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INDUSTRIAL FLOTATION PROCESS MODELLING : RTD MEASUREMENT BY RADIOACTIVE TRACER TECHNIQUE

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Abstract: The residence time of liquid and solid was determined in flotation banks of mechanical cells and flotation columns at Salvador Division, Codelco-Chile. The estimation was developed from residence time distribution (RTD) measurements using radioactive tracers. Results allowed the estimation of effective pulp volumes in flotation banks and columns. It was found that industrial flotation banks are properly modeled by a number of perfect mixers in series equivalent to the number of cells in the bank. However, large size flotation columns are closer to a perfect mixer. Measurement of RTD also allowed the identification of operating control problems such as solid sedimentation and excessive tailing valves manipulation. *Copyright © 2002 IFAC*

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1. INTRODUCTION

Mineral species recovery in continuous industrial flotation equipments can be described by the general equation,

$$R = R_\infty \int_0^\infty \int_0^\infty (1 - e^{-kt}) F(k)E(t)\, dk\, dt \qquad (1)$$

where R_∞ represents the maximum flotation recovery at infinite time. The term $(1 - e^{-kt})$ represents the mineral recovery of a first order process with invariant kinetic constant k, as a time function. F(k) is the kinetic constant distribution function for mineral species with different flotation rates, and E(t) is the residence time distribution function for continuous processes with different mixing characteristics. The objective to determine the residence time distribution (RTD) was to evaluate the effective residence time of the liquid and solid phases in industrial flotation equipments. Thus, the mineral recovery or the flotation kinetics of mineral species can be predicted from Equation (1).

1.1 Impulse Response Method

In a multiphase system with segregation, estimation of the mean residence time of each phase is related to the effective volume occupied by each phase. In plant practice, the volume occupied by each phase (liquid, solid or gas) is unknown and varies with operational conditions. Alternatively, in order to evaluate the effective residence time of the liquid and solid in flotation equipments, the impulse response method was used in this work. This method is a dynamic identification procedure which consists of introducing a small amount of tracer (close to an impulse) into the system operating at steady state, and to register the transient response (tracer concentration). The response corresponds to the transfer function of the system, and it is of interest for dynamic identification (order, noise) and studying control systems, as well as for dynamic modelling (short-circuiting, recirculation) (Yianatos and Bergh, 1992).

1.2 Application of the Radioactive Tracer Technique

Measurement of residence time distribution in different industrial flotation cell arrays was developed by the radioactive tracer technique. This technique allows the non-invasive tracer detection, and also is adaptable to different kind of equipment (Goodall and O'Connor, 1989; Blet et al., 1999). However, the application of the radioactive tracer technique for industrial flotation characterization is rather scarce.

The procedure consists of selecting a liquid or solid tracer that allows on-line RTD data acquisition. The way the tracer is injected into the feed is critical in order to generate a pulse signal (closer to impulse). Therefore, a pneumatic system of high reliability was used for this objective. Sensors for tracer activity detection were installed on the discharge pipes of each equipment to obtain the process time response, thus allowing the simultaneous data acquisition of up to 12 control points, with a minimum period of 50 milliseconds. Br-82 was used as liquid tracer element, while different types of minerals were used to trace the solid :

a) non floatable irradiated solid
b) non floatable irradiated solid per size class

2. EXPERIMENTAL WORK

The residence time distribution of the following flotation circuits of the bulk Cu-Mo concentrator of Division Salvador was determined.

a) Rougher flotation circuit, consisting of five parallel banks of Wemco cells
b) Scavenger flotation circuit, consisting of two parallel banks of Dorr-Oliver cells
c) Cu/Mo cleaning circuit, consisting of two parallel flotation columns

2.1 Liquid and Solid Residence Time Distribution in Rougher Wemco Cells

Residence time distribution of liquid and solid was measured simultaneously for the five rougher banks of Wemco cells. Each bank consists of 9 cells, 42.5 m^3 each, in arrangement 3-3-3.

Tests were developed with a solution of Br-82 as liquid tracer and activated tailings as solid tracer. Table 1 shows a summary of the mean residence time of liquid and solid estimated from the measured RTD for the five rougher banks.

Table 1. Liquid and solid residence time

Rougher Bank	Liquid, Br-82 Mean time min	Activated Solid Mean time min
1	35.9	34.7
2	40.3	38.3
3	40.8	39.2
4 *	54.7	49.9
5	36.7	35.0

* abnormal operation, large concentrate overflow

In general, it was observed that the solid mean residence time was lower than the liquid with a difference of 1 to 2 minutes, which approximately corresponds to a 5% lower mean time. The difference can be explained due to gravity segregation affecting the mineral, particularly in coarse classes (>100-200 microns). Also, solid particles have less probability of internal circulation than the liquid because of the lower entrainment into the froth.

In example, Figure 1 shows the liquid residence time distribution in rougher bank 5, where the mean residence time was 36.7 minutes.

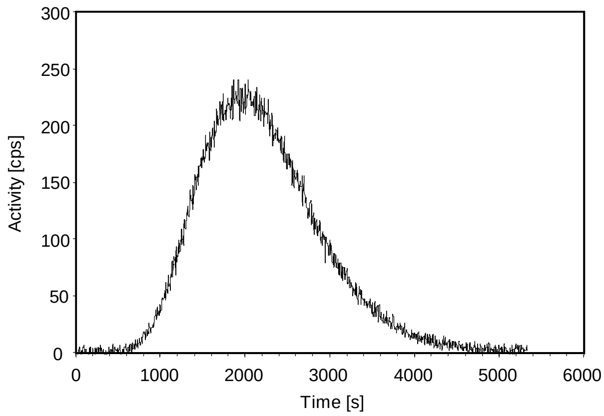

Fig. 1. Liquid RTD in rougher bank (9 cells)

Figure 2 shows the residence time distribution of the irradiated solid tailing in rougher bank 5, where the mean residence time was 35.0 minutes.

2.2 RTD Modelling of Industrial Flotation Cells

It was found that the liquid RTD can be properly represented by a model of N perfect mixers in series with pure delay, according to the normalized distribution function,

$$E(t) = \frac{(t)^{N-1} \exp[-t/(\tau/N)]}{(\tau/N)^N \; \Gamma(N)} \qquad (2)$$

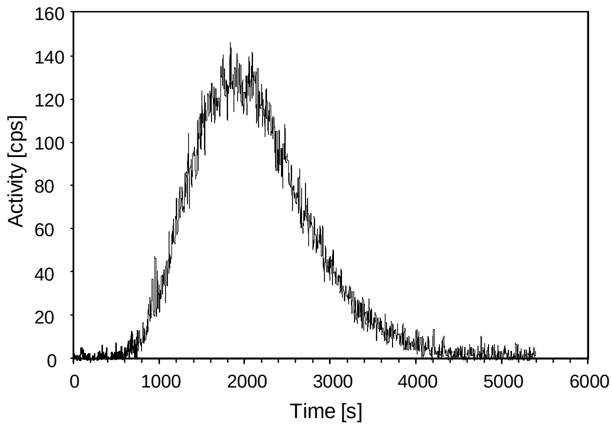

Fig. 2. Solid RTD in rougher bank (9 cells)

Equation (2) allows the fitting of non integer values of N. In the case of rougher banks the best fit was obtained for an N value approximately equal to 9, which corresponds to the number of cells in the bank. In example, Figure 3 shows the model fitting for the data of rougher bank 1.

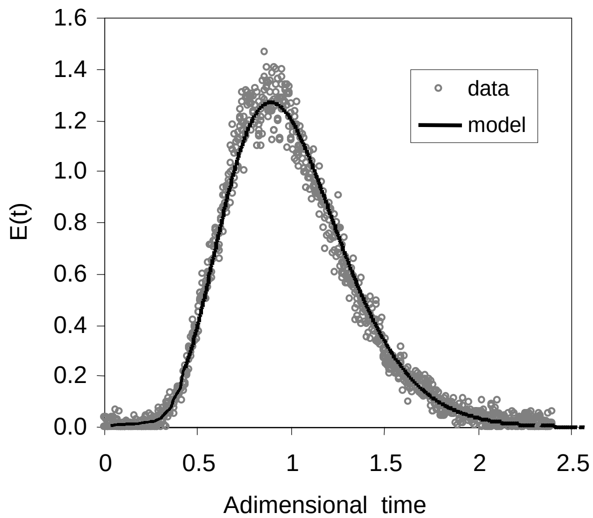

Fig. 3. Model fitting for N perfect mixers in series

2.3 Mean Residence Time and Effective Volume Estimation in Rougher Flotation

In order to estimate the mean residence time of the pulp in the rougher circuit a flotation process simulator considering the cell design characteristics, the flotation kinetic and the species mass balance in each cell of the bank was used. Then, knowing the flowrate and feed characteristics of each bank the flotation kinetic was adjusted according to the actual mass balance, i.e. grades and solid percentage in concentrate and tailings.

The mean residence time of each cell was estimated from the volumetric pulp flowrate entering the cell and the effective cell volume. The maximum pulp volume V_p in each cell was estimated by,

$$V_p = V_c(1 - \varepsilon_{gp}) + V_e(1 - \varepsilon_{ge}) \qquad (3)$$

considering the froth depth equal to 0.1 m, cell volume $V_c = 40.9$ m^3, froth volume $V_e = 1.6$ m^3, pulp air holdup $\varepsilon_{ge} = 17\%$ and froth air holdup $\varepsilon_{ge} = 85\%$ (Yianatos et al, 2000). In this way V_p resulted equal to 34.2 m^3, and the percentage of effective cell volume V_{ef} can be estimated by,

$$V_{ef} = 100\, V_p / V_t \qquad (4)$$

which corresponds to 80.4% of the nominal cell volume V_t. The actual value can be lower depending on the existence of dead zones or solid settling. The total pulp residence time in each bank was calculated by adding the flotation time of 9 cells in the bank plus the time related to the transport delay.

Also, the solid mass flowrate was measured by weight-meters located under the grinding feed belts. During the liquid tracer tests the total mass flowrate was 1450 tph which corresponded to 2895 m^3/h volumetric flowrate.

Table 2 shows the mass and volumetric flowrates adjusted according to the mean residence time directly measured in each rougher bank. In case of rougher bank 4 the operation was not normal in some cells because of solid sedimentation problems which constrained the tailings discharge thus generating a larger concentrate overflow. This condition was confirmed while developing kinetic tests and RTD per size classes in the same flotation circuit.

Table 2. Flowrate distribution in rougher banks

Bank N°	Mass flowrate tph	Volumetric flowrate m^3/h	Total Time min	RTD time min
1	315	629	36.1	35.9
2	284	567	40.6	40.3
3	281	561	40.9	40.8
4 *	286	571	54.3	54.7
5	309	617	36.8	36.7

* abnormal operation, larger concentrate overflow

According to this results it can be observed that the flowrate distribution among the five rougher banks varies from 19 to 21%, which is not a significant difference from the point of view of the circuit balance and its control. It was found that in case there is no dead zones or significant solid settling, the total mass flowrate should be 1475 tph, which is approximately 2% larger than the tonnage directly measured by weight-meters. Then, considering that the error in the plant feed tonnage measurement was about ± 2%, it was concluded that the effective cell volume can decrease from 80.4% to 78.8% of the nominal cell volume, mainly because of solid sedimentation in the flotation cell.

2.4 Particle Size Effect on Solid Residence Time

The effect of particle size on the solid residence time was evaluated. Tests were developed on banks 3 and 4 of the rougher flotation circuit.

Activated tailing was used as solid tracer in the following size classes : Coarse (+150 μm), Medium (-150+45 μm) and Fine (-45 μm).

Table 3 shows a summary of the mean residence time per size class estimated from solid RTD in rougher bank 3.

Table 3. Solid residence time per size class

Size class	fine	medium	coarse
Time, min	41.5	39.5	38.2

The effect of particle size showed that mean residence time of fine mineral in rougher bank 3 was similar to the liquid (40.9 min.), the mean residence time of medium size mineral was similar to the global solid (39.2 min.) and the coarse mineral showed a mean residence time lower than medium. This result confirms that increasing particle size decreases the mean residence time of minerals. Solid segregation in flotation cells is partially due to the larger gravitational effect on coarse particles and also because of lower internal circulation of solid by entrainment into the froth. Difference between mean residence time of liquid and overall solid was approximately 5%, which showed a reasonable homogeneity in the pulp suspension. In case of flotation columns, however, a mineral with an average size of 100 microns has a residence time equal to a half of the liquid residence time (Dobby and Finch, 1985; Yianatos and Bergh, 1992).

2.5 Mean Residence Time and Effective Volume Estimation in Scavenger Flotation

The scavenger circuit consists of two parallel banks of 8 Dorr-Oliver cells, 42.5m^3, in arrangement 2x2x2x2, and the total cell volume was V_T equal to 680 m^3. The volume for transportation between cells was estimated in 34 m^3.

Table 4 shows the results of estimating mean residence time of liquid and solid in both flotation banks from RTD measurements. It can be observed that results are consistent and they show a minor difference between liquid and solid. The smaller difference is due to the mineral size because in this stage the mineral has been reground and the size is finer with an 80% lower than 45 microns, also because of the larger circulation of fine particles into the froth by entrainment.

Table 4. Solid and Liquid Residence Time

Scavenger Bank	Liquid Br-82 Time, min	Solid Tailing Time, min
1	21.6	23.1
2	21.2	21.9

The volumetric feed flowrate was directly estimated from pulp velocity measurements by using radioactive liquid tracer and was about 1200 m^3/h. For this flowrate the nominal pulp residence time in the total cells volume would be 39.1 minutes, considering a normal increase of about 15% in the mean residence time of the pulp. This increase in residence time is related to the fraction of pulp which is gradually recovered into the concentrate thus decreasing the remaining pulp flowrate. Comparison of the nominal residence time (39.1 min) with the direct measurement of mean residence time (21.4 min) less the transport delay time in the cells transfer boxes (1.6 min), showed that the effective flotation volume was about 50.6 % of the total cells volume. This result is similar to the estimation of the V_P volume occupied by pulp in all cells calculated by equation (3). In this case, considering the froth depth measurement equal to 1.1 m, V_c = 392 m^3, V_e = 288 m^3, ε_{gp} = 18% y ε_{ge} = 90%, the resulting V_p was 350 m^3, thus the effective cell volume estimated by equation (4) corresponds to 51.5% of the total cells volume.

The froth depth was directly measured and it was larger than one meter in all cells. This non-normal condition was principally due to limitations in the air flowrate control and the pulp level control system, both regulated by manual modification of the valves set point.

2.6 Liquid and Solid Residence Time in Flotation Columns

The mean residence time of flotation columns in the cleaning circuit was determined. The circuit consists of two rectangular columns, 2x6x13m, operating in parallel. The total column volume V_T was 312 m^3. Figure 4 shows the liquid residence time distribution in column 1, where the mean residence time was equal to 10.7 minutes.

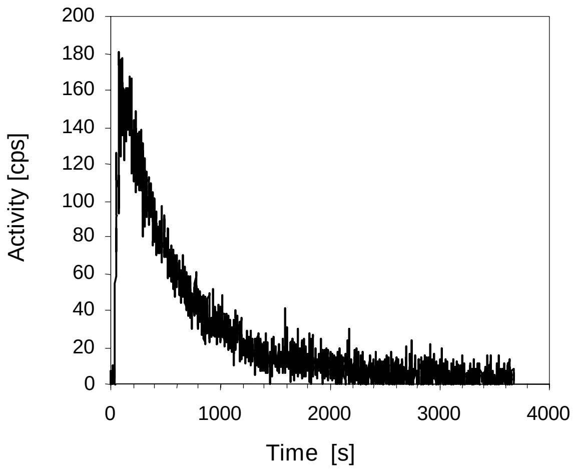

Fig. 4. Liquid RTD in flotation column, 2x6x13m

Figure 5 shows the solid residence time distribution in column 1, where the mean residence time was 8.9 minutes.

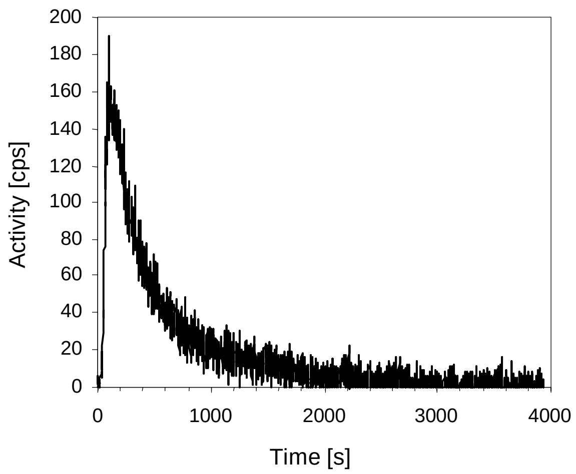

Fig. 5. Solid RTD in flotation column, 2x6x13m

Table 5 shows the estimation of the mean residence time of liquid and solid for both columns from RTD measurements. The residence time of solid tailings, with particle size d_{80} = 20-30 microns, was equal to 83% of the liquid residence time. This result was in good agreement with previous studies relating the liquid and solid residence time as a function of particle size (Dobby and Finch, 1985; Yianatos and Bergh, 1992).

The common feed flowrate to both columns was measured on-line with a magnetic flowmeter. This flowmeter was calibrated by measuring the pulp velocity using liquid radioactive tracer. Results showed that the magnetic flowmeter reading was adequate with an error lower than 1%.

Table 5. Liquid and Solid Mean Time in Columns

Type of injection	Column	Liquid, Br-82 Time, min	Tailings Time, min
internal	1	10.7	8.9
common	1	10.6	-
	2	12.6	-

The volumetric flowrate measured during experiments under common tracer injection to both columns was 1087 m^3/h. Also the wash water flowrate added over the top of both columns was 148 m^3/h. Then, considering the columns total volume was 312 m^3, a nominal average time of 15.1 min was obtained. Consequently, according to the actual mean residence time of the pulp in both columns (11.6 min) the effective volume occupied by pulp was 77.3% of the total volume.

The estimate of the volume V_P occupied by pulp in both columns was 241 m^3, according to equation (3), considering the measurement of the froth depth equal to 1m, V_c= 288m^3, V_E = 24 m^3, ε_{gP}= 18 % y ε_{gE}= 80% (Yianatos et al, 1998). This result corresponds to an effective volume of 77.2% of the total volume of both columns, according to equation (4), which is in good agreement with estimation based on RTD measurement.

2.7 RTD Modelling of Industrial Flotation Columns

The liquid RTD of industrial flotation columns can be described by equation (2), using a model of less than 2 perfect mixers in series. However, a better fit was obtained considering a model consisting of one large perfect mixer (residence time τ_L) and two small perfect mixers in series (residence time τ_S) represented by the following equation,

$$E(t) = \frac{[\,-t/\tau_S - \alpha\,]\exp[\,-t/\tau_S\,] + \alpha\exp[\,-t/\tau_L\,]}{(\tau_L - \tau_S)} \qquad (5)$$

where,

$$\alpha = \tau_L / (\tau_L - \tau_S) \qquad (6)$$

The column local mixing conditions generated near the feed input and the bubble generation zones can be related with the two small mixers. On the other hand, the collection zone extended from the feed entrance to the bubble generation level can be related to the single and large mixer. This result confirm that despite the presence of baffles, the mixing condition in an industrial column was closer to a large perfect mixer, due to air bubble circulation, mainly for high air flowrates.

2.8 Operating Control Problems

Measurement of residence time distribution in industrial equipment and estimation of the pulp flowrates in pipes from measurements of pulp velocities using radioactive tracers, allowed the observation of different operating problems, in example: pipes containing air pockets (air accumulation), improper flowmeter readings because of lack of calibration, solid sedimentation problems and pulp level control problems which cause excessive manipulation of valves used to regulate the discharge flowrate in flotation cells and columns.

3. CONCLUSIONS

Measurement of residence time distribution allowed the estimation of pulp effective volumes of 50-80% of the total cell volume, in rougher and scavenger flotation cell banks, and effective volumes of 77% of the total volume in flotation columns.

In flotation cell banks it was found that mean residence time of solid (23%+212 microns) was 5% lower than liquid. Also, it was shown that the mineral residence time decreases by increasing the particle size. In example, coarse mineral (> 150 microns) showed a residence time 8% lower than the finest mineral (< 45 microns). This effect, however, is less significant than the one observed in flotation columns. Thus, the common use of liquid RTD for flotation modelling purposes must be considered with caution.

It was found that the RTD of industrial mechanical cells banks are well described by a number of perfect mixers in series equivalent to the actual number of cells in the bank.

The RTD of large size industrial flotation columns was closer to a perfect mixer. Thus, the column mixing conditions showed a significant pulp short-circuiting into tailings which strongly decreases the mineral recovery.

RTD measurement also allowed the observation and analysis of common operating problems such as solid sedimentation in flotation equipments and excessive manipulation of discharge flowrate valves used for pulp level control.

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