CHARACTERIZATION OF A LARGE SIZE FLOTATION CELL BANK

Juan Yianatos, F. Henríquez
Depto. Procesos Quím., Biotecnológicos y Amb.
Universidad Santa María - Chile
juan.yianatos@usm.cl

A. Oroz
Superintendencia de Ingeniería Metalúrgica
Minera Escondida – Chile

ABSTRACT

Design and operating conditions of a large size mechanical flotation cell bank were evaluated by comparing it with the actual operating conditions in a plant. The objective was to determine scale-up factors, which are still mainly based on empirical rules. Experiments were conducted on the rougher flotation circuit at Minera Escondida Ltd. Co. The circuit consisted of self-aerated mechanical cells of 160 m$^3$, arranged in 6 parallel banks with 9 cells each.

From samples, local mass balances around each cell of the bank and adjusted overall mass balances, the flotation kinetics was evaluated for the rougher circuit. This information was used to fit different kinetic flotation models, and was found that the rectangular distribution function was the most appropriate to describe the distributed rate constant for industrial operation. Thus, a rougher flotation simulator was developed to describe the actual operation. Feed samples were taken in parallel from the rougher circuit and batch kinetics tests were performed. The metallurgical behavior was then modeled at a laboratory scale in order to determine the scale-up factors between laboratory and plant. The scale factors observed for large sized cells, 160 m$^3$, were found reasonably similar to those previously determined for self-aerated mechanical cells, also arranged in banks of 9 cells, but of lower size (42.5 m$^3$).

In general, the rougher flotation operation was found to reach the predicted metallurgical target, and that the optimal separability criteria was also respected.
INTRODUCTION

In the last decade, flotation equipments have shown a significant increase in size. Since the 80’s mechanical cells size has increased 10 times, reaching levels over 200 m$^3$ per cell. At the present moment, most of the new concentrators in Chile have been fully equipped with 140-160 m$^3$ cells. On the other hand, pneumatic flotation columns of more than 260 m$^3$ are now in operation. Thus, the dramatic increase in size of flotation equipments poses new challenges related to equipment design, in terms of mixing and pulp circulation, as well as optimizing bubble generation and dispersion, and the froth mineral transport.

Laguna Seca concentrator.

The Laguna Seca concentrator belongs to the mining company Minera Escondida Ltd. and is located in Los Andes pre-cordillera, Chile, 3200 m above sea level and 160 km from Antofagasta city. The plant’s capacity is 110,000 tpd of ore with a feed grade of 1.27% copper and produces a final concentrate of 40% copper. The flotation circuit, Figure 1, consists of a rougher stage, with 6 parallel banks of 9 Wemco cells, each 160 m$^3$. The rougher concentrate together with the scavenger concentrate goes to a regrinding circuit consisting of three parallel vertical mills. The rougher and scavenger tailings are the plant’s final tail.

The first cleaner feed combines the regrinding mill discharge together with the second cleaner tail. The first cleaner tailing goes to the scavenger circuit. Both circuits consist of 5 parallel banks with 4 Wemco cells, each 160 m$^3$. The first cleaner concentrate goes to the second cleaner circuit consisting of 8 parallel flotation columns, each 4.5 m in diameter and 13.5 m in height.

Rougher flotation cells.

Mechanical cells are self-aerated circular in shape and are 6.8 m in diameter 4.4 m in height giving a total volume of 160 m$^3$. The mixer is a vertical cylindrical rotor located near the pulp/froth interface. The rougher circuit consists of 6 parallel banks each of 9 cells arranged as $1x1x1x1x2x2x1$ for level control purposes. The first 3 cells of each bank were provided with 10 radial internal launders, while the remaining cells with 8 radial launders.
Control system.

The targets of the rougher stage are to maximize the copper recovery, avoid the overloading of the regrinding and cleaning circuits, and to keep a minimum concentrate grade. Rougher concentrate and tailings grade are monitored while pulp level and frother dosage are manipulated variables. On the other hand air aspiration valves are kept fully open.

EXPERIMENTAL

In order to evaluate the rougher flotation bank performance sampling from each cell was developed.

Operating conditions

Table I shows the main operating conditions during the rougher flotation sampling.

<table>
<thead>
<tr>
<th>Table I. Rougher operating conditions</th>
</tr>
</thead>
<tbody>
<tr>
<td>Date : 18/12/03</td>
</tr>
<tr>
<td>Rougher feed : Total copper, % : 1,27 ± 0,20</td>
</tr>
<tr>
<td>Rougher feed : Total iron, % : 1,89 ± 0,16</td>
</tr>
<tr>
<td>Ratio : Cu/Fe : 0,67</td>
</tr>
<tr>
<td>Rougher feed : Tonnage (tph) : 4,516,0 ± 6,9</td>
</tr>
<tr>
<td>Rougher feed : solids % : 31,4</td>
</tr>
<tr>
<td>Primary Collector AP-3758 (g/t) : 21,0</td>
</tr>
<tr>
<td>Secondary Collector SF-114 (g/t) : 9,0</td>
</tr>
<tr>
<td>Frother (30% pin oil, 70% X-133) (g/t) : 14,0</td>
</tr>
</tbody>
</table>

The overall mass flowrate was assumed to be proportionally distributed in the parallel operating rougher flotation banks.

The plant feed tonnage during the sampling period is shown in Figure 2. Around the set point the operation was stable and the variations were normal.

![Figure 2. Flotation feed tonnage](image-url)
Procedure

Batch tests

A rougher feed sample was collected while the rougher kinetics study was conducted. The sample was used to develop batch flotation tests with the same feed characteristics in order to estimate the laboratory based kinetic parameters like maximum recovery ($R_{\text{MAX}}$) and the flotation rate constant ($k_{\text{MAX}}$).

Concentrate sampling

The concentrate samples were taken directly from the cell lip using a modified standard sampler of 20 cm in length.

The standard sampler was modified, adding a 7.5 cm extension, in order to hang it from the cell lip, making it possible to collect the full amount of concentrate in that perimeter. Here, the sampled concentrate grade was assumed to be representative for each cell. Samples were taken with a 5 minutes delay to account for the pulp residence time in each cell. Figure 4 shows the location of the concentrate sampler and internal launders in the rougher cells.

![Top view of a flotation cell](image)

**Figure 4.** Sampler location for concentrate overflow

Tails sampling

The tails sampling was performed in the last cell of the rougher flotation bank.

The concentrate mass flowrate per cell was estimated by measuring the mass overflowing over the sampling perimeter for a time period. The overall rougher concentrate, estimated from the overall mass balance, was then distributed proportionally to the measured mass flowrate in each cell.

All samples were analyzed for copper, iron and non soluble minerals grade, solid percentage and specific gravity.

RESULTS AND DISCUSSION

Batch flotation and plant data adjustment

Table II shows the results of the batch flotation tests of the samples taken from the rougher feed.
Table II. Batch flotation results from rougher feed

<table>
<thead>
<tr>
<th>Time, min</th>
<th>Dry weight, g</th>
<th>Cu, %</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>52.9</td>
<td>18.09</td>
</tr>
<tr>
<td>2</td>
<td>25.6</td>
<td>7.68</td>
</tr>
<tr>
<td>12</td>
<td>107.6</td>
<td>1.28</td>
</tr>
<tr>
<td>Tail</td>
<td>919.2</td>
<td>0.11</td>
</tr>
<tr>
<td>Total</td>
<td>1,105.3</td>
<td>1.26</td>
</tr>
</tbody>
</table>

Using two kinetic models, the above data and data obtained from the rougher kinetic sampling the flotation operation was characterized. Table III shows the tested models, for both batch laboratory and plant operation. These models correspond to the classical first order model which has an invariant rate constant and the model which consider a rectangular rate constant distribution.

From an hydrodynamic point of view the rougher bank performance was assumed can be well described using the tank in series model proposed by Yianatos et al. (1) which also considers rate constants of rectangular distribution, Yianatos et al. (2).

Table III. Flotation models for rougher kinetics

<table>
<thead>
<tr>
<th>Model</th>
<th>Equation</th>
</tr>
</thead>
<tbody>
<tr>
<td>Batch, F(k) : rectangular</td>
<td>[ R = R_\infty \left( 1 - \left( \frac{1-e^{-kt}}{kt} \right) \right) ]</td>
</tr>
<tr>
<td>Batch, k : invariant</td>
<td>[ R = R_\infty (1 - e^{-kt}) ]</td>
</tr>
<tr>
<td>Plant, N perfect mixers in series, F(k) : rectangular</td>
<td>[ R = R_\infty \left[ 1 - \left( \frac{1 - (1+k_{\text{max}} \tau_P)^{1-N}}{(N-1)k_{\text{max}} \tau_P} \right) \right] ]</td>
</tr>
</tbody>
</table>

where,

- \( R \) : mineral recovery at time \( t \), %.
- \( R_\infty \) : maximum recovery at infinite time, %.
- \( k \) : flotation kinetic constant, \( \text{min}^{-1} \).
- \( k_{\text{max}} \) : maximum flotation kinetic constant for a rectangular distribution, \( \text{min}^{-1} \).
- \( N \) : number of perfect mixers in series.
- \( \tau_P \) : cell residence time, min.

A flotation bank simulator to characterize plant operation which allowed the use of the N perfect mixers in series model with rectangular distribution F(k), but accounting for pulp condition and flowrate changes along the bank of cells was developed.

Table IV shows the results obtained from adjusting the corresponding model. The plant design parameters are also shown (3).

Table IV. Kinetic parameters for rougher flotation

<table>
<thead>
<tr>
<th></th>
<th>Batch</th>
<th>Plant</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>k invariant</td>
<td>F(k) rectangular</td>
</tr>
<tr>
<td>( R_{\text{MAX}} )</td>
<td>91,5</td>
<td>95,4</td>
</tr>
<tr>
<td>( k, \text{min}^{-1} )</td>
<td>1,34</td>
<td></td>
</tr>
<tr>
<td>( k_{\text{MAX}}, \text{min}^{-1} )</td>
<td>3,53</td>
<td>1,24</td>
</tr>
</tbody>
</table>

PROCEMIN 2004, Santiago, Chile. www.procemin.cl
Figure 5 shows a good agreement between the experimental data and the flotation model. In this case the mineral performance was described by a rectangular F(k) distribution which accounts for the mineral quality loss from the feed entrance down towards the final tailings. However, other factors differing from mineral characteristics also affect the overall flotation performance. For this case, the addition of a secondary collector in cell 4 and the significant froth depth decrease in the last cell 9, are both disturbances which alter the smooth performance predicted by the constant kinetic parameter model. In fact, the effect of these kinds of disturbances is equivalent to introducing a permanent or local change in the "overall" flotation rate constant. This effect became apparent when the performance of the last 2 bank cells was predicted from the data fitting for the first 7 cells.

![Figure 5](image1.png)

**Figure 5.** Plant recovery data and simulation results.

Figure 6 shows the cell's froth depth varying from 7 to 24 cm along the flotation bank during the sampling period. Cell 1 and 9 showed the lower froth depth, thus contributing to a higher recovery in the first cell and a significant entrainment in the last cell.

![Figure 6](image2.png)

**Figure 6.** Froth depth along the rougher flotation bank

Figure 7 shows the local copper concentrate grade per cell and the cumulative concentrate grade along the rougher bank, compared with the model prediction.
A discontinuity between cells 3 and 4 can be observed in Figure 7. This effect can be partially related to the decrease in the number of internal launders, from 10 to 8, in these cells. There was also a decrease in froth depth for cell 3 and 4, producing an increase in the overall volumetric flowrate, which can be observed in Figure 8, thus increasing the potential for solids entrainment. On the other hand, it was appreciated that the optimal separability criteria was respected because, downwards along the bank, the local concentrate grade decreased reaching the value near the feed grade.
Figure 9 shows the local copper recovery per cell, and the model prediction, using the flotation bank simulator.

![Graph showing local copper recovery per cell](image)

Cells 5 and 7 from Figure 8 show a smaller concentrate overflow, compared with cells 6 and 8, which are respectively located on the same bank. This result can partially be related to a sampling constraint, because due to access problems the sampling points for these two cells were in opposite positions to the other cells. The difference also can not be related to the operation itself because the froth depth and air flowrate were similar. On the other hand, the larger increase in the solid overflow of the last cell can be related to a significant decrease in the froth depth as shown in Figure 6.

**Carrying capacity of the first cell.**

For a bank of cells it is normal for the first cell to have the higher recovery and grade percentage, see Figures 7 and 9. So, an expected limited condition, in terms of mineral transport, could be reach by increasing cell size. This occurs because in self aerated cells, increasing the cell size would decrease the specific air rate, the superficial air rate, the bubble surface flux, and the froth recovery.

In order to evaluate the actual carrying capacity, the mass flowrate at the pulp/froth interface level was estimated. According to Alexander et al. (4), a froth recovery factor of 40% was assumed for this purpose.
Figure 10 shows the size distribution, in terms of the cumulative passing percent, for the first cell concentrate. The $d_{80}$ was around 78 microns.

![Figure 10. Solids size distribution in the first cell concentrate](image)

Figure 11 shows, for an hypothetical minimum superficial air rate of 0.4 cm/s, for different particle and bubble sizes, an estimate of the maximum theoretical carrying capacity at the pulp/froth interface. In the case of large mechanical flotation cells and due to the presence of the froth crowder, the cross section decreases while bubbles move from the interface to the overflow.

![Figure 11. Maximum theoretical carrying capacity at a superficial air rate of 0.4 cm/s.](image)

According to this result the operating condition of the first cell (29.8 tph at the interface level), showed in black point in Figure 11, is far from the maximum theoretical carrying capacity limit. A typical bubble size range of 1-2 mm was assumed for this case, Yianatos et al. (2). Homogeneous bubble swarm was also assumed together with the fact that all the bubbles were 50% loaded with particles of 80 microns.
Figure 12 shows the volumetric flowrate at the overflow level versus the froth depth for the 9 cells in the bank.

![Graph showing volumetric flowrate vs. froth depth]

Figure 12. Volumetric concentrate flowrate versus froth depth.

Here it was observed that the first cell and the last cell of a bank show the higher volumetric flowrates. This was caused because, the first cell has the larger amount of floatable solids and the last cell has the lower froth depth.

**Metallurgical characterization**

According to the automatic data collected every two hours, the operating conditions of the rougher bank during testing day are shown in Figure 13. Here, the fractional mass recovery into the concentrate, C/F, is presented against the enrichment ratio (concentrate grade/feed grade) for the rougher bank.

Here a significant spread was observed in the fractional mass flowrate as well as in terms of the enrichment ratio for the concentrate, during a shift. Particularly, two kinds of operating conditions can be observed. Firstly, a rather stable operation with rougher recoveries around 90±4%, which are the normal values for this operation typically running with 5 rougher flotation banks in parallel. Secondly, due to operational troubles in one bank, an abnormal condition was observed when the circuit was temporary operating with only 4 banks. The rougher circuit then decreased the mineral throughput from 4,500 to 3,500 tph making it possible for it to recover the normal performance for this condition.
Figure 13. Concentrate mass recovery versus enrichment ratio in rougher bank

**Rougher scale-up evaluation**

The scale-up problem is complex and this is due to the different factors affecting the hydrodynamics, kinetics and transport conditions in the pulp and froth zones, Yianatos et al. (5). The development of large size cells requires an in plant practical evaluation to verify the validity of the scale-up factors.

In order to estimate scale-up factors from batch to plant operation a key factor is, for comparison purposes, the selection of the recovery. The simplest approach is to consider the overall rougher recovery and compare this value with the batch test under the same conditions. However, final cells show a rather flat condition in normal flotation bank operation, because their contribution to the overall recovery is relatively small. Thus, there is a large uncertainty when relating time and recovery from plant operation. Also, individual cell operation can significantly affect the overall residence time in the bank. In order to avoid these constraints, a new approach has been proposed, one where the comparison between batch and plant operation is made at the optimum separation point and can be observed from the corresponding separability curves, Yianatos et al. (5). This approach was tested at the El Salvador concentrator in a rougher circuit consisting of 5 banks with 9 cells, each 42.5 m\(^3\), using data from plant and batch tests over a period of 9 months.

Because the data was limited here, an alternative approach was developed. In this new approach the flotation bank operation is firstly modeled so as to have an expected smooth behavior for the recovery versus time relationship, both in plant and batch scale. The point of comparison was then selected considering a dimensionless recovery \(\eta_P\).

Table V shows the results, derived from the adjusted model using the rougher simulator, of cumulative recovery for time along the bank of cells,

<table>
<thead>
<tr>
<th>Table V. Rougher cumulative recovery</th>
</tr>
</thead>
<tbody>
<tr>
<td>Cell</td>
</tr>
<tr>
<td>------</td>
</tr>
<tr>
<td>1</td>
</tr>
<tr>
<td>2</td>
</tr>
<tr>
<td>3</td>
</tr>
<tr>
<td>4</td>
</tr>
<tr>
<td>5</td>
</tr>
<tr>
<td>6</td>
</tr>
<tr>
<td>7</td>
</tr>
<tr>
<td>8</td>
</tr>
<tr>
<td>9</td>
</tr>
</tbody>
</table>
Tables IV and V show that, the copper recovery at infinite time was 90.1%, and the final copper recovery in the rougher bank was 87.3%. Thus, the dimensionless recovery, \( \eta_p \), achieved in plant operation was,

\[
\eta_p = \frac{R_R}{R_\infty} = \frac{87.3}{90.1} = 0.97
\]

where \( R_R \) is the final cumulative recovery in the flotation bank, achieved at an average residence time \( \tau_P \) of 28.9 min. On the other hand, Table IV showed an infinite time recovery of 95.4% for the batch flotation model. Thus, for the same dimensionless recovery observed in the plant, the recovery for batch operation was, \( R_{eq} \):

\[
R_{eq} = \eta_p R_\infty = 0.97 \times 95.4 = 92.4
\]

The corresponding time for the same dimensionless recovery in a batch operation was \( \tau_B \) equal to 8.8 min, considering the batch rectangular kinetic model, shown in Table III.

In order to compare the kinetic behavior for both operations, the following dimensionless relation was used,

\[
[k_{max} \tau]_{PLANT} = [k_{max} \tau]_{BATCH}
\]

where \( k_{max} \), \( min^{-1} \), represents the kinetic constant for plant and batch operation, and \( \tau \) is the corresponding residence time. Thus, in order to keep an equivalent kinetic condition in both scales, the following relation must be respected,

\[
\frac{k_{max,B}}{k_{max,P}} \frac{\tau_P}{\tau_B} = constant
\]

where the "\( P \)" and "\( B \)" indices respectively correspond to plant and batch operation. Thus, for this operation the scale-up factor was,

\[
\frac{\tau_P}{\tau_B} = \frac{28.9}{8.8} = 3.3
\]

This result agreed with the scale-up factors observed from other studies developed in 42.5 m\(^3\) cells, Yianatos et al. (5).

On the other hand, a direct comparison between the \( k_{max} \) parameters for both models from Table IV gave,

\[
\frac{k_{max,B}}{k_{max,P}} = \frac{3.5}{1.2} = 2.9
\]

The difference observed between the ratios \( k_{max} \) and \( \tau \), equations (5) and (6) is related to the fact that the infinite recovery is different for both operations. Thus, the comparison assuming a dimensionless recovery is more realistic.

An estimate of the impact the mixing condition has upon the time scale-up factor, can be observed in Figure 14. From here, the results observed in the range 85-95% recovery became independent of the final recovery, after 4 perfect mixers in series. In the present study the rougher bank consisted of 9 cells in series, which is closer to the perfect mixers. Thus, from Figure 14 and with respect to the batch operation, a time factor of 1.15 can be attributed to the mixing conditions in the
bank. A comparison of this value with the overall scale-up factor, observed from plant data, clearly shows that the mixing condition has a minor impact on the overall scale-up factor in a bank of more than 4 cells. Thus, the main components of the scale-up factor are due to differences in pulp and froth mineral transport, bubble size distribution, gas residence time and effective contact between particles and bubbles, among others, Deglon (6).

**Figure 14.** Time scale-up factors at different mixing conditions and same kinetic rate constant

![Graph showing time scale-up factors](image)

**CONCLUSIONS**

A large size rougher flotation bank, consisting of 9 self-aerated cells each of 160m$^3$ was characterized.

The rectangular kinetic flotation model was found appropriate in describing the flotation bank performance.

There is no limitation in the carrying capacity, despite the large amount of solids recovered in the first cell, this was verified.

The rougher flotation operation reached the metallurgical target predicted from design and operates near the optimal separability criteria.

A scale-up factor of 3.3 was determined to relate the laboratory batch flotation to the plant operation in a bank of 160 m$^3$ cells.

**ACKNOWLEDGEMENTS**

The authors are grateful to Minera Escondida Ltda. for providing access to their plant and for valuable assistance in the experimental work. Funding for process modelling and control research was provided by CONICYT, project Fondecyt 1040100, and Santa María University, project 270322.

PROCEMIN 2004, Santiago, Chile. www.procemin.cl
REFERENCES


