improvement and extension
of the use of flotation columns
flowsheet optimization

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In bibliography, this report should be cited as follows:

The report presents the work carried out by the partners since the mid term of the project (April 30th, 1994).

The salient issues of the work carried out during the period have been the following:

- **TASK 1: Improving selectivity through the use of gases**

  End of the work carried out with gases during the column flotation stage, as a consequence of the conclusions of the mid term meeting.

- **TASK 2: Froth Control**

  Numerous tests have been carried out directly inside the APIRSA industrial plant, with the opportunity of the industrial tests carried out with BRGM and CISA pilot plant columns.

  Different interesting results have been noticed, but the poor reproducibility of the tests does not allow easy conclusions.

- **TASK 3: Study of alternative bubble generation systems**

  The industrial tests carried out with the Microcel TM technology, during the 4th semester of the project, both in Sweden (zinc flotation from Vieille Montagne) and in Spain (Zn cleaner and Zn rougher) have not been fully analysed, but the first results are very encouraging, confirming that column metallurgical performance are very dependant on bubble generation technology.

  As a result of the IMPEXFLOTTCOL project, LKAB, on the basis of the test programme carried out during the 2nd and the 3rd semesters, and APIRSA, on the basis of the test programme carried out during the 4th semester, are both considering, in their pending expansion programme, to implement flotation columns equipped with the new Microcel technology.

- **TASK 4: Flowsheet tests and selection**

  On the fine grained ore, the work carried out at APIRSA, by the three partners (APIRSA, BRGM and Control Int S.A.) is still under analysis.

  Nevertheless, the following conclusions could yet be submitted:

  . possibility to produce in only one stage a Zn final concentrate with a good grade (> 47 % Zn) and a recovery (> 35 %) which could be considered as acceptable taking into consideration the poor liberation (before regrinding step);

  . possibility to simplify the Cu circuit with the production of a better quality concentrate.
Concerning the flotation of coarse grained ores, the experimental work carried out on Zn ores at Ammeberg Mining A.B. is still under evaluation.

- TASK 5: Ionic Flotation

The work carried out shows the effectiveness of column flotation to float molybdenum precipitate, with defined bubble diameters.
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1. TECHNICAL OVERVIEW
CRITICAL OVERVIEW OF PROGRESS

1.1. TASK 1

During the 4th semester, as decided during the mid-term management meeting, the experimental work carried out in corresponding task has been settled.

1.1.1. Work carried out by Lulea University

Reporting of the work carried out by Lulea University has been completed; the corresponding summary is as follows:

It was shown in a previous technical report that the chemical environment created by different grinding media dominates the pulp potential, and affects the Cu/Zn flotation behaviour. The adequate pulp potential was the key factor to attain good Cu/Zn selectivity. The effect of reactive gases on the pulp environment and on the Cu/Zn flotation performance is the current research target. Thus, several flotation tests in a mechanical laboratory flotation cell were done on a fine-grained complex sulphide ore with high pyrite and pyrrhotite content. After grinding in a stainless steel mill only, the pulp environment was modified by using reactive gases such as 5% hydrogen in the preconditioning and flotation steps and sulphur dioxide instead of sodium bisulphite as Zn-depressant. A three-factor statistical design with two levels of preconditioning, three levels of flotation gas type, and two levels of modifiers was used. Also, a few tests without any Zn-depressant at all, and varying dosage of sulphur dioxide were carried out. The interpretation of the reaction mechanism in the flotation system is discussed in relation to pulp chemistry data measurement, which was processed by a mutivariate analysis method. Statistical results indicate that after grinding in a stainless steel mill, the use of a reactive gas with reducing characteristics in the pre-treatment and flotation can vary the pulp potential only within a limited extent, and does not dramatically affect the Cu-Pb/Zn flotation behaviour. Flotation of copper and lead minerals takes place in the pulp potential range of -80 mV to +50 mV vs SCE. However, the flotation gases seem to affect the flotability of silver minerals to some extent. Best silver recovery was attained under the situation of sulphur dioxide as Zn-depressant and air flotation. Furthermore, it is shown that a Zn-depressant is necessary for good Cu-Pb/Zn selectivity, and 80 ml/min sulphur dioxide can successfully substitute 200 g/t sodium for Zn-depressant, which points out a way for the utilisation of surplus sulphur dioxide in sulphuric acid factories.

1.1.2. Experimental work carried out by Apirsa and BRGM

The experimental work carried out by Apirsa and BRGM has been yet previously reported.

A consolidation of the work carried out by the different partners will be consolidated in a Technical Report (T3) to be presented during the next semester.
1.2. TASK 2 : FROTH CONTROL

Tests were carried out at the BRGM pilot plant and in the APIRSA plant, both with Aznalcollar ore.

1.2.1. Pilot plant results

The pilot plant column was used for Cu roughing and special attention was devoted to Cu/Pb selectivity.
The effect of addition of chemical reagents in wash water and in bubble generator has been tested.

a) Effect of MgSO$_4$ in wash water

The promising test realized in January 1994 was not confirmed: the effect of magnesium sulfate is not yet clear.

b) Effect of collector (AXK) in wash water

Addition of collector (100 g/t) in wash water has a clear and surprisingly effect: Cu recovery drops from 68% to 47%, while Cu and Pb grades are increased: respectively 2.9 and 7.9% without collector; 3.9 and 9.6% with collector. Cu/Pb selectivity increases slightly from 1.5 to 1.9. Cu/Zn and Cu/Fe selectivities are more significantly increased: respectively from 1.9 to 9.0 and from 24.7 to 41.0.

The size of the bubbles in the froth zone is around 5 mm with collector and 2 mm without collector: this explains why the recovery decreases. The role of collector in decreasing the bubble size has to be explained.

c) Effect of collector in bubble generator (Microcel system)

Standard collector dosage in the feed was 150 g/t.
Addition of 30 g/t of AXK was carried out either in the feed, either in the bubble generator.
As in the wash water, collector in the bubble generator increases the size of the bubbles, compared with those when collector is added in the feed.

The effect is the increasing of Cu and Pb grades: respectively from 2.7 to 4.4% and from 6.1 to 10.6%. Cu/Pb selectivity index gains 0.7 points (2.1 to 2.8).

Zn grade remains constant and Fe grade decreases slightly.

Surprisingly, Cu recovery increases also from 65% to 71%.
1.2.2. Tests in Apirsa plant

Tests were carried out with the CISA column (30 cm diameter, 7 m high) on the third Zn cleaning, and with the BRGM column (6 cm diameter, 6 m high) on Zn roughing, Cu first cleaning. The froth sampling apparatus has been tested on the Cu cleaning.

a) Third Zn cleaning (CISA column)

Influence of physical parameters (wash water rate, height of froth, air rate) on column performance was studied through a factorial design. A very important variability of results has been observed because of the non-controlled parameters (pulp density, size distribution). It has been observed also that all points were approximatively on the same grade-recovery curve.

• Effect of diminution of the volume of the froth

While the height of froth remains constant, the volume of the froth is decreased by immersion of a kind of floater. Its dimensions were calculated so that the volume of the froth was reduced of one half. The results are presented in the following table.

<table>
<thead>
<tr>
<th></th>
<th>Factorial design (11 points)</th>
<th>Volume 100% (3 points)</th>
<th>Volume 50% (3 points)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Selectivity index</td>
<td>1.86</td>
<td>1.94</td>
<td>2.14</td>
</tr>
<tr>
<td>Zn recovery (%)</td>
<td>43.1</td>
<td>32.3</td>
<td>42.1</td>
</tr>
<tr>
<td>Zn grade % (Concentrate - Feed)</td>
<td>6.4</td>
<td>7.1</td>
<td>7.4</td>
</tr>
</tbody>
</table>

It seems that reducing the volume of the froth causes a slight increase in selectivity. An explanation could be the decrease of the residence time of particles in the froth, while washing remains efficient.

• Effect of reagents in wash water or in bubble generator

No clear effect of collector nor MgSO₄ was observed in wash water or in bubble generator.

b) Zn roughing (BRGM column)

MgSO₄ was not tested at this step.

Collector (EXNa) was added in wash water and in bubble generator. The effects were compared with the addition of collector in the feed. The following table shows that adding collector in the bubble generator could be a more efficient way to increase recovery without detrimental effect on the grade of the concentrate.
Flotation columns: improvement and extension of their use

<table>
<thead>
<tr>
<th></th>
<th>No collector (3 points)</th>
<th>Collector in the feed (2 points)</th>
<th>Collector in the bubble generator (turbo-air) (3 points)</th>
<th>Collector in wash water (1 point)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Selectivity index</td>
<td>38.7</td>
<td>30.5</td>
<td>34.0</td>
<td>23</td>
</tr>
<tr>
<td>Zn Recovery (%)</td>
<td>52.3</td>
<td>75.0</td>
<td>62.7</td>
<td>60.8</td>
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<tr>
<td>Zn Grade %</td>
<td>40.6</td>
<td>29.9</td>
<td>35.4</td>
<td>33.5</td>
</tr>
</tbody>
</table>

Testing of froth sampling apparatus

First tests with the froth sampling apparatus were carried out at Apirsa. Some difficulties were encountered while moving the cutting plates. However, samples of solids could be collected at five different heights. Following tables show the increased grades of copper and decreased grades of lead when distance to pulp/froth interface increases.

<table>
<thead>
<tr>
<th>Distance from pulp/froth interface (cm)</th>
<th>5</th>
<th>15</th>
<th>25</th>
<th>35</th>
<th>45</th>
</tr>
</thead>
<tbody>
<tr>
<td>Mass of solid volume of froth (g/l)</td>
<td>16.2</td>
<td>13.2</td>
<td>11.3</td>
<td>9.3</td>
<td>9.9</td>
</tr>
<tr>
<td>Cu grade (%)</td>
<td>18.6</td>
<td>16.3</td>
<td>21.3</td>
<td>22.2</td>
<td>23.8</td>
</tr>
<tr>
<td>Pb grade (%)</td>
<td>1.65</td>
<td>1.81</td>
<td>1.51</td>
<td>1.39</td>
<td>1.31</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Distance from pulp/froth interface (cm)</th>
<th>-5</th>
<th>5</th>
<th>15</th>
<th>25</th>
<th>35</th>
</tr>
</thead>
<tbody>
<tr>
<td>Mass of solid volume of froth (g/l)</td>
<td>18.5</td>
<td>11.3</td>
<td>7.6</td>
<td>7.9</td>
<td>8.6</td>
</tr>
<tr>
<td>Cu grade %</td>
<td>16.2</td>
<td>22.1</td>
<td>23.1</td>
<td>24.0</td>
<td>25.4</td>
</tr>
<tr>
<td>Pb grade %</td>
<td>2.92</td>
<td>2.98</td>
<td>2.95</td>
<td>2.96</td>
<td>2.76</td>
</tr>
</tbody>
</table>

1.2.3. Conclusions

Zn cleaning (Apirsa): significant improvement (around ten points) of the recovery was observed on three tests when the volume of the froth was increased.

Zn rougher (Apirsa): slight advantage of introducing collector in the bubble generator (Turbo-air) rather than in the conditioner.

Cu rougher (in BRGM): selectivity Cu/Pb increases when collector is added in the bubble generator (Microcel) rather than in the conditioner.

To confirm some of these results, laboratory batch tests will be carried out.
1.3. TASK 3: STUDY OF ALTERNATIVE BUBBLE GENERATION SYSTEMS

Sub-task 3.3: Testing with various mineral systems under operational plant conditions

The programme for industrial testing for comparing alternative bubble generation systems has been finalized (provided it will not be disturbed anymore by changes made by mining companies). It will include:

- with the pilot column manufactured within the frame of Sub-task 3.2:
  - two complex sulphide base metal plants: APIRSA (Spain) - Zn cleaner, Zn rougher, GUEMASSA (Morocco) - Cu cleaner, probably Zn or Pb circuits,
  - one phosphate plant: KHOURIBGA,
  - one fluorite plant: EL HAMMAM;

- with the MITU mobile pilot plant, in Sweden:
  - one iron ore apatite flotation plant (LKAB),
  - one zinc flotation plant (Union Miniere).

This programme provides a relatively comprehensive coverage of various realms of applications (mineral types, particle size, ...). It will generate a guideline database for appropriate selection of bubble generation technology.

The first testing campaign with the pilot column manufactured within the frame of Sub-task 3.2 has been conducted from June to October 1994 in the APIRSA plant at Aznalcollar in Spain. The data need to be reducted before to be ready for analysis. It confirms that column metallurgical performance is highly dependent on bubble generation technology. The Microcel™ technology turns out to be more amenable to be adapted for the flotation of very fine particle size distribution. Recovery up to 80 % has been achieved in the zinc cleaning circuit.

The next testing campaign will take place in the complex sulfide (Cu, Pb & Zn) plant of Guemassa (ONA group) in Morocco. As scheduled in the last report, the pilot column was shipped and integrated in the Zinc circuit of the APIRSA plant in June 1994. The slurry tests were conducted on the roughing and cleaning stages of the Zinc circuit. The complete data analysis will be presented in the next report.

The next testing campaign will take place in the complex sulfide plant of Guemassa. The column should be installed on the copper circuit by the end of November and the first testing should start in December. The next sites for the column testing will be also located in Morocco and will include the El Hammam plant (ONA group) which is processing fluorite and a phosphate-based operation within the OCP group.

Within the frame of task 4.2, the MITU mobile pilot plant was installed in the Zingruvan plant at Union Miniere (Sweden). A specific assistance was given by Control International in August for the start-up of the Microcel system. Some comparative data between the porous sparger and the Microcel system should be available in the next report.
1.4. TASK 4: FLOWSHEET TEST AND SELECTION

1.4.1. Tests on very fined grained ores

The work has been based on the pilot plant operations carried out in the Apirsa plant, by a joint Apirsa-BRGM and CISA team, from October 3rd to November 9th.

Two pilot plant size columns have been transported from Orléans and erected inside the industrial plant; it was then possible to process flows proceeding from different points of the flowsheet.

The BRGM column (D=6.35 cm, H= 6m) have been tested in different parts of the circuits: Zn rougher, Cu rougher, Cu first cleaner.

a) Zn rougher (fig. 1)

Column has been tested for the Zn roughing, alone and also included in circuits with pilot scale conventional cells, acting either as cleaner or as scavenger. The results show the possibility for column to operate at roughing stage for zinc and to produce final concentrate. Good grades (>47%) with acceptable recoveries (>35%) have been obtained. A factorial program has been carried out to determine the parameters required to reproduce this kind of results. It seems possible to describe three flowsheets for the Zinc circuit:

- Flowsheet 1

The column is used for roughing and produces a final concentrate; the tailing are processed in a conventional circuit. In this case, the volume to be treated in column is important. The main advantage is the production of a final concentrate. One of the drawback could be the possible decrease of flexibility with one column compared to banks of cells (to be checked).

Zn rougher (1/2)

- On BRGM column
  - Possibility to obtain a final concentrate in one step (Zn > 47%, Rec > 35%)
  - Three flowsheets proposed for the Zn circuit

1 (CISA Ø 30 cm x 10 m and BRGM Ø 6.35 cm x 6 m columns).
Fig. 1 - Flowsheet of Zn circuit
• Flowsheet 2

The concentrate of the three first roughing cells feeds the column, which produces a final concentrate. The column tailings are sent to the first cleaning cells. For this flowsheet the volume of pulp to be treated is quite smaller. The main advantage is the lower circulating load in the conventional circuit. The other advantage is the direct production of a final concentrate.

The corresponding rough results as well as the full corresponding flowsheets are presented in Tables 1 and 2.

• Flowsheet 3

The column is used exactly as conventional cells in roughing stage. Column tailings are treated in scavenging cells, and concentrates are sent to the first cleaner. In this case the main advantage compare to conventional cells is the lower energy consumption. No test has been completed following such circuit.
### Table 1 - Pilot plant tests at Apirsa - Zn circuit results.

<table>
<thead>
<tr>
<th>Test number</th>
<th>Feed flowrate</th>
<th>Wash water flowrate</th>
<th>Air rate</th>
<th>MIBC dosage</th>
<th>Pulp level</th>
<th>Froth height</th>
<th>pH</th>
<th>Zn grade</th>
<th>Solids flowrate</th>
<th>Zn grade</th>
<th>Solids flowrate</th>
<th>Zn grade</th>
<th>Solids recovery</th>
<th>Solids flowrate</th>
<th>Zn grade</th>
<th>Observation</th>
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<tr>
<td>PC1</td>
<td>25</td>
<td>400</td>
<td>3500</td>
<td>0</td>
<td>525</td>
<td>37</td>
<td>-</td>
<td>5.77</td>
<td>26</td>
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<td>21.77</td>
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</table>

**Fixed conditions:**
- Bubble generator = Turbo air
- Conditioning ENH = 40 g/t
- CuSO₄ = 400 g/t

**Observation**:
- Preliminary test
- Factorial program

**Notes**:
- Feed (Measured) and Feed (Calculated) columns represent the actual and calculated values of feed materials.
- Concentrate and Tailings columns show the recovery and grade of Zn in the concentrate and tailings respectively.
- The values indicate the efficiency of the flotation process at different conditions.
Table 2 - Pilot pant tests at Apirsa - Cu circuit results and froth control on Zn circuit.
b) Cu circuit (fig. 2)

The first conclusion concerning roughing is that, very few tests having been made, a final concentrate has not been obtained, but however the results are better than with the conventional cells: the selectivity is improved.

**Cu rougher**

- Very few tests (Only 4)
- No final concentrate has been obtained
- Good selectivity in any case

<table>
<thead>
<tr>
<th></th>
<th>PC27</th>
<th>PC28</th>
<th>PC29</th>
<th>PC30</th>
<th>Industrial rougher concentrate</th>
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<tbody>
<tr>
<td>%Cu</td>
<td>15.67</td>
<td>10.93</td>
<td>5.55</td>
<td>5.96</td>
<td>2.55</td>
</tr>
<tr>
<td>%Pb</td>
<td>1.85</td>
<td>1.39</td>
<td>1.51</td>
<td>1.5</td>
<td>2.35</td>
</tr>
<tr>
<td>Ag (ppm)</td>
<td>878</td>
<td>585</td>
<td>347</td>
<td>334</td>
<td>169</td>
</tr>
<tr>
<td>Cu rec.</td>
<td>24.2</td>
<td>25.6</td>
<td>37.9</td>
<td>35.9</td>
<td>-</td>
</tr>
</tbody>
</table>

For the first cleaning stage, a final concentrate is easy to get. These concentrates are very good in term of grades (22<Cu<25.5, Pb<2.5, Zn<3). More analysis of the impurities will be carried out. In terms of recovery, results are also good (between 25 and 35). The following table shows the comparison of the current industrial concentrate composition and the PC41 concentrate.

<table>
<thead>
<tr>
<th></th>
<th>Industrial concentrate</th>
<th>PC41 concentrate</th>
</tr>
</thead>
<tbody>
<tr>
<td>% Cu</td>
<td>19.91</td>
<td>25.58</td>
</tr>
<tr>
<td>% Pb</td>
<td>4.5</td>
<td>1.09</td>
</tr>
<tr>
<td>% Zn</td>
<td>7.83</td>
<td>1.51</td>
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<tr>
<td>Ag (ppm)</td>
<td>1361</td>
<td>1424</td>
</tr>
<tr>
<td>% Sb</td>
<td>1.77</td>
<td>0.97</td>
</tr>
</tbody>
</table>
Grinding and Copper Flotation Circuits

Fig. 2 - Flowsheet of Cu circuit.
Flotation columns: improvement and extension of their use

Cu first cleaner

- Final concentrate obtained
- Very good selectivity compared to the industrial concentrate
- Interesting recovery

<table>
<thead>
<tr>
<th></th>
<th>Industrial concentrate</th>
<th>PC41 concentrate</th>
<th>PC44 concentrate</th>
</tr>
</thead>
<tbody>
<tr>
<td>%Cu</td>
<td>19.91</td>
<td>25.58</td>
<td>20.15</td>
</tr>
<tr>
<td>%Pb</td>
<td>4.5</td>
<td>1.09</td>
<td>2.35</td>
</tr>
<tr>
<td>%Zn</td>
<td>7.83</td>
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<td>Ag (ppm)</td>
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<tr>
<td>Cu rec.</td>
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</table>

Modelling

Column modelling is in progress

Sala Agitated mill (SAM) modelling.

Following a campaign of pilot tests carried out with Apirsa ore at BRGM, the importance of percentage of solids, feed flow rate and load of grinding media (cylpebs) have been quantified.

Tests on Residence Time Distribution have been also carried out.

As a conclusion, a model, which is a combination of simplified kinetic and energetic approach has been developed. The model is currently tested.

For reason of equipment availability no industrial tests have been completed with SAM mill inside Apirsa plant.

As a provisional conclusion, we can say that the pilot plant campaign carried out at Apirsa:

- has demonstrated in an industrial environment the possibility of obtaining directly after just a roughing stage a good quality Zn concentrate. The recovery (40-50 % in the best cases), which could be considered as modest if compared to the results obtained previously on a coarse grained ore as Chessy (recovery 95-98 %), seems in the Apirsa case very attractive for a simplification of the flowsheet, with positive impact on both investment and operating cost. Such recovery on Apirsa ore could be explained by the liberation state of the Zn bearers at that stage of the process (in the industrial process, a regrinding stage is required in the downstream part of the flowsheet).

In the case of the Cu circuit such demonstration has not been possible, possibly for a lack of time;

- has demonstrated, in the case of the Cu circuit a very significant improvement of the Cu concentrate quality.
The pros and cons of tests in an industrial plant have been heavily enjoyed and suffered:

- **pros**: work on fresh pulp, in industrial conditions, with operators input.

- **cons**:
  - importance of the experimental work in terms of equipment needs, manpower involved,
  - variations of the quality of the flows,
  - difficulties of arranging reflection and analysis periods between the experimental tests.

### 1.4.2. Tests on coarse grained ores

The testwork at Ammeberg Mining AB (previously Union Minière Sverige AB) is now finished but the results are still under evaluation. The pilot plant will be moved to Boliden Mineral AB in Boliden for further testwork at the central concentrator in Boliden. A final report of all pilot plant activities will be compiled and submitted to the next Management Committee Meeting in March, 1995.

A new test-program for further pilot plant tests in Boliden has been prepared and it will be finalized during December. Pilot plant tests will be carried out with columns as first roughers in the CuPb-flotation of Petiknäs ore. Three columns will be run in parallel to be able to study the influence of different operating parameters on the column performance. Different bubble generators will be studied, complementary to other work, carried out in Task 3 (Study of alternative bubble generation systems).

### 1.4.3. Economic evaluation (cf. addendum 1)

Since the last report and according to the Impexflotcol WORK PROGRAMME (see Human resources allocation table 1/32) activities of UTS-DINMA partner were mainly directed toward the objectives of Task 4 (Flowsheet test and selection) Sub-task 4.3 (Economic evaluation).

In particular:

A procedure was derived that allows the selection of the configuration and the sizing of complex cell-based flotation circuits. The method is driven by an user-defined financial objective function that incorporates metallurgical efficiency through concentrate sales revenues and elements of both capital and operating costs. The suggested design method is likely to be superior to the traditional empirical procedures that point to maximise metallurgical performances without including in a formalised way the cost side of the circuit; technically feasible solutions too costly for the metallurgical results they allow are rejected and the best techno-economic solution is iteratively searched.
Traditionally flotation circuits are sized with the objective of obtaining a threshold metallurgical recovery and grade. The problem is often dealt with the aid of steady state simulators allowing to evaluate alternative flowsheet configurations. Prior work on the algorithms for configuration / optimisation of complex froth flotation circuit has been recently reviewed by Yingling J. (1993). Economic aspects are surely taken into consideration during plant design: as an example, plants based on larger cells cost less and to some extent allow lower operating costs.

However the economic side and its technical consequences are not clearly formalised into a decision procedure permitting to rank alternative processing solutions. So that it is not possible to answer questions as:

- is the cost (capital and operating) of extra cells in a rougher or scavenger bank compensated by the extra recovery it allows?
- is the increase in grade that an extra bank of cleaner allows justified by the related extra costs of the asset?
- which is the best cell-number-size combination from a techno-economic point of view?
- is the extra cost of using smaller flotation cells while maintaining constant the overall bank volume offset by the better metallurgical results related to a restrained "short-circuiting" effect?

A framework was set out for a scaling up procedure that allows to derive the configuration of a flotation circuit that is optimal according to both a metallurgical and a financial point of view. Indeed any metallurgical result is obtained at a cost and the two aspects should be weighted one versus the other. The procedure can help ranking processing solutions on the ground of their financial efficiency.

Engineers can analyse complex processing problems and make better and quicker judgements if they can use more accurate information available through the use of computerised design procedures. These procedures support rather than replace engineer's judgement and improve the effectiveness not the efficiency of the design process. They simply solve repetitive tasks for which standard and unique solutions exist. Effectiveness is also improved by allowing engineers to perform tasks in less time, with less effort and at a lower cost and improving the quality of the data they use to take final decisions. Several 'what if' scenarios can be examined and the final decision can be taken on a trial and error process.

Future work will lead to the extension of the procedure to the design of hybrid plants that is plants including columns in the flowsheet.

1.5. TASK 5: APPLICATION OF COLUMNS TO IONIC FLOTATION

1.5.1. Ionic flotation of molybdenum

a) Reminder

Previous studies have shown that the recovery of "Mo-α-Box" precipitate depends on both bubble diameter and number.
Use of various types of bubble generators allows the production of different bubble diameters. An other way can be the change of running parameters as recirculation pump speed (recycling rate), air flow rate, frother concentration....

In the conclusion of the former report, a classification of bubble generators has been proposed. In the case of molybdenum flotation from solutions with 2 to 5 $10^{-3}$ M/l concentration, the more efficient generators are Flotaire and Microcel types. With these generators, the bubble diameter range from 0,3 to 1 mm. The gas holdup can vary from 3 to 26%. The working of these generators is stable compared to Imox and the distribution of bubble sizes is more homogenous.

In literature, the molybdenum ionic flotation in column was not studied either in laboratory or in pilot plant scales. The difficulty encountered with molybdenum flotation proceeds from the no-hydrophobicity of the reagent used for the precipitate flotation.

To solve this problem, we proposed the flotation with strictly definite bubble diameters. In this case, the molybdenum precipitate flotation is governed by the trapping of flocs by bubble under the influence of microturbulences. A semi-experimental method allows the bubble diameter determination with a permanent control of the gas holdup.

b) Experimental procedures

- **Apparatus** (fig. 3)

  The molybdenum is previously precipitated in a slowly agitated tank with a controlled speed. This tank is fed with:

  - a variable speed peristaltic pump for the molybdenum sulfuric solution,
  - a variable speed peristaltic pump for the collector.

  The flotation column is a 75 mm diameter by 3 m high unit constituted by 6 pieces of 50 cm high with the following equipment:

  - a variable speed peristaltic pump to feed the column with the "Mo-BOx" precipitate suspension,
  - a variable speed peristaltic pump for the epurated solution,
  - choice of bubble generators,
  - a control panel with air rotameters, water rotameters, air and water pressure regulators and a controller for monitoring level.

  The column is used without any washwater. The bubble generators are disposed in the closed circuit of the recirculation pump placed on the column basis. The stability of working is done by the regulation of the speed of the output pump as the speed of the feed pump is fixed manually.

  Gas holdup is measured with two water filled manometers placed at different levels on the column $\Delta L$ is the distance between the pressure tapping points $= 700$ mm.

  The two types of bubble generators principally used are the Flotaire (external sparger) and the Microcel (semi-internal sparger).
Sodium molybdate + Sulfuric acid
Conditioning tank (10 minutes)
Flocks formation
Floated product (precipitate)
\( \alpha \)-benzoïne-oxime

Feed flowrate - 1 l/min
Flocks in froth
Few flocks
Clear solution
Unfloated product (clear solution)

Fig. 3 - Schematic diagram of ionic flotation in column.
• Operating

- initial concentration of reagents:
  . sulfuric acid and α Benzoine-oxime = 0.1 M/l,
  . sodium molybdate = from 2 to 5 $10^{-3}$ M/l;

- feed flowrates:
  . sulfuric molybdate = from 0.833 to 1.25 l/min,
  . α benzoine-oxime solution = from 0.033 to 0.22 l/min,
  . precipitate suspension = from 0.833 to 1.25 l/min;

- conditioning in agitated tank = 10.6 min. To maintain a conditioning time of 10.6 min, a molybdenum precipitate suspension of 10 liters is initially prepared in the agitated tank. For a test, the total volume of precipitate suspension is 40 liters to perform flotation during 10 min;

- frothers: C7 (polyglycols) and Montanol (aliphatic alcohol) from Hoechst, Aerofroth 65 (polyglycols) from Cyanamid in concentration from 0 to 20 mg/l.

The following parameters are controlled during test:

  - feed flowrate ($D_{\text{air}}$)
  - air flowrate ($D_{\text{air}}$)
  - manometric pressure difference ($\Delta H$)
  - solid concentration in the feed and in the epurated solution
  - water output in the froth
  - recirculation pump flowrate

The following parameters are calculated:

- recovery of precipitate after centrifuging and weighting of samples from the feed and the solution output.
  
  \[ R = \frac{(P_{\text{initial}} - P_{\text{final}})}{P_{\text{initial}}} \]

- gaz holdup by the manometric pressure difference ($\Delta H$)
  
  \[ \varepsilon_g = 1 - \left(\frac{\rho_w}{\rho_{\text{susp}}}\right) \left(1 - \frac{\Delta H}{\Delta L}\right) \]

In our case, $\rho_w$ is very close from $\rho_{\text{susp}}$. So we can write $r_w = \rho_{\text{susp}}$ and equation then reduces to

\[ \varepsilon_g = \frac{\Delta H}{\Delta L} \]

with $\varepsilon_g$ = gaz holdup, $\rho_w$ et $\rho_{\text{susp}}$ respectively the densities of water and suspension, $\Delta H$ the manometric pressure difference during the test.

- mean diameters of bubble

To estimate the mean diameter of bubble, a mathematical expansion from the method of Finch and Dobby [4] was used.
During test, the following parameters are measured: $\Delta H$, $D_{air}$ et $D_{al}$. For calculations we need to determine the gas holdup ($\varepsilon_g$) and the superficial velocity [flowrate/column cross sectional area ($S_c$)] of gas ($J_g$) and liquid ($J_l$).

By definition:

$$J_g = \frac{D_{air}}{S_c}$$

$$J_l = \frac{D_{l}}{S_c}$$

and by calculation $\varepsilon_g = \frac{\Delta H}{\Delta L}$

The slip velocity then can be calculated experimentally. The slip velocity is the velocity of the gas phase relative to the liquid phase and is defined as:

$$\left(\frac{U_{sg}}{\varepsilon_g}\right)_{exp} = \frac{J_g}{\varepsilon_g} + \frac{J_l}{1 - \varepsilon_g}$$

The slip velocity is related to the system variables. For bubble sizes with $d_b \leq 2$ mm ($Re_b \leq 500$), a suitable expression is an adaptation of the multi-species hindered settling equation of Masliyah [5] written below for the gas-slurry system:

$$\left(\frac{U_{sg}}{\varepsilon_g}\right)_{cal} = \frac{gd_b^2(q_{sl} - \varphi_b)F(1 - \varepsilon_g)}{18\mu_{sl}} \left(1 + 0.15Re_{bs}^{-0.687}\right)$$

where

$$Re_{bs} = \frac{dbU_{sg}q_{sl}(1 - \varepsilon_g)}{\mu_{sl}}$$

et $F(1 - \varepsilon_g) = (1 \varepsilon_g)^{m-1}$

From Richardson and Zaki (1954):

$$m = [4.45 + 18 \frac{dc}{db}]Re_b^{-0.1}$$

if $1 < Re_b < 200$

$$m = 4.45 Re_b^{-0.1}$$

if $200 < Re_b < 500$

et $Re_b = \frac{d_bU_{b}q_{sl}}{\mu_{sl}}$

where $d_b$ is the bubble diameter

$\mu_{sl}$ the slurry viscosity

$U_{b}$ the terminal rise velocity of a single bubble.
As there is no theoretic relation between $d_b$ and $U_b$, the problem of the determination of $d_b$ is more difficult. Dobby and Finch use another empiric equation between $U_b$ and $U_{sg}$. For our calculation, the $U_b$ determination was done independently of $U_{sg}$.

In literature, several empiric relations could be found between the relative velocity of bubbles and their diameters (for example Clift and al., 1978). By modeling these data for bubble totally covered by reagent, a polynomial regression equation can be found:

$$U_b = A d_b^3 + B d_b^2 + C d_b$$

with a confidence interval of 95% A, B and C are known constants.

As $U_b = f(d_b)$, when the bubble diameter $d_b$ can be estimated, $U_b$, $Re_b$, $m$, $F$ and finally the relative velocity ($U_{sg})_{cal}$ can be calculated.

This calculated value of $U_{sg}$ has to be compared with the experimental determination ($U_{sg})_{exp}$.

If the two velocities are identical, the mean bubble diameter is determined. Otherwise estimation has to be fit since conformity.

For each test, the estimation of bubble diameter was done. The permanent control of water levels in manometer tubes allow to follow the gas holdup and the bubble diameter evolution.

c) Results and comments

Two molybdenum concentrations have been studied ($2$ and $5 \times 10^{-3}$ M/l).

Several parameters that act on bubble diameter and gas holdup have been tested. They are:

- the air flowrate,
- the feed pump flowrate,
- the recirculation pump flowrate,
- the frother concentration.

The systematic study of each parameter, all other parameters kept constant, on the bubble diameter is finally the result of a combination of all studied parameters.

The first series of tests in the pilot column shows that, for the $2 \times 10^{-3}$ M/l Mo concentration, the bubble diameter influences strongly the recovery. A similar tendency has been observed in the laboratory microcell.

Recoveries over 95% to 100% can be reached for bubbles diameters from 0,3 to 0,4 mm (fig. 4). Total flotation of the precipitate is obtained for relatively high frother concentration (30 mg/l of Aerofroth 65). In these conditions, gas holdup ranges from 17,1 to 21,4%.
To increase the bubble diameter, the frother concentration has to be decreased. Then bubble with diameters of 0.65 mm could be generated but recovery falls down to 80%. If the recovery has to be maintained at a high level of 95%, it is necessary to increase the recirculation pump flowrate to 72 l/h with a frother concentration of 12.5 mg/l. Then gas holdup reaches 18.6%.

But consequently, the water content of the froth increases. Indeed, without frother the water entrainment with the froth is low 0.2 liter for a test (10mn- 40 l feed). If the gas holdup increases from 10% to 21.4% for the same gas flowrate (correlatively and logically the bubble diameter is divided by 2), loss of water by the froth reaches 4 to 5 liters. The precipitate flotation is then total.

The decrease of the gas flowrate and the use of a frother that presents flocculant properties allow a very low water content in the froth with, in the same time, a Mo recovery from 95 to 99% (tests 5 and 6 of the table).

An increase in the feed flowrate disturbs the flotation process and induces a longitudinal circulation of the suspension. If, moreover, the gas flowrate is increased, the flotation recovery is strongly decreased [for example from a recovery of 94.8% with a gas flowrate of 45.7 l/h (test N°7), only 78.5% of the molybdenum is recovered in the froth with a gas flowrate of 86.6 l/h(test N°8)].

To summarize, for a concentration of 2 $10^{-3}$ M/l of molybdenum and in a 3" column, the ionic flotation is highly efficient for feed flowrate from 50 to 75 l/h and gas flowrate from 45 to 67 l/h with a recovery about 95% and without any dilution of the flocs in the froth. The gas holdup may be limited to avoid turbulences and pulp circulations in the column that induce coalescences and deterioration of the bubble sizes distribution.

Flotation of molybdenum precipitates from 5 $10^{-3}$ M/l solution. The Flotaire and Microcel bubbles spargers were tested.

With the Flotaire generator and with Aerofroth 65 as frother (15 mg/l), the Mo recovery is 94.9% (gas holdup = 18.6% and mean bubbles diameter 0.38 mm). A decrease of the gas flowrate induces a decrease of the bubble diameter and, in the same time, of the recovery to 84.5%. Water entrainment in the froth is always high in these conditions.

A systematic study of the bubble diameter generated by a Microcel sparger with a constant gas flowrate shows that the flotation recovery decreases in the same time with the bubble diameter (fig. 5). The bubble diameter varies when the recirculation pump flowrate changes.

An increase in the gas flowrate from 56.5 to 67.3 l/h gives a total flotation of molybdenum for bubbles diameter of 0.5 mm. In this case, the froth is quite dry.

An increase in the recirculation pump velocity increases the gas holdup to 16.4%. Then, the bubble diameter is smaller and flotation recovery decreases from 99 to 80.9% (fig. 6).

These last results confirm the studies in the laboratory microcell. With high molybdenum concentration, flotation occurs with definite bubble diameter (> 450 μm) when happens the microturbulence effect.
Fig. 4 - Precipitate recovery as a function of the bubble diameter

\[ C(\text{Mo}) = 2 \times 10^{-3} \text{ M/l, } D_{\text{air}} = 50 \text{ l/h} \]

Fig. 5 - Precipitate recovery as a function of the bubble diameter.

\[ C(\text{Mo}) = 5 \times 10^{-3} \text{ M/l, } D_{\text{air}} = 56.5 \text{ l/h, } D_{\text{air}} = 50 \text{ l/h} \]
1.5.2. Complementary study on the precipitation of Cr (VI) in sulphuric acid media

In the previous report it was shown that ion flotation is governed by different mechanisms, mainly the chromium concentration and the molar ratio (f) of collector to Cr(VI), which affect its efficiency. This was also reported in the literature by some authors who gave different values of $\varnothing$ for optimal conditions of ionic flotation of Cr(VI) with quaternary ammonium collectors (Grieves, 1969; Jdid, 1985).

The residual concentration of collector in solution, which do not react with Cr, affects the ion flotation by electrostatic and physicochemical interactions that may occur between the bubbles and the precipitate and the bubbles and the solution. Furthermore, the stability of the foam may be also affected.

The $\varnothing$ ratio depends on the nature of the ionic species of Cr(VI) in the sulphuric acid solution. According to Pourbaix diagram (Pourbaix, 1962), two species can be formed as a function of chromium concentration: $\text{HCrO}_4^-$ and $\text{Cr}_2\text{O}_7^{2-}$. From the thermodynamic data, the determination of Cr(VI) species in $\text{H}_2\text{SO}_4$ media shows the existence of $\text{HCrO}_4^-$, $\text{Cr}_2\text{O}_7^{2-}$, $\text{CrO}_3\text{SO}_4^{3-}$, $\text{H}_2\text{CrO}_4\text{aq.}$ and $\text{CrO}_4^{2-}$. The predominance of each species is given figure 7, for

---

**Fig. 6 - Precipitate recovery as a function of the bubble diameter.**
chromium concentrations of $10^{-4}$, $6 \times 10^{-4}$ and $10^{-3}$ M, and at ionic strength of 1 M. It can be deduced from this figure that 76 to 86% of Cr(VI) forms HCrO$_4^-$, and the proportion of this species decreases when Cr concentration increases, because of the formation of Cr$_2$O$_7^{2-}$ species. However, at pH = 1 to 1.5 only 4 to 5% of dichromate ions are present in sulphuric acid solutions, whereas the species CrO$_3$SO$_4^{2-}$ represents about 18 to 23%.

Then, taking into account these results, it can be written the following reactions for the precipitation of Cr(VI) by cationic collectors (Col) in sulphuric acid media:

$$\text{HCrO}_4^- + \text{Col} = \text{HCrO}_4\text{Col} \quad [1]$$

$$\text{Cr}_2\text{O}_7^{2-} + 2 \text{Col} = (\text{Cr}_2\text{O}_7)(\text{Col})_2 \quad [2]$$

$$\text{CrO}_3\text{SO}_4^{2-} + 2 \text{Col} = (\text{CrO}_3\text{SO}_4)(\text{Col})_2 \quad [3]$$

To elucidate these mechanisms, the precipitate "Cr(VI)-HDTA" (HDTA = Hexa Decyl Trimethyl Ammonium) was characterized by infrared spectroscopy, for $\theta = 1$ and 2. The obtained infrared spectra, figure 8, exhibits strong absorption bands at 780 cm$^{-1}$ corresponding to dichromate species, weak doublet bands at 912 and 888 cm$^{-1}$ corresponding to chromate species and sharped bands at 680 - 610 and 1130 - 1080 cm$^{-1}$

Figure 8 corresponding to sulphate species, probably CrO$_3$SO$_4^{2-}$. Then it can be deduced that the three reactions occur during chromium precipitation with HDTA in sulphuric acid media, but the reaction 2 seems to be predominant, when comparing the intensities of the absorption bands.

Furthermore, during the precipitation of Cr(VI), the pH (fig. 9) and the equilibrium potential (fig. 10) of the solution decrease, from 1.5 to 1 and from 665 to 600 mV, respectively. The diminution of pH would result from the deprotonation of HCrO$_4^-$ according to the reaction 4:

$$\text{HCrO}_4^- = \text{CrO}_4^{2-} + \text{H}^+ \quad [4]$$

Since the species CrO$_4^{2-}$ is unstable at these pH and potential values (Pourbaix, 1962), there is a dimerisation of chromate ions as following (reaction 5):

$$\text{CrO}_4^{2-} = 1/2 \text{Cr}_2\text{O}_7^{2-} + 1/2 \text{H}_2\text{O} \quad [5]$$

Then, reactions 4 and 5 explain that the reaction 2 is predominant.

In the other hand, the decrease of equilibrium potential would result from redox reactions between Cr(VI) and the collector and/or the alcohol (the collector is dissolved in ethanol before its introduction in the sulphuric solution containing Chromium) and this phenomenon is probably responsible of the bubble formation during Cr(VI) precipitation as it has been reported in the last report.
Fig. 7 - Cr (VI) species distribution as a function of pH.
Total Cr Concentration, M/l:
a) $-1 \times 10^{-4}$; b) $-6 \times 10^{-4}$; c) $-10^{-3}$
Fig. 8 - Infrared spectra of Cr (VI) precipitate with HDTA - Br.
Fig. 9 - pH variation as a function of time during Cr (VI) precipitation using ODTA-Br.

Fig. 10 - Eh variation as a function of time during Cr (VI) precipitation using ODTA-Br (F2).
1.6. PLANNED ACTIVITIES

The actual accomplishments for the two first years of the project could be considered as fully satisfactory.

If the results obtained on task 1 have not been as fruitful as sought when starting the project, the results obtained in Tasks 3, 4 and 5 are promising.

Task 2 (froth control) appears as an important task which from a scientific and industrial point of view will not be completed inside the present programme. The results yet obtained demonstrate the importance of froth control, but also the difficulty for obtaining presently industrial application.

Task 3 will be carried on, following schedule, on different ores, in Morocco (polymetallic ore from Guemassa phosphate ore from Knourihga and fluorite ore from El Hammam).

The experimental part of Task 4 has been almost completed. Many analyses and interpretation work remain to be carried out.

The economic analysis of the results has also to be done.

It has to be noticed that additional testwork in Task 4 will be carried out in Sweden and will be financed directly with additional funding from the industrial partner (Boliden).
2. EXPLOITATION ASPECTS

2.1. PRACTICAL APPLICATIONS OF PROJECT RESULTS

For the time being, two practical industrial applications of the project results (in particular Tasks 3 and 4) should be mentioned:

- LKAB, after evaluation of the results of the pilot plant testwork in its facilities, and after a feasibility study proving the economy and process benefit of column flotation has included the purchase of column flotation equipment for the Kiruna dephosphorization circuit in 1995 investment budget (the column will be included in a circuit producing 4 MT per year of iron pellets, for the cleaning of an apatite concentrate);

- following the first results of the tests carried out at Apirsa, the company is considering to retrofit an existing column with the Microcel bubble generator.

After the final analysis of the results of the pilot plant campaign, Apirsa will also consider the possible impacts of the introduction of columns inside the new "Los Frailes project".

Sucessful results will probably have a positive impact on the other iberian complexe sulphide plants.

Oner another hand, positive impact of the demonstration work to be carried out in Morocco should be expected.

In the field of ionic flotation, it can be said that the technical feasibility of the process has been established.

2.2. PUBLICATIONS

A paper comparing flotation with mechanical cells and columns has been prepared by Apirsa and Trieste University. It has been decided to postpone its publication to allow the introduction of the results of the last pilot plant campaign.

A paper intitled "A Method for a Financially Efficient Design of Cell-based Flotation Circuit by Trieste University and BRGM" has been proposed for publication in "Int. Journal of Mineral Processing" (see addendum 1).
2.3. PROJECT MANAGEMENT ASPECT

2.3.1. Resources for the period

<table>
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<th>% of full corresponding programme* cost</th>
<th>Expenses during the first two years</th>
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* on the basis of the exchange rate of December 28th, 1994, as in column (2).

Such cost statement (with the reserve of the exchange rate which is not corresponding to the full period) shows:

- the low consumption of resources by Apirsa which is directly linked to the interruption during more than a full year of the commercial operation;

- the high consumption of resources by Control International S.A., directly linked to the purchase of the new pilot plant column.

It is mostly probable that the expenses of Apirsa and BRGM during the month of November, in relationship with the pilot plant operation should correct their low consumption.

2.3.2. Management meetings

A project management meeting corresponding to the 2nd year of the project has been held at Trieste University on November 24th 1994.

The next one is scheduled for March 24th at Paris.
REFERENCES


ADDENDUM 1

A METHOD FOR A FINANCIALLY EFFICIENT DESIGN OF CELL-BASED FLOTATION CIRCUITS

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² Process Simulation Group, Research Division, BRGM, Orleans, France
³ Economic Evaluation Group, BRGM, Orleans, France.
SYMBOLS

τ average residence time in a cell unit
α difference in number of cells between two subsequent cleaning stages
c concentrate
C operating cost
E corrective coefficient for the increase in pulp volume due to aeration
f feed
F feed throughput
I capital cost of one flotation cell
K first-order lumped (mass) rate constant; [time⁻¹]
k first-order species-specific rate constant; [time⁻¹]
N number of cells
np years of project life or service
p fraction of metal paid
q metal quotation
Q feed flowrate
r discount rate
R recovery
Rfc refining costs
t tails
Trc smelter treatment costs
u metal deduction units
V volume
Y yield
Z vector of independent variables to optimise
w fraction solid
γ solid specific gravity

SUBSCRIPTS & INDICES

∞ asymptotical variable value approached as time → ∞
g grade
OV overall
PL plant
RG rougher
s cleaning stage
S last cleaning stage
SC scavenger
B bank
C cell unit
SB rougher-scavenger sub-system of a more complex circuit
ABSTRACT

A procedure is derived that allows the selection of the configuration and the sizing of complex cell-based flotation circuits. The method is driven by an user-defined financial objective function that incorporates metallurgical efficiency through concentrate sales revenues and elements of both capital and operating costs. The suggested design method is likely to be superior to the traditional empirical procedures that point to maximise metallurgical performances without including in a formalised way the cost side of the circuit; technically feasible solutions too costly for the metallurgical results they allow are rejected and the best techno-economic solution is iteratively searched. An example application is used to illustrate the innovative method.
INTRODUCTION

Traditionally flotation circuits are sized with the objective of obtaining a threshold metallurgical recovery and grade. The problem is often dealt with the aid of steady state simulators allowing to evaluate alternative flowsheet configurations. Prior work on the algorithms for configuration/optimisation of complex froth flotation circuit has been recently reviewed by Yingling J. (1993). Economic aspects are surely taken into consideration during plant design: as an example, plants based on larger cells cost less and to some extent allow lower operating costs. However the economic side and its technical consequences are not clearly formalised into a decision procedure permitting to rank alternative processing solutions. So that it is not possible to answer questions as (Schena G. 1994)

- is the cost (capital and operating) of an extra cells in a rougher or scavenger bank compensated by the extra recovery it allows?
- is the increase in grade that an extra bank of cleaner allows justified by the related extra costs of the asset?
- which is the best cell-number-size combination from a techno-economical point of view?
- is the extra cost of using smaller flotation cells while maintaining constant the overall bank volume offset by the better metallurgical results related to a restrained “short-circuiting” effect?

Here a framework is set out for a scaling up procedure that allows to derive the configuration of a flotation circuit that is optimal according to both a metallurgical and a financial point of view. Indeed any metallurgical result is obtained at a cost and the two aspects should be weighted one versus the other. The procedure can help to rank processing solutions on the ground of their financial efficiency. A symbolic block diagram showing the succession of the tasks to be completed for an optimum design is reported in Figure 1. The suggested procedure is also demonstrated with an application example.

Engineers can analyse complex processing problems and make better and quicker judgements if they can use more accurate information available through the use of computerised design procedures. These procedures support rather than replace engineer’s judgement and improve the effectiveness not the efficiency of the design process. They simply solve repetitive tasks for which standard and unique solutions exist. Effectiveness is also improved by allowing engineers to perform tasks in less time, with less effort and at a lower cost and improving the quality of the data they use to take final decisions. Several ‘what if’ scenarios can be examined and the final decision can be taken on a trial and error process.
Figure 1. Symbolic diagram of the decision procedure for the selection of the best flotation circuit configuration.
MODELLING FLOTATION METALLURGY

Recovery

The recovery of a flotation bank can be modelled by assuming that

- the change in pulp concentration in a batch flotation process is described by a first-order differential equation
- the flotation rate is constant with respect to time and its value is \( k \)
- the bank of flotation cells is a series of \( N \) perfect mixers
- the flowrate is constant on the cells of the bank that is \( \tau \) is constant throughout the bank
- the recovery tends asymptotically to a maximum value \( R_\infty \)

The continuous-flow recovery equation for a mono-phase model based on these assumptions is:

\[
R = \left[ 1 - \frac{1}{(1 + k \cdot \tau)^N} \right] \cdot R_\infty
\]  

The hypothesis of constant flowrate is not essential if a simulator rather than the simplified Equation 1 is used to calculate the bank recovery. However the use of a bank- rather than a single cell-approach permits to keep the dimension of the mathematical problem within a more reasonable size. The decrease in flowrate can be neglected with confidence when flotation is applied to base metal ores and the volume floated is a little proportion of the feed to the cell.

The overall recovery of a rougher-cleaning system is:

\[
R_{ov} = \frac{R_{RG} \cdot R_{CL}}{R_{RG} \cdot R_{CL} + (1 - R_{RG})} \tag{2}
\]

where \( R_{RG} \) and \( R_{CL} \) are the rougher and cleaning bank recovery.

The overall recovery of a rougher-scavengers system is:

\[
R_{ov} = \frac{R_{RG}}{R_{RG} \cdot R_{SC} + (1 - R_{SC})} \tag{3}
\]

where \( R_{SC} \) is the scavenger bank recovery.

The two relationship can be generalised to the complex rougher-scavenger-S cleaning stages system \((s=1,...,S)\) shown in Figure 2. An approximate overall circuit recovery is:

\[
R_{ov} = \frac{R_{RG}}{R_{SC} \cdot R_{RG} + (1 - R_{SC})} \cdot \prod_{s=1}^{S} \frac{R_{s}}{1 - R_{s-1} \cdot (1 - R_{s})} \tag{4}
\]

where \( R_{s} \) is the recovery of the sth cleaning bank and:

\[
R_{s=0} = R_{SB} \quad \text{if} \; S>0 \quad \text{and} \quad \Pi = 1 \quad \text{if} \; S=0 \tag{5}
\]

\( R_{SB} \) is the overall recovery of the rougher-scavenger sub-system of the complex circuit. Equation 4 can be re-written as:
where:

\[ R_{3} = 0 = R_{SB} \text{ if } S > 0 \quad \text{and} \quad \Pi = 1 \text{ if } S = 0 \] (7)

Rather than as a simple limit of a summation of partial recovery terms the overall recovery relationship can be derived in a more conventional way by writing the solution of a set of material balance equations (Villeneuve J. 1992).

**Yield**

The correspondent circuit yield is:

\[ Y_{OV} = \frac{Y_{RG}}{Y_{SC} \cdot Y_{RG} + (1 - Y_{SC}) \cdot \prod_{s=1}^{S} \frac{R_{s}}{1 - R_{s-1} \cdot (1 - Y_{s})}} \] (8)

where:

\[ R_{s=0} = Y_{SB} \text{ if } S > 0 \quad \text{and} \quad \Pi = 1 \text{ if } S = 0 \]

However the availability of a simulator allows an iterative solution of the circuit and the use the global recovery expressions as those of Equations 2,3 and the approximated Equation 4 can be avoided.

The yield of the single bank of N cells can be expressed as:

\[ Y = \left[ 1 - \frac{1}{(1 + K \cdot \tau)^{N}} \right] \cdot Y_{m} \] (9)

where K is the first-order mass rate constant ; since concentration occurs : k > K or k/K > 1 and R/Y >1. The derivation of K does not require extra experimental work.

**Grades**

The concentrate grade is:

\[ c_{g} = \frac{R_{OV} \cdot f_{g}}{Y_{OV}} \] (10)

where \( f_{g} \) is the circuit feed grade.

The tail grade:

\[ t_{g} = \frac{f_{g} - Y_{OV} \cdot c_{g}}{(1 - Y_{OV})} \] (11)
MODELLING FLOTATION CIRCUIT ECONOMICS

Plant revenues

For a base metal the revenues from concentrate sales are:

\[
\text{Revenues} = Y_{OV} \times p \times (c_g - u) \times (q - Rfc) - Y_{OV} \times Trc \quad \text{$/tonne RoM} \quad (12)
\]

where Rfc and Trc are the refinery and treatment charges, q is the metal quotation, u the grade deductions and p the fraction of metal paid. Values of the relevant costs and metal quotation are published on specialised periodical. The revenues through yield and grade incorporate metallurgical efficiency.

Operating and capital costs

The operating costs of the flotation system can be expressed as sun of the costs related to the rougher, scavenger and cleaner cells are:

\[
\text{Operating cost} = N_{RG} \cdot C_{RG} + N_{SC} \cdot C_{SC} + \sum_{n=1}^{S} N_n \cdot C_s \quad \text{$/year} \quad (13)
\]

The operating costs of a cell in a bank are mainly related to the cost of energy and reagents. For the purpose of including design-specific costs in the procedure objective function only energy related costs are considered. It can be assumed that depending on the type of cells 1.5 - 2.2 kWh per cubic meter of effective cell volume are necessary.

For the size of cells in the rougher stage it is usual practice to assume:

\[
V_{RG} = \frac{Q_{PL} \cdot E}{\tau} \quad (14)
\]

where Q_{PL} is the plant pulp flowrate and E a factor for taking into account the increase in volume due to aeration. Typical values of \( \tau \) are available. For most base metal applications effective average residence time is between 1 and 2 minutes per cell. For the calculation of effective flotation time the volume of flotation mechanism should be detracted from the gross cells volume. Effective flotation time is further reduced by the volume occupied by the air dispersed into the pulp as accounted by trough E.

The cells volume is an independent variable in a green-field scenario. Cells are available when an existing plant should be optimised. Similarly in a case of plant capacity expansion where a number of cells are available. In other situations the cells size for the rougher stage can be easy selected by the approximate calculation indicated in equation 14. The size of the cleaning cells is generally smaller than that of the rougher-scavenger cells. The problem of selecting cell size is further simplified by the fact that manufactures have a limited number of cell sizes available in their production line.

The flotation plant capital cost is:
Capital cost $ = N_{RG} \cdot I_{RG} + N_{SC} \cdot I_{SC} + \sum_{i=1}^{S} N_{i} \cdot I_{i}$

that is the sum of the capital cost of all the cells of the circuit.
The capital cost $I$ of a cell can be evaluated using cost estimation relationships of the type

$I = A \cdot V^{B}$

where $A$ and $B$ are two constants.

The full circuit capital cost can be distributed as a uniform cost throughout the $np$ years of the project life assuming a constant rate of interest $r$ as:

$\text{Equivalent annual cost} = (\text{Capital cost}) \cdot \frac{r \cdot (1 + r)^n}{(1 + r)^n - 1}$

The flotation related cost index per unit of RoM is:

$\text{Costs} = (\text{Operating cost} + \text{Equ.an.cost}) / \text{tonnes RoM per year}$

S/tonne RoM

(18)

The cost index include sole circuit related cost items. The segment of the costs that are or though to be constant with respect to the circuit configuration does not affect the optimisation procedure.

Revenues can be set against cost and the operating margin index derived. The costs item derived from Equations 16-18 should be interpreted as optimisation indices; indeed the actual costs have a more complex structure but the aim here is to isolate the most relevant flotation dependent terms that can be used as variables in the optimisation procedure.

OPTIMAL DESIGN PROCEDURE

Optimisation for complete circuit design

The margin to optimise can be expressed as:

$\text{Margin}(Z) = \text{Revenues}(Z) - \text{Costs}(Z) - \text{Other costs}$

S/tonne RoM

(19)

where the elements of the vector $Z$ are all the adjustable variable of the optimisation procedure. The "other costs" item includes the cost for RoM production and the other processing costs that are not strictly related to the flotation circuit configuration. However this numerical value - constant with respect to circuit configuration - is not required for the solution of the optimisation problem.

Generally:

$\text{dim}(Z) = S_{max} + 3$

(20)

Indeed the optimisation procedure searches for the best number of cells in the rougher and scavenges banks, the best number $S$ of cleaning stages and of cells in each of them. In other terms:
\[
\{Z\} = \{N_{RG}, N_{SC}, S, N_{s=1,S}\}
\]

(21)

By maximising the margin with respect to the independent variables (elements of the vector \(Z\)) the best circuit configuration can be derived.

Constraints

Additional constrains can be used to better the model. For example if smelting and refining installations do not accept concentrates bellow a certain metal grade:

\[
c_g \geq c_{\min}
\]

(22)

Optimisation for sizing the banks

A number of cases can be examined. For example with a number of cleaning stages set \textit{a priori} to \(S\):

Maximise : \(\text{Margin}(Z)\) \quad \text{$/\text{tonne RoM}$}

(23)

\[
dim(Z) = S + 2
\]

(24)

\[
\{Z\} = \{N_{RG}, N_{SC}, N_{s=1,S}\}
\]

(25)

Short cuts

The optimising procedure can be simplified by decreasing the number of independent variables. Generally one cleaning flotation stage has a number of cells lower than the previous stage and higher than the next cleaning stage. If the difference in number of flotation cells between two consecutive cleaning bank is \(\alpha\):

\[
N_s = N_{RG} - \alpha \cdot s
\]

(26)

\(\alpha \in \mathbb{N}\) set of integers.

Hence:

\[
S = \text{floor}(\frac{N_{RG} - 1}{\alpha})
\]

(27)

and

\[
dim(Z) = 3
\]

(28)
\{ Z \} = \{ y_{ro}, N_{sc}, \alpha \} (29)

**Solution algorithm**

The maximisation problem is solved with conventional methods of constrained optimisation. The independent variables of the vector $Z$ are integer. Sequential search techniques that do not require derivatives, allow the inclusion of constraints and tend to find the global maximum are the suitable tools for the application. Also simplified searching techniques that inspect the feasible region of the independent variables and evaluate the value of the objective function replacing as candidate for the solution the point having the highest recorded function value at that point of time by a new point with a higher value are applicable but perhaps more computing intensive.

The proposed algorithm optimises a given network structure, namely the rougher-scavenger-S cleaning stage structure. The _a priori_ selection of the network structure avoids to end up with peculiar unrealistic solutions like those involving 'split steam interconnections'. These arise from optimisation procedures applied to complex unconstrained superstructures (Yingling J. 1993). The unexpected results can be explained by a failure to include in the objective function an element of cost related to the existence of a stream so that the proliferation of interconnection is not penalised. Indeed in real world extra streams have disadvantages; they require pumping, piping and add complexity to the technical solution.
**CIRCUIT MASS BALANCE**

**Bank mass feed-rate**

Once the solution has been found the mass balance can be completed and the solid feed rate $F$ to each flotation stage derived by back ward calculation. For the $S$ flotation stage:

$$F_s = \frac{Y_{OR}}{Y_s}$$ (30)

and for the preceding cleaning stages:

$$F_{s-1} = \frac{F_s}{Y_{s-1}}$$ (31)

and

$$F_{s-2} = \frac{F_{s-1} - (1 - Y_{s}) \cdot F_s}{Y_{s-2}}$$ (32)

The recursive expression is:

$$F_s = \frac{F_{s+1} - (1 - Y_{s+2}) \cdot F_{s+2}}{Y_s}$$ (33)

**Bank volumetric flow-rate**

The volumetric flowrate can also be derived:

$$Q_s = F_s \cdot \left(\frac{1}{\gamma} - \frac{1}{w_s} + 1\right)$$ (34)

where $w_s$ is the solid fraction of solids in the pulp and $\gamma$ the solid specific gravity.

**Cell volume**

The cell volume can be re-calculated:

$$V = \frac{Q_s}{\tau}$$ (35)
POSSIBLE MODEL REFINEMENTS

The procedure can be refined by improving the flotation model. In this work a mono-phase description has been used but the inclusion of a two-phase (that is pulp and froth) model to describe cell recovery can improve flotation description. Cell- rather than bank-based models allow to release the constant cell flowrate assumption. For such a purpose the use of a circuit independent simulator is advisable.

Stage specific rather than general rate constants can also improve the description of the flotation circuit behaviour. In the cleaning stages and in the scavenger section the flotation rate constants can be rather different from the values determined with a batch test on the liberated RoM. Generally $k_{CL} > k_{RG} > k_{SC}$ and $k > K$. However operating conditions as the intensity of aeration in the various flotation stages can determine different k values (Forssberg E.). Reagent addition can also vary the value of flotation rate constants.

Species- and/or size-specific rate constants can further improve the reliability of the model and help to discover possible high circulating load behaviours of some type of particles that can be detrimental for the flotation system and suggest circuit modifications. This refinement requires the rate constant of each considered sub-population.

The inclusion of an element of choice capable to select between one or more bank of cells and one column permits the evaluation of hybrid circuits. Indeed in some cases columns allow to reach in one step better grades than banks of cells in two or more flotation stages but with lower recovery and different costs. Work is in progress to include this capability in the simulation software.

Also the costing procedures can be further ameliorated; as an example including capital costs for flotation circuit equipment as pumps and the related energy costs. For the solution of green field problems solutions specific indirect costs as floor area requirements, building high, instrumentation and control needs can also be included as cost items. The inclusion of penalties for indesirerate metals and prices for valuable sub-products can improve the formula for revenues estimation.

The procedure has a high potential for the improvements of the friendness of general purpose steady state simulation packages (Villeneuve J. 1992). The state of the art simulators simply perform the simulation of flowsheets selected by the design engineers. However the financial based design makes possible a quick and automatic derivation of a default flotation circuit which can be the staring point for later circuit improvements of flotation circuit thus allowing intelligent computer assisted flowsheeting.
APPLICATION EXAMPLE

Capital and operating cost values

The following Table 1. can be used as a guide for preliminary evaluation of flotation cell capital cost. Alternatively the cost of a small size (volume lower than 20 cubic metres) flotation cell can be derived using the cost function \( I_c = A V^B \) where \( A = 5684 \), \( B = 0.6478 \) and the gross volume \( V \) is in cubic meters. The relationship was derived by power regression on available data updated to 1994.

<table>
<thead>
<tr>
<th>effective volume cubic meters</th>
<th>gross volume c.f.</th>
<th>$US</th>
</tr>
</thead>
<tbody>
<tr>
<td>3</td>
<td>100</td>
<td>15000</td>
</tr>
<tr>
<td>4</td>
<td>150</td>
<td>20000</td>
</tr>
<tr>
<td>8</td>
<td>300</td>
<td>25000</td>
</tr>
<tr>
<td>16</td>
<td>600</td>
<td>35000</td>
</tr>
<tr>
<td>38</td>
<td>1400</td>
<td>95000</td>
</tr>
<tr>
<td>100</td>
<td>3700</td>
<td>195000</td>
</tr>
</tbody>
</table>

Table 1. Indicative values of flotation cells capital costs inclusive of motors and drive.

The power regression on the data of Table 1 gives \( A = 6275 \) and \( B = 0.7244 \).

A more precise evaluation of a bank capital cost should include the cost of junction boxes (JB), feed (FB) and discharge (DB) boxes:

\[ I_B = I_c \cdot N_B + N_B / 4 \cdot I_{JB} + I_{FB} + I_{DB} \]

where it was assumed the need of one junction box each 4 cells.

Operating costs can be estimated by taking into account that the power consumption of one large size efficient flotation mechanism can be as low as 1.5 kW \( \text{ca} \) per cubic meter of cell volume including the power required for the air blower. For small size cells a value in the range 1.8-2.2 kW per cubic meter is appropriate. Current (IIfst Quarter 1994) average industrial energy cost in Europe is 6 pence per kWh \( \text{ca} \). that is US $ 9 cents per kWh \( \text{ca} \).

Problem statement

It is required to design a flotation circuit to treat a chalcopyrite porphyry-copper mineral preconcentrated in the concentrator rougher section and re-ground to final liberate the species. The rougher concentrate is 4 percent copper and the overflow of the closed circuit re-grinding mill is of 4000 tonnes per day of dry solid. Batch flotation tests were performed on the liberated ore and repeated on the batch products. The following rate constant values derived by best fitting on mass and grade data collected during the batch tests were used to demonstrate the procedure. The smelter requires minimum 28 percent copper in the flotation concentrate.
Simulation

To illustrate the procedure - rather than performing a full multivariable optimisation leading to the derivation of the complete circuit configuration - the number of cells in the rougher and scavenging section where fixed and the sole number cleaning stages and cells in each cleaning stage were derived by optimisation. This simplification allows to show graphically and step by step the searching procedure. Table 3 summarises some application example specification and simulation assumptions.

Table 2. Values of flotation rate constants and maximum metal recovery.

<table>
<thead>
<tr>
<th>Stage</th>
<th>$k_{CU} \text{ [min}^{-1}\text{]}$</th>
<th>$R_{\infty}$</th>
</tr>
</thead>
<tbody>
<tr>
<td>rougher</td>
<td>0.4</td>
<td>0.98</td>
</tr>
<tr>
<td>scavenger</td>
<td>0.3</td>
<td>0.96</td>
</tr>
<tr>
<td>cleaners</td>
<td>0.5</td>
<td>0.85</td>
</tr>
</tbody>
</table>

Table 3. Design specifications.

<table>
<thead>
<tr>
<th>Design data :</th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Flotation circuit feed rate</td>
<td>4000 Tpd</td>
</tr>
<tr>
<td>feed grade</td>
<td>4 percent</td>
</tr>
<tr>
<td>minimum concentrate grade</td>
<td>28 percent</td>
</tr>
<tr>
<td>average residence time in a cell unit</td>
<td>1.3 minute</td>
</tr>
<tr>
<td>size of available cells</td>
<td>16 cubic m</td>
</tr>
<tr>
<td>number of cells in the rougher bank</td>
<td>9</td>
</tr>
<tr>
<td>number of cells in the scavenger bank</td>
<td>6</td>
</tr>
<tr>
<td>cleaning arrangement</td>
<td>1</td>
</tr>
</tbody>
</table>

Table 4 reports financial and cost data used for the simulation like metal quotations etc.

Table 4. Financial and cost data

<table>
<thead>
<tr>
<th>Description</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>copper metal quotation</td>
<td>2450 $/tonne</td>
</tr>
<tr>
<td>grade unit deduction</td>
<td>1.5 %</td>
</tr>
<tr>
<td>refining costs</td>
<td>$ 200 per tonne metal</td>
</tr>
<tr>
<td>treatment costs</td>
<td>$ 85 per tonne concentrate</td>
</tr>
<tr>
<td>other production costs</td>
<td>$ 6 per tonne RoM</td>
</tr>
<tr>
<td>discount rate</td>
<td>13 %</td>
</tr>
<tr>
<td>project/plant life</td>
<td>10 years</td>
</tr>
<tr>
<td>energy costs</td>
<td>0.09 $/kWh</td>
</tr>
<tr>
<td>cells energy intensity</td>
<td>1.9 kWh per cubic m</td>
</tr>
</tbody>
</table>
Figure 3 shows the increase in grade of the concentrate versus the number of cleaning stages. However only the section of the curve with values higher than the minimum grade required by the smelter is of interest.

Figure 4 shows the decrease in metal recovery as the number of cleaning stages of the flotation circuit increases.

Figure 5 shows that as the recovery decreases also the circuit yield decreases but with a different law.

Figure 6 shows that the revenues from concentrate sales decrease with the number of stages. Here the relevant section refers to grades greater than the minimum paid by the smelter.

Figure 7 shows that the process related costs increase as the circuits becomes more complicated and finally Figure 8 that three is the minimum number of stages to achieve the maximum margin with the constraints posed by the smelter; the number of cells in each cleaning stages is given by Equation 24.

**Optimisation**

The simulation can be repeated with different values of number of cells in the rougher and scavenger banks and scaling down factors (a) for the number of cells in two consecutive flotation cleaning stages. The solution with the highest value of the objective function maximum singles out the overall best plant configuration. The simulation was repeated of a number of circuits arrangements but the overall margin ($47.47) could not be improved. In Table 5 two other solutions (a and b) are reported. Therefore the solution reported in Table 3 is the candidate to be the best.

<table>
<thead>
<tr>
<th></th>
<th>solution a</th>
<th>solution b</th>
</tr>
</thead>
<tbody>
<tr>
<td>number of cells in the rougher bank</td>
<td>9</td>
<td>8</td>
</tr>
<tr>
<td>number of cells in the scavenger bank</td>
<td>6</td>
<td>7</td>
</tr>
<tr>
<td>cleaning arrangement (α value)</td>
<td>2</td>
<td>1</td>
</tr>
<tr>
<td>objective function value</td>
<td>43.7</td>
<td>46.4</td>
</tr>
</tbody>
</table>

**Table 5. Alternative' next best' circuit solutions**

**CONCLUSIONS**

Flotation is the most extensively used process for the concentration of minerals. Today circuits are often designed by the use of empirical methods and rarely studied by the aid of simulators. This statement does not suggest that traditional design procedures lead to inefficient flowsheeting. Indeed process related costs are taken into consideration and the maximisation grades and revenues is the objective of the traditional empirical procedures as well. However these well established methods rest widely on the experience of the process engineers. This calls for the development of appropriate methods for flotation circuit design and optimisation to facilitate and speed up design procedures. The most successful designer is the one that maximises the profitability of the processing operation. This can be achieved by reducing costs and increasing revenues. A method was proposed that includes explicitly the maximisation of the profitability of a flotation plant into design and optimisation procedures. The availability of these procedures in a process simulator can provide quickly the designer with a "default" flotation circuit for the solution of the processing problem and
transform the current simulators that merely document engineer's choices into intelligent tools for computer assisted design. The work of the experienced engineers will still be needed to refine computer generated solutions but facilitated and speeded up. The amount of data required by the model is limited and most of time available anyway to the designer that uses traditional design systems.
ACKNOWLEDGEMENTS

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REFERENCES

Schena G. Flotation-cell size and -bank length calculation according to techno-economic principles. Paper submitted to Aufbereitung Technik, June 1994


Figure captions

Figure 1. Symbolic diagram of the decision procedure for the selection of the best flotation circuit configuration.

Figure 2. Rougher-scavenger-S cleaning stages flotation system.

Figure 3. Concentrate grade as a function of the number of cleaning stages.

Figure 4. Recovery versus number of cleaning stages

Figure 5. Yield versus number of cleaning stages

Figure 6. Revenues from concentrate sales versus number of cleaning stages

Figure 7. Flotation related costs versus number of cleaning stages

Figure 8. Margin versus number of cleaning stages