salt solution. About 750 grams of LiCl salt was dissolved in four litres of water by mechanical agitation to conduct the RTD test for each cell. The stimulus-response method was used by introducing the LiCl salt solution in the cell feed box. Samples of the tailings from the tails box were taken at regular intervals to obtain the exit age distribution E(t), or the RTD of fluid (Levenspiel, 1972). The LiCl solution tests were based on the assumption that the solids in the cell approximate the behaviour of the liquid phase. This assumption has been shown to be valid for fine particles (Finch and Dobby, 1990).

The residence time distribution profiles were plotted for each cell tested. The E(t) distribution for each plot was obtained by normalising the distribution curve of lithium concentration against time so that the total area under the curve is unity. A CSTR model was fitted to the RTD profile for each of the plots. This shows a close to perfectly mixed regime with a small component of plug flow. The plug flow component is more evident when the time record of the tracer in the exit stream, known as F(t) distribution (Levenspiel, 1972), is plotted. The following equation was fitted to the F(t) distribution to obtain the parameters, viz fraction of plug flow (P) and mean residence time (τ):

$$F9t) = 1 - \exp\left[-\left(\frac{1}{1-P}\right)\left(\frac{t-P\tau}{\tau}\right)\right] \qquad (1)$$

The parameter values P and τ are shown in Table 3 along with the fraction of the cell 'dead volume' obtained from the F(t) distribution. Table 3 shows that the parameter P values, which represent the fraction of the plug flow, are greater for OK38 cells compared to the MX14 cells, suggesting relatively higher plug flow characteristics for the OK38 cells. This may be because the impeller is suspended higher above the tank bottom in the OK38 cell. The dead volume in the cell was found to be around 13 per cent for all cell types, which could be due to settling of solids on the bottom of the tank.

It is also evident from Table 3 that the cell residence time increased down the bank for both the rougher and cleaner cells. This is because of strong pulling rate of concentrates in the rougher and cleaner cells. More importantly, the liquid residence times in the cleaner cells were almost half of those in the rougher cells. This may partly account for the low unit recoveries in the cleaner cells compared to the rougher cells observed in the plant. The RTD studies indicate that the total residence time in the cleaner circuit is lower than in the rougher circuit. The ratio of residence times for the cleaner and rougher circuits is therefore less than unity, which may have important implications on the overall plant recovery and selectivity. Sutherland (1981) has shown that in many concentrators the optimum recovery and selectivity is obtained when the ratio of residence times for the cleaner and rougher circuits is equal to one. For the Red Dog zinc flotation circuit the ideal ratio is expected to be greater than one because regrinding of the rougher concentrate will slow down the flotation kinetics in the cleaners. These observations suggest that consideration should be given to increasing the residence time in the zinc cleaner circuit.

Froth flow behaviour

Froth recovery, R_f, is commonly used to quantify the efficiency of recovery of valuable minerals in the froth (Finch and Dobby, 1990; Gorain *et al*, 1998). Mathematically, R_f is represented as the ratio of the overall flotation rate constant (k) and collection zone rate constant (k_c). The k values can be directly obtained from recovery (R) and cell residence time (τ) using a perfectly mixed model given by Levenspiel (1972) as shown below:

$$k = \frac{R}{(1-R)\tau} \qquad (2)$$

and

$$R_f = \frac{k}{k_c} \qquad (3)$$

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 was obtained by regressing flotation rate constant against froth depth and then extrapolating this line to intercept the 'Y axis' (Feteris *et al*, 1987).

In this work the R_f values were calculated for zinc rougher cells 2 (OK50) and 3 (OK38) for which rate constant data at different froth depths were collected. Table 4 shows the estimated k_c and R_f values for cells 2 and 3. The range of R_f values obtained for the OK50 cell was found to be greater than for the OK38 cell. R_f values as high as 90 per cent were obtained for the OK50 cell, which may be due to better design of its launders resulting in better froth transportation. The R_f values for the OK50 and OK38 cells (shown by asterisks in Table 4), under normal operating conditions, were around 60 per cent and 53 per cent, respectively. These values are lower than expected in a typical rougher circuit which suggests that significant improvements in R_f are possible in the existing cells.

It is to be noted that froth recovery value of a cell is an indication of efficiency of removal of particles in the froth in that cell and comparing R_f values of different cells in different applications could be deceptive. This is because the collection zone kinetics in different cells will vary as a function of impeller design and stator, mixing, cell and feed operating conditions. It is possible to have similar froth recovery values but different overall mineral recovery and therefore concentrate tonnages in two cells treating the same feed.

A better evaluation of froth transportation in cells of different designs is possible by comparing the air specific froth tonnages, T_{as}, in the cells. T_{as}, in essence, normalises the cell concentrate flow rate with respect to different cell operating variables that influence froth transportation behaviour in large cells. T_{as} can be mathematically defined as shown below:

$$T_{as} = \frac{Q_f}{LAT_f} \qquad (4)$$

where Q_f = the concentrate flowrate (m³/h), L is the lip length (m), A is the effective froth area (m²), T_f is the froth residence time (s)

$$T_f = \frac{D}{J_g} \qquad (5)$$

where D = froth depth (cm) and J_g = superficial gas velocity (cm/s)

Air specific froth tonnage, T_{as}, is plotted against zinc or lead recovery and selectivity in Figures 1 and 2, respectively, for the cells surveyed.

The data show strong relationships between air specific froth tonnage and recovery and selectivity which are independent of cell type. This suggests that T_{as} is an important criterion to characterise froth transportation behaviour.

Identifying cell operating conditions for optimum metallurgical performance

Air specific froth tonnage, T_{as}, though a very useful concept, is difficult to implement for optimising cell performance in plants. This is because of the difficulty in measuring cell concentrate flowrates. In addition, T_{as} is not an independent entity and is dependent on cell operation as can be seen in Figure 3. A strong relationship can be observed between air specific froth tonnage and froth residence time, T_f, for the lead and zinc rougher circuits.

TABLE 4

Calculated froth recovery values for zinc rougher cells 2 and 3.

Zinc rougher cell 2 (OK50)					Zinc rougher cell 3(OK38)				
Froth depth	J_g	k	k_c (fitted)	R_f	Froth depth	J_g	k	k_c (fitted)	R_f
35	0.44	0.055	0.069	0.80	47.5	0.7	0.023	0.066	0.34
20	0.44	0.061	0.069	0.89	37.5	0.7	0.034	0.066	0.52
35	0.75	0.068	0.173	0.39	25.0	0.7	0.043	0.066	0.65
20	0.75	0.113	0.173	0.65	47.5	0.9	0.028	0.085	0.32
35	1	0.110	0.300	0.37	37.5	0.9	0.045	0.085	0.53
20	1	0.227	0.300	0.76	25.0	0.9	0.054	0.085	0.64
35	1.36	0.152	0.255	**0.60***	47.5	1.2	0.063	0.104	0.60
20	1.36	0.196	0.255	0.77	37.5	1.2	0.073	0.104	0.70
					25.0	1.2	0.082	0.104	0.79
					47.5	1.4	0.067	0.127	**0.53***
					37.5	1.4	0.083	0.127	0.66
					25.0	1.4	0.095	0.1266	0.75

J_g = superficial gas velocity (cm/s), k = flotation rate constant (1/min), k_c = collection zone rate constant (1/min), R_f = froth recovery
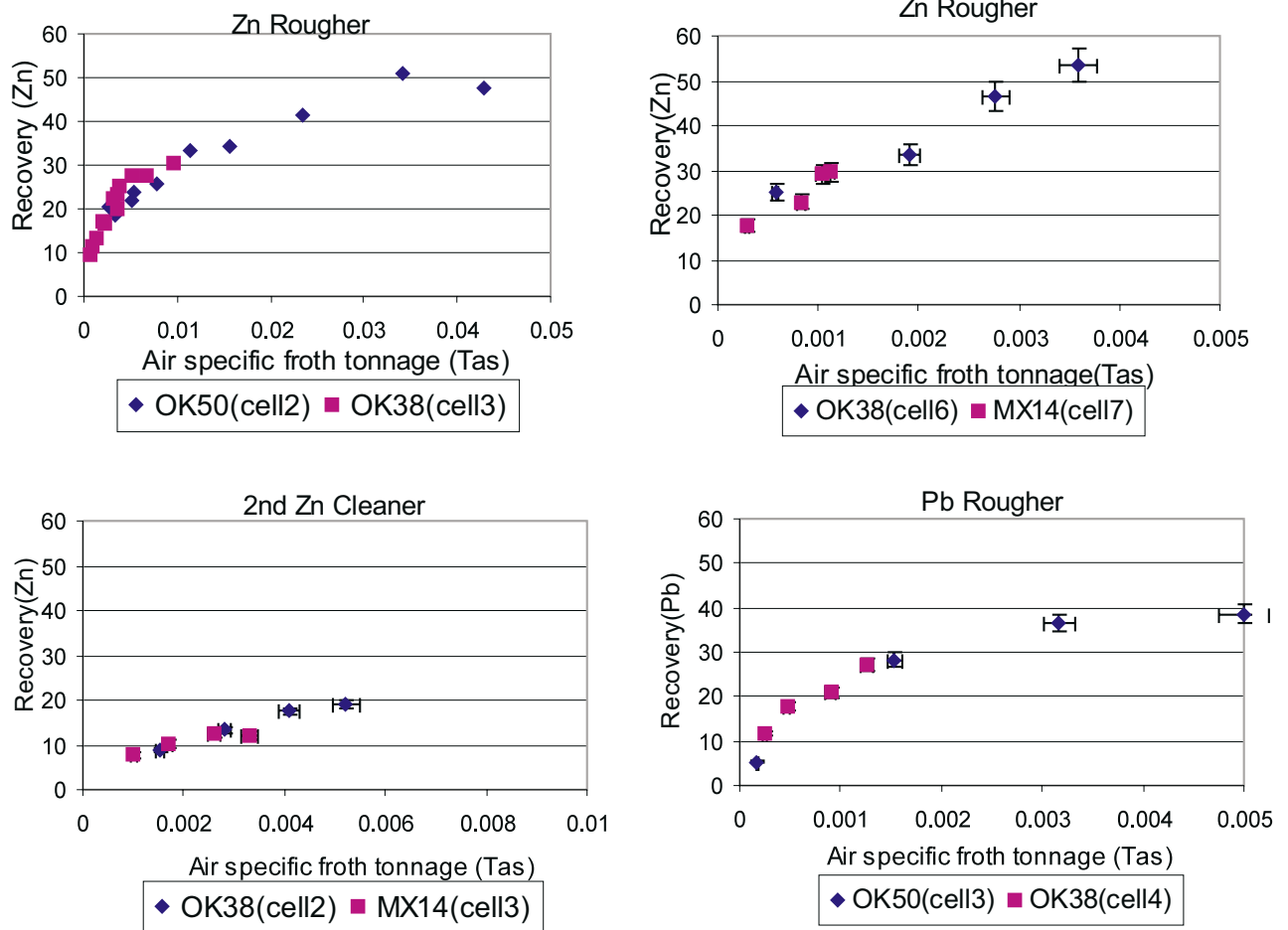
FIG 1 - The effect of air specific froth tonnage on recovery.

Froth residence time, T_f, on the other hand, is a variable which is dependent 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 shows the plot of recovery against froth residence time for the test cells in different circuits. All the plots show a decrease in recovery with an increase in froth residence time. This increase in froth residence time is associated with a deeper froth depth or lower air rate or both. Both these changes are known to decrease recovery. The operating values of froth residence time in cleaners are normally greater (50 - 120 s) than in roughers (20 - 60 s). Cleaner cells are normally operated at deeper froth depths and lower air rates compared to the rougher cells.

Figure 4 shows that the recovery values for the OK50 cells are greater than for the OK38 cells which are in turn greater than the MX14 cells for both lead and the zinc roughers. In the zinc cleaners the recovery values are similar for both OK38 and MX14 cells at froth residence times greater than about 70 seconds. This suggests that the recovery of froth dominates the overall flotation kinetics for both cells in the zinc cleaner circuit. At froth residence time values lower than 70 seconds, the OK38 cell outperforms the MX14 cell, which can be attributed to the poor dispersion of MX14 cells at high air rates.

The effect of froth residence time on the recovery of pyrite follows a trend similar to the recovery of zinc. Figure 5 shows the plot of selectivity as a function of froth residence time for the different cells. An increase in froth residence time resulted in a gradual increase in selectivity up to a point above which it appears to remain constant. This increase in selectivity with increase in froth residence time is likely due to the increase in drainage of entrained pyrite in the froth at either deeper froth depth or lower air rate or both. This drainage is expected to increase to a maximum until a certain froth depth when the froth begins to collapse.

The selectivity values for the zinc rougher cells are significantly greater than for the zinc cleaners. The selectivity values range from about 3 - 15 for the rougher cells and 2 - 3 for the cleaner cells. The low values of selectivity in the cleaners may be attributed to the slow flotation kinetics of fines generated after regrinding. This is an important observation which deserves further investigation to quantify the effects of regrinding of the rougher concentrate on liberation and flotation characteristics.

The froth residence times that result in the best selectivity, recovery and optimum metallurgy for different cells are given in Table 5, which also shows the froth residence time values for the present cell operation. It is to be emphasised that the optimum metallurgy shown in Table 5 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.
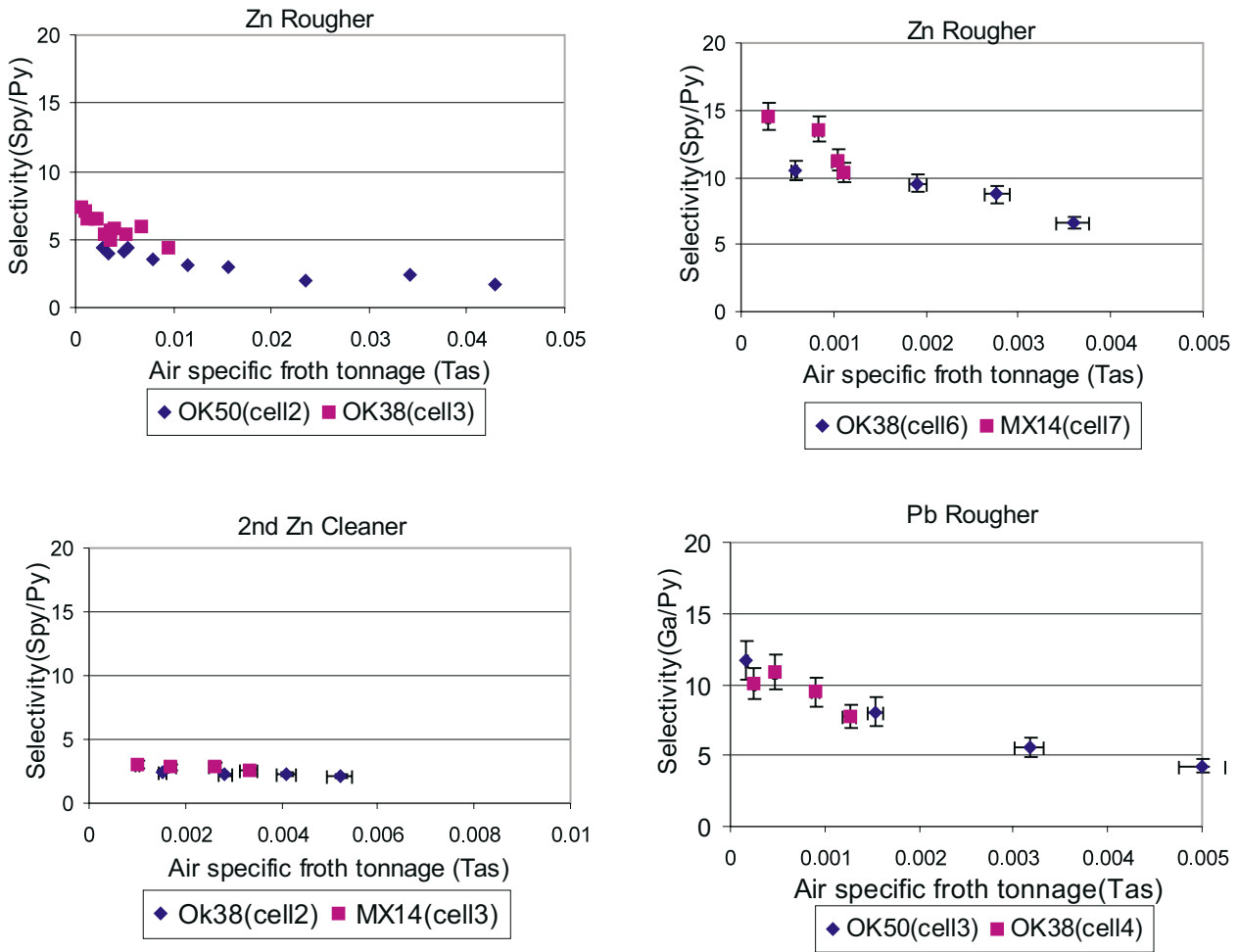
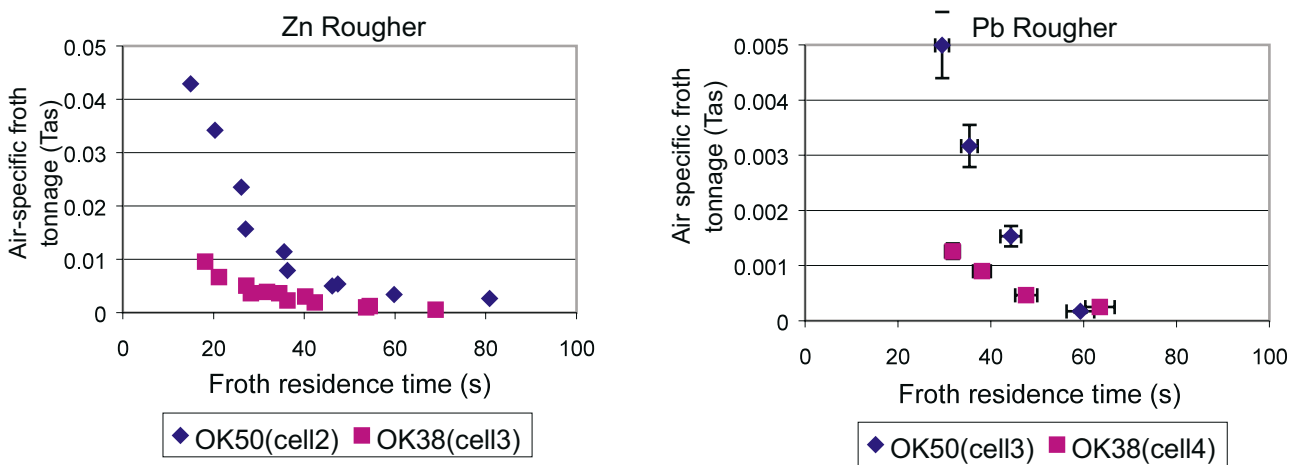FIG 2 - The effect of air specific froth tonnage on selectivity.

FIG 3 - The relationship between air specific froth tonnage and froth residence time.
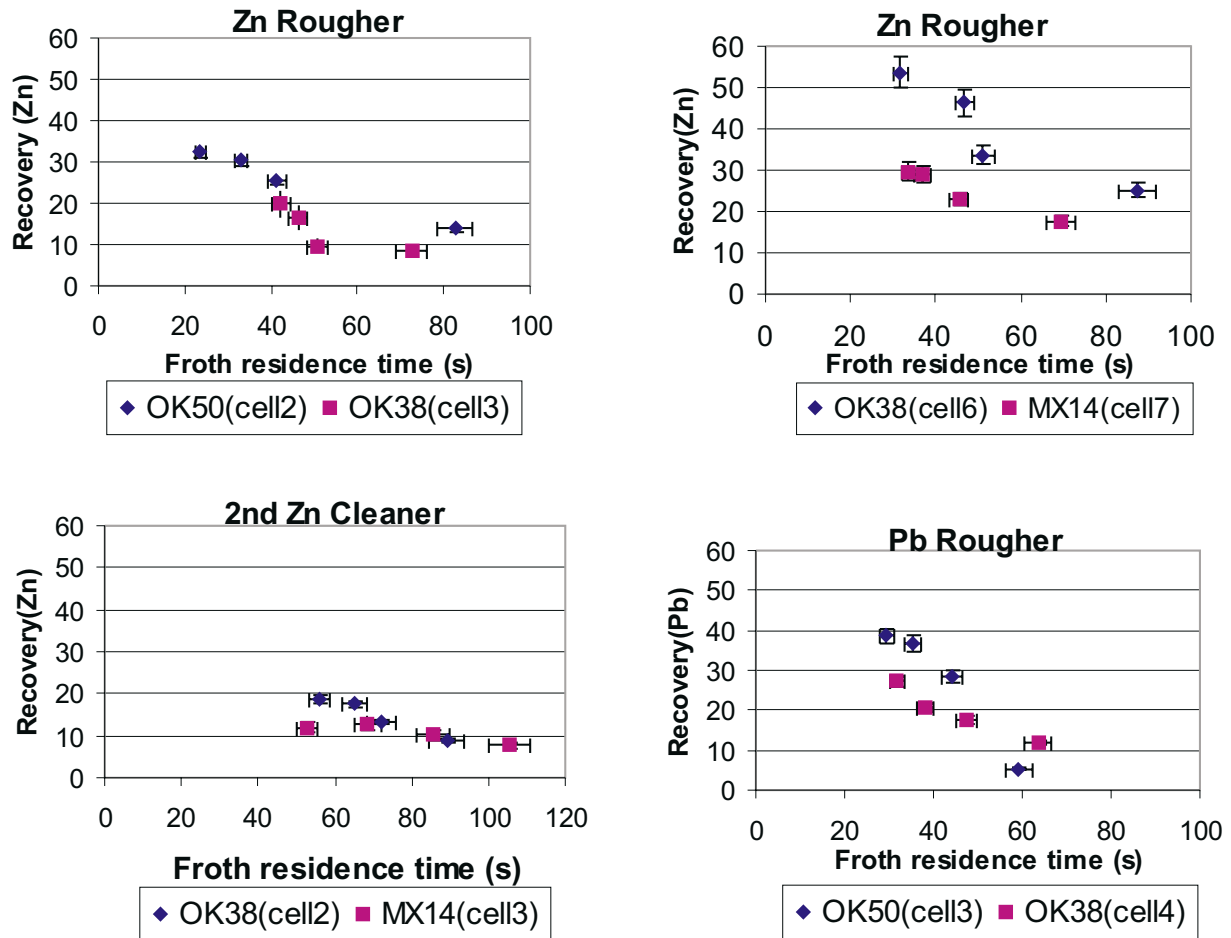
FIG 4 - The effect of froth residence time on recovery.

TABLE 5

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

Cell type	Circuit	T_f (s)	T_f (s)	T_f (s)	T_f (s)
		Presently Operating	Best Selectivity	Best Recovery	Optimum Metallurgy
Cell 2 OK50	Zn Rough	29	60	30	50
Cell 3 OK38	Zn Rough	31	60	40	45
Cell 6 OK38	Zn Rough	31	60	30	45
Cell 7 MX14	Zn Rough	55	60	40	45
Cell2 OK38	2nd Zn Clean	45	75	60	65
Cell3 MX14	2nd Zn Clean	54	75	70	70
Cell3 OK50	Pb Rough	42	50	35	45
Cell4 OK38	Pb Rough	53	60	30	40

The best recovery values were obtained for froth residence times ranging from 30 to 40 seconds for the rougher and 60 to 70 seconds for the cleaner cells. The best selectivity values were obtained for the froth residence times around 60 and 70 seconds for the rougher and cleaner cells respectively. This is not unexpected as these 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 65 to 70 seconds for cleaning. The present cell froth residence times are much lower than required for optimum performance for both zinc roughers and cleaners but are close for lead roughers. An exception is cell 7 (MX14), which is operating at a greater froth residence time than required.
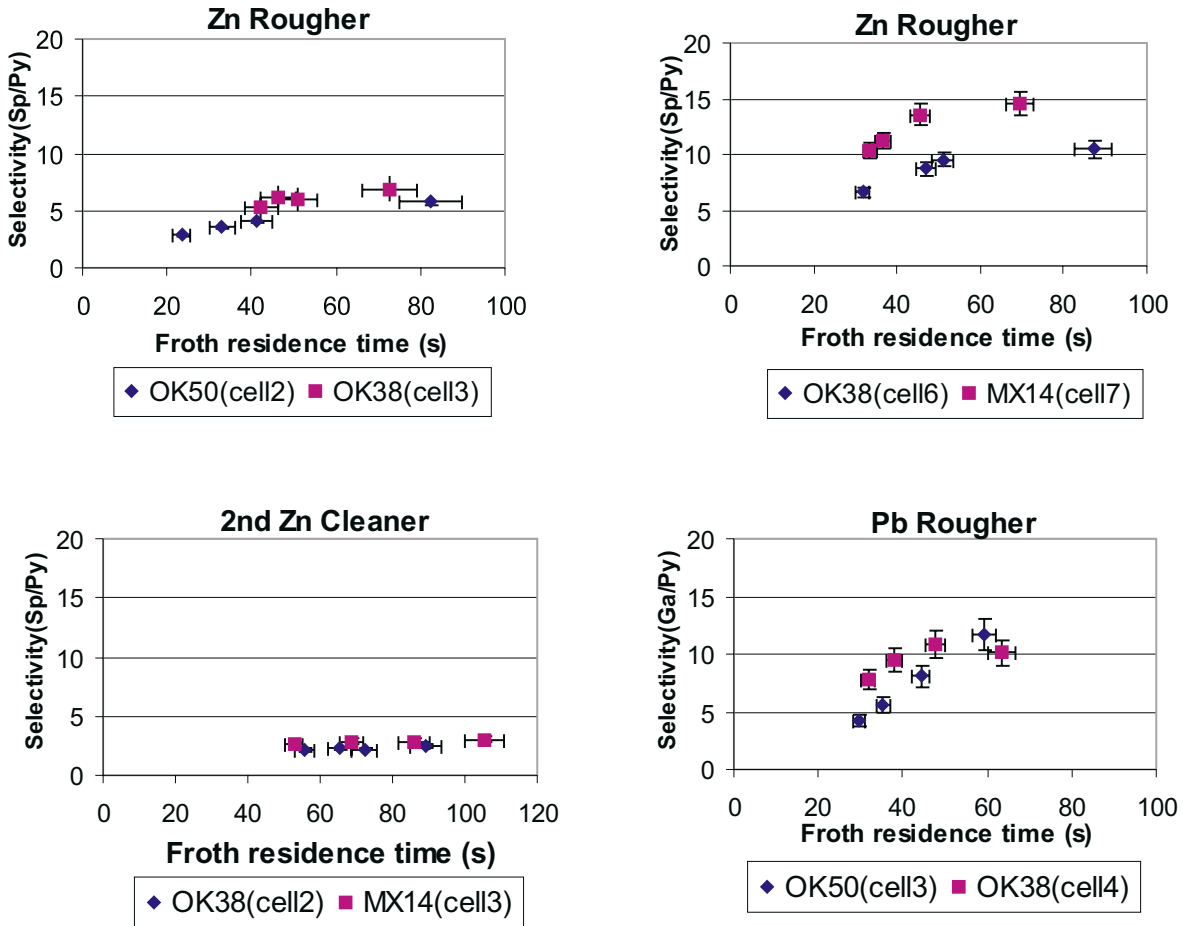
FIG 5 - The effect of froth residence time on selectivity.

Any strategy to adjust froth residence times must take into account factors such as: ability to disperse, ability to transport froth, limits that can be achieved in a cell, power consumption, ease of control, solids suspension and mixing characteristics, maintenance, cost effectiveness, effect on downstream cells and circuit. The author would like to emphasise that the above strategies will vary with cell type and duty, launder design and circuit configuration.

CONCLUSIONS

An extensive study on hydrodynamics and gas dispersion properties was carried out in Outokumpu OK50, OK38 and Maxwell MX14 cells at the Red Dog concentrator in Alaska. The aim of this investigation was to evaluate the operating and metallurgical characteristics of flotation cells to optimise the performance of the Red Dog flotation circuit.

Evaluation of the following cell hydrodynamics and gas dispersion properties provided valuable information on the operating characteristics of the large cells at Red Dog:

- superficial gas velocity,
- bubble surface area flux,
- gas hold-up,
- power consumption,
- residence time distribution, and
- froth flow behaviour.

This study provided insights into the operating limits of different cell types, which are useful in comparing and selecting cells for different duties. The froth transportation behaviour in different cells was characterised by estimating their froth recovery, R_f, values. A better representation of the froth transportation behaviour in cells of different sizes, designs and launder configuration was possible when the concentrate flowrate was normalised with respect to cell lip length, effective froth area and froth residence time, which is termed as 'air specific froth tonnage, T_{as}'. Studies at the Red Dog concentrator showed that T_{as} has a strong correlation with both recovery and selectivity and this relationship is independent of cell type and launder configuration. T_{as}, though effective in characterising froth flow behaviour, is not useful in optimising cell performance.

Froth residence time provides a convenient way of selecting the cell operating conditions in a flotation circuit. It is dependent on froth depth and superficial gas velocity which are easy to vary and measure in plants. Both mineral recovery and selectivity can be related to froth residence times, which can be used to select the best cell conditions for optimum circuit performance. The Red Dog tests showed that optimum metallurgical performance was obtained for cell froth residence times ranging from 40 to 50 seconds for the zinc and lead rougher circuit and 65 to 70 seconds for the zinc cleaner circuit. A comparison of these values with the normal plant cell operating values suggests that the latter are much lower than required for optimum performance for both zinc rougher and cleaner circuits.

ACKNOWLEDGEMENTS

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