Stellingen behorende bij het proefschrift van B.J.P.M. Huls

1. Eh-pH diagrams are being used extensively in hydrometallurgy. The hydro-metallurgist knows the conditions to conduct a leach. In the future the mineral dressing engineer needs to adapt to and to apply this concept before designing a flotation experiment.

2. To use mathematical modelling properly and simulation in mineral processing, it takes a person well skilled and trained for many years in this method, combined with accurate measurements of a process in steady state.

3. The recovery in a column flotation cell can be maximized by allowing the pulp interface to vary, depending on the mass of floatable mineral reporting to the column feed.

4. The fastest payback in mineral dressing is process control; automation is required to a degree, but generally only eases the mode of operation.

5. The application of process controll in Mineral Dressing operations lags behind the petrochemical and chemical industries, and even behind pyrometallurgical and hydrometallurgical installations. Only recently have plant operators become aware of the necessity thereof.

6. There is a relation between climate and world conflicts.

7. Passive sport exertion creates hysteria and vehemence.

8. The recognized equation for carrying capacity does not consider buoyancy effects of loaded particles. Carrying capacity is directly related to a function of buoyancy.


9. The greenhouse effect is good to promote tourism on the Russian tundra.
INTERACTION BETWEEN GRINDING AND FLOTATION, SPECIFICALLY COLUMN FLOTATION, AND COLUMN SCALE-UP
INTERACTION BETWEEN GRINDING AND FLOTATION, SPECIFICALLY COLUMN FLOTATION, AND COLUMN SCALE-UP

PROEFSCHRIFT

TER VERKRIJGING VAN DE GRAAD VAN DOCTOR AAN DE TECHNISCHE UNIVERSITEIT DELFT, OP GEZAG VAN DE RECTOR MAGNIFICUS, PROF. DRS. P.A.SCHENCK, IN HET OPENBAAR TE VERDEDIGEN TEN OVERSTAAN VAN EEN COMMISSIE AANGEWEZEN DOOR HET COLLEGE VAN DEKANEN, OP DONDERDAG 6 SEPTEMBER 1990 TE 14.00 UUR

DOOR

BERNARDUS JOSEPHUS PAULUS MARIA HULS
MIJNINGENIEUR
GEBOREN TE LEIDSCHENDAM

druk: wibro dissertatiedrukkerij, helmond.
Dit proefschrift is goedgekeurd door de promotor

Prof. ir. W.P.C. Duyvesteyn PhD
Aan mijn ouders

Aan Gerda

Aan Eldwin en Natalie
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INTRODUCTION

Geologists are inclined to distinguish a mineralization to its origin in principal, and to further classify to mineral combination. Such information undoubtedly is useful for the mining and metallurgical engineers when designing flowsheet details, such as reagents. At this place it is more apt to divide mineral resources into categories that directly indicate which series of sequential unit process steps are being used for upgrading. Broad groupings are ores, rock salts, inorganic raw materials and energy resources. In more detail, one may distinguish

- Base Metal Sulphides
- Industrial Minerals
- Light Metals
- Precious Metals
- Energy Resource Material
- Oxides

Overlap may exist and it can even be argued that 'Iron Ores' deserves its own category. Similarly, someone may favour the group 'Steel Modifiers', which would include for example nickel, chromium, molybdenum, and manganese ores.

For each of the above mentioned groups the concentration process is inherently different. The common steps for Sulphides are crushing, grinding, classification, flotation, dewatering and sometimes magnetic separation. Magnetic separation may be used to separate pyrrhotite from pentlandite, but also refers to the use of high-gradient magnetic separation to remove sphalerite with high iron content from a copper-lead separation circuit. At times gravity separation methods are employed, but mostly to separate large particles of ore from a different category.

Typical beneficiation methods for Industrial Minerals are crushing, grinding, screening, dewatering and drying. This group, however, encompasses a large variety of minerals, some of which demand special treatment. Native sulphur, for example, is recovered from the earth by solution mining; graphite can best be concentrated by column flotation. Unit processes such as magnetic separation (conventional, high intensity, or electro-magnetic) and flotation, are normally included in the flowsheet to eliminate impurities. Minerals of the Light Metals category, if not produced as a byproduct, are mostly limited to crushing, screening, washing, drying and hand sorting or cobbing. Many bauxites are upgraded by washing or wet screening to remove sand and some clay minerals. Metallurgical and chemical grade bauxite that must be transported appreciable distances are often dried before shipment. Precious Metals are recovered by gravity separation, if coarse, or float with base metal sulphides to be separated during a subsequent pyrometallurgical or hydrometallurgical process.
The main method of upgrading employed for Energy Resource Materials are crushing, screening and gravity separation. Fine particles are liberated by grinding and are now often recovered by column flotation. If uranium minerals are to be included in this group, then these differ in treatment as leaching is standard practice. Crushing, screening, washing, and drying also are principal unit process steps for Oxides. Gravity separation and various forms of magnetic separation are employed for chromium and manganese oxides and for rare earths.

Most unit processes are known and practised for many years. Advances have been made in various fields. Agricola, in De Re Metallica, gave a descriptive treatise on ore sorting, crushing, screening and washing as preparation methods for subsequent metal separation. Figure 1 illustrates a crushing and screening operation in the 16th century. The crusher in use is described in the following manner:

"A block of oak timber six feet long, two feet and a palm square, is laid on the ground. In the middle of this is a fixed mortar-box, two feet and six digits long, one foot and six digits deep; the front, which might be called a mouth, lies open; the bottom is covered with a plate of iron, a palm thick and two palms and as many digits wide, each end of which is wedged into the timber with broad wedges, and the front and back part of it are fixed to the timber with iron nails". The crushing itself is executed by stamps and is described as follows: "Each stamp has a tappet at the back, which requires to be daubed with grease on the lower side that it can be raised more easily. For each stamp there are on a cam-shaft, two cams, rounded on the outer end, which alternately raise the stamp, in order that, by its dropping into the mortar, it may with its iron head pound and crush the rock which has been thrown under it. To the cam-shaft is fixed a water-wheel whose buckets are turned by water-power. Instead of doors, the mouth of the mortar has a board, which is fitted into notches cut out of the front of the block. This board can be raised, in order that when the mouth is open, the workman can remove with the shovel the fine sand, and likewise the coarse sand and broken rock, into which the rocks have been crushed; this board can be lowered, so that the mouth thus being closed, the fresh rock thrown in may be crushed with the iron-shod stamps".

The crushing action was fulfilled in a circuit closed by screens, as becomes apparent from the following paragraph:

"Broken rocks or stone, or the coarse or fine sand, are thrown by the workman into a box, which is open on the top and the front, and is three feet long and nearly a foot and a half wide. Its sides are sloping and made of planks, but its bottom is made of iron wire netting, and fastened with wire or two iron rods, which are fixed to the two side planks. This bottom has openings, through which broken rock of the size of a hazel nut cannot pass; the pieces which are too large to pass through are removed by the workman, who again places them under stamps, while those which have passed through, together with the coarse and fine sand, he collects in a large vessel and keeps for the washing".
Fig. 1. Crushing and screening in the 16th century. (From Agricola's De Re Metallica)

It would be an onerous job for writer and reader to analyze in detail every progress made in the improvement of equipment to separate minerals more effectively and economically. This manuscript, therefore, will focus on changes in philosophy to operate unit processes, specifically in base metal sulphide plants. These philosophies do not have to be restricted to one mineral category alone, but are believed to be equally applicable to other categories. Furthermore, it has as objective to emphasize the importance of process control, a modern concept in the mineral industry, not necessarily modern anywhere in the chemical or
petrochemical industries. Process control was made possible especially following the spell of low prices of metals in the late seventies, early eighties. At that time management was forced to look at more economic alternatives and cost cutting was partially realized by personnel lay-off. Plants could only run, however, if a combination of automation and process control would take the place of or assist plant floor operators. It should be borne in mind that first attempts of process control were made in the late sixties; only since about 1978 was instrumentation improved to such degree that process control became successful.

Process control will be discussed further in Chapter 1.4. It pertains to maintaining operating parameters within predefined limits all the time. This cannot be achieved without constant and consistent monitoring a circuit, while comparing process variables to set points. A deviation would result in a corrective action the magnitude of which depends on the size of deviation.

In Figure 2 in Chapter 2.1, one may compare a rudimentary PID for a grinding circuit that incorporates sufficient equipment to operate the circuit. Not shown are the high and low temperature and pressure alarms or other equipment that could detect possible malfunctioning of the hardware involved. Here, the attendant decides on how the circuit runs. Figure 9 of the same chapter shows a PID of the same circuit, but includes more process monitoring equipment. A particle size set point is dialled in, and the control strategy ensures that size is maintained through automatically varying the throughput of the circuit. A different loop maintains the level in the pump box. Not shown are the control loops dealing with constraints such as mill power overload, high pump box level, high cyclone underflow density, high pump amperage and varying pump box level. When a constraint is reached, corresponding digital alarm signals suspend the size control strategy. Automatically, and gradually, operating parameters are being altered without operator interference. This method ensures continuing equilibrium in the circuit, even though constraints are reached. That is what makes Figure 9 different from Figure 2 where necessary human interruption invariably is late in response, which results in circuit imbalance. In this case the circuit is not controlled (properly).
PART ONE

1.1 Mineral Processing: Important Process Parameters

Mineral Processing is the first step in obtaining final products or concentrates from natural minerals or ores that need to meet specific requirements as to grade and/or particle size [1]. The grain size of run of mine ore may vary from 0 to 1000 mm, the valuable components often are small in mass or volume, and the ore usually contains one or more minerals that are intimately intergrown. To meet product specifications generally two technical problems need to be resolved:

1) minerals need to be concentrated to comply with a grade suitable for subsequent processing. This implies separation of useless, sometimes detrimental components. The process becomes more complicated if more than one valuable mineral has to be concentrated separately.
2) minerals need to be concentrated to particle size. Size ranges may be narrow depending on the application of the product.

Parameters that determine the final lay-out of a process flowsheet may be distinguished into two main influence spheres: those that are imposed upon by external factors, and those that are inherent to the very ore body. To the first group belong general type of ore body, mine method, geography, environment, client, and overall economics. The second group is characterized by the occurrence of minerals in a deposit in relation to each other, to size, and to value, that governs the selection of equipment. Factors associated with each influence sphere are expounded below.

External factors that influence a process lay-out

Traditionally, the determining factors to exploit an ore deposit were referred to in brief as the four M's: Mineral, Market, Money, and Management [2]. It is beyond the purpose of the current discussion, however, to include such important influences as law, management, financing, and politics. The external factors under consideration here purport directly to the physical appearance of a processing plant that enables upgrading of an ore to a desired product.

The form in which an element occurs in nature implies a certain concentration procedure. A most famous nickel find is the massive nickel-copper sulphide deposit along the base of the Sudbury (Ontario, Canada) layered mafic intrusive. The average nickel content is around 1.1 to 1.8 %. Beneficiating techniques such as flotation and magnetic separation play key roles in preparing the ore for subsequent extractive metallurgy. Tropical weathering of ultramafic intrusives results in lateritic garnierite deposits (e.g. Dominican Republic) [3]. Although such ore assays roughly equivalent in nickel (1.6 %), it is not amenable to most of the standard beneficiation methods. This is due to the manner in which its nickel content is chemically disseminated. Screening, rubbing, and hammer milling to reject oversized, less weathered fragments is the extent of physically concentrating nickel in laterites [4].
Ore bodies may exhibit an enriched cap formed by weathering or redissolution with primary ore deeper in the ground. Each zone contains different minerals, which could alter drastically the size or lay-out of a concentrator (Ok-Tedi, Papua New Guinea), or at least could prompt a change in the reagent scheme (Cuajone, Peru).

Processing methods sometimes depend on the methods used in the mine. Very low grade ores are extracted underground by block-caving, a method not selective per se. But large quantities demand larger and more process equipment. Inherent selective underground mining methods are cut-and-fill and square-set mining of vein deposits. Ore withdrawal usually is controlled to grade. Shrinkage stoping often implied long-time exposure of ore to the atmosphere. Especially pyrites would oxidize at the expense of flotation results. In coal the trend is continuous mining, but more dust is generated compared to more manual winning methods. The obvious way to control dust is by water sprays, which makes the cleaning of coal with air nearly impossible. Wet cleaning of coal is less selective because of fine coal adherence to the coarse product, and requires a subsequent dewatering stage. This in turn is impeded by the higher fines content. The industry will be inclined to increase cyclone usage and adopt column flotation as the final coal cleaning stage.

Geography may also have an impact on the selection of the concentration process. The terrain affects pumping, snowfall defines plant compactness (some concentrators even go underground as did Andina, Chile), water imposes constraints on equipment protective material and its temperature affects process metallurgy, remoteness limits the choice of equipment in terms of size and manufacturer, and manpower adaptability may restrict the degree of automation and control [6]. The Lupin milling circuit in the Canadian Arctic was determined by the size and load that a Hercules airplane could carry. Polaris and Red Dog, both also in the North, were built in sections that were barged up to their respective location. Space was at premium which limited the selection of equipment. At Red Dog the additional constraint is water: the supply is limited and no effluence is allowed. Column cells in the cleaning circuit feature launders angled at 60° to allow gravitational flow of high density concentrate directly to the pressure filters.

Not only the ore grade and the economic value of the mineral may help in the process choice, but also the vicinity of a smelter. A nearby smelter may allow low grade concentrate production to boost recovery [8]. Assume that a next door copper smelter would accept 50,000 tonnes of copper concentrate containing 25 % copper. To ship this concentrate over a distance of 960 km it must be upgraded to 27 % copper, which is at the expense of 0.8 % in recovery, given a known grade-recovery relationship. If a company needs to transport copper concentrate over a distance of 2500 km, the drop in metal recovery for a 30 % grade concentrate would fall by $ 2.08 MM. However, savings in transportation costs would amount to (2500 * 2061 =) $ 5.15 MM, under the assumptions stated in the Table below.

Capital and operating costs escalate rapidly in the order gravity concentration of alluvial deposits (e.g. dredging, sluicing), gravity concentration of lode ores, flotation, simple hydrometallurgical processes and complicated hydrometallurgical and pyrometallurgical processes [9]. This sequence has changed putting simple hydrometallurgical processes ahead
<table>
<thead>
<tr>
<th>GRADE</th>
<th>REC</th>
<th>TONNES</th>
<th>TONNES</th>
<th>TRUCKING</th>
<th>METAL REV.($/MM)</th>
<th>MAX DIST km</th>
</tr>
</thead>
<tbody>
<tr>
<td>% Cu</td>
<td>% Cu</td>
<td>DRY</td>
<td>8% moist</td>
<td>$/km</td>
<td></td>
<td>千米</td>
</tr>
<tr>
<td>25</td>
<td>92.8</td>
<td>50,000</td>
<td>54,000</td>
<td>2700</td>
<td>24.75</td>
<td>NEXT DOOR</td>
</tr>
<tr>
<td>27</td>
<td>92</td>
<td>45,897</td>
<td>49,569</td>
<td>2478</td>
<td>24.54</td>
<td>960</td>
</tr>
<tr>
<td>30</td>
<td>85</td>
<td>38,165</td>
<td>41,218</td>
<td>2061</td>
<td>22.67</td>
<td>3255</td>
</tr>
</tbody>
</table>

TRUCKING:$0.05/tonne/km, not incl. handling charges
REVENUE: assume total recovery in handling, smelting and refining is 90%; copper @ $1/lb; not incl. smelter charges.

of flotation, due to recent advances made in heap leaching-electrowinning of especially copper ores. Nevertheless, the point being made is to concentrate valuable mineral as much and as soon as possible to minimize the throughput of subsequent, more expensive unit processes. A thorough evaluation of the ore in the initial stages of the development of an ore body is absolutely necessary to scope the range of equipment to be installed. Within each ore processing category the range of process sophistication is wide and hence the capital and operating costs.

The client may dictate the type of unit operations, particularly in the case of industrial minerals. Exotic minerals are sold to specific buyers and processing techniques may have to be tailored to meet specifications imposed by the buyer. A summary of the influence of product specifications on process techniques was given by Thom [7]. When producing precious stones, for example, care should be taken not to crush these unnecessarily small; mica must be produced in large, flawless sheets; asbestos was produced in long fibres.

To build new concentrators the engineers have to be very aware of the disposal of tailings and other effluents. Environmental considerations may dictate the selection or number of unit processes. When building a concentrator engineers have to incorporate adequate disposal plans of tailings and other effluents. Gold plants, for example, use cyanide, but tailings effluents high in cyanide and heavy metal concentration require treatment. Before treatment with ferrous sulphate, Hemlo Gold (Ontario, Canada) recycles effluents to the grinding circuit first where cyanide and heavy metals are partially removed. Even existing plants have to adapt their flowsheet to accommodate new legislature on environment. The 'Sulphur-dioxide Abatement Program’ in Ontario, Canada requires two major nickel producers to review and alter their flowsheets in order to increase the rejection of pyrrhotite, an iron sulphide low in nickel (0.6%). Each company will need to be in compliance by 1994.

Factors inherent to the very ore body

As has been alluded to, upgrading can be established in various ways. Methods basically are distinguished between physical and chemical processes. Physical properties of a mineral often lead directly to a process for separating that mineral from others. The magnetic
susceptibility of monoclinic pyrrhotite allows effective removal of this mineral from pentlandite by magnetic separation; gold, with a specific gravity of 19.3 can be recovered by spirals, or other gravity separation techniques, away from quartz (s.g. of 2.65) or other sulphides (s.g. below 8); coal with a specific gravity of about 1.3 can be floated from heavier ash-containing impurities by immersing the material in a heavy medium. The ash-forming constituents, being heavier than the coal substance, increase the density of coal and can be removed as a refuse sink by selecting a suitable specific gravity of the separating fluid. Physical-chemical reactions taking place in surface chemistry (flotation, electrostatic fields) are considered a physical process. Large differences in physical properties are required, which sometimes are aided by modifying agents (reagents, electric field, heavy medium).

Chemical processes, such as leaching and roasting, are resorted to when physical processes become uneconomic or are impossible. In chemical processes, the mineral structure does not change just on the surface, as occurs with flotation, but throughout. Examples are the dissolution of gold and silver in cyanide, and the removal of sulphur from sulphides during roasting. Nickel laterites may be upgraded by discarding hard waste rock, but the low nickel content in the weathered serpentine limits the concentrate grade to about 3%. Only by a change in composition via roasting or leaching is winning of nickel possible.

For physical separation the valuable material must be liberated from host rock or other components. This process depends on how the ore deposit was formed. One of the oldest classifications is based on the form assumed by the mineral deposit [3]: veins, replacements and disseminations. A copper sulphide may be porphyry (disseminated), a stratiform deposit, from hydrothermal origin (veins), or a massive sulphide. When of hydrothermal origin, the ore body usually contains a complexity of fine-grained minerals. Examples are the copper-zinc-lead-silver mines in Butte, Montana and at Cerro de Pasco in Peru. Finer grind is required and more final products are produced. A typical copper porphyry in the Cordillera of North and South America is simpler in concept and less equipment is involved. Figure 2 depicts examples of flowsheets for simple and complex ores.

Exposure to weathering is a common phenomenon. Copper minerals may be present in various stages of transition, from chalcopyrite (CuFeS₂) to bornite (Cu₉FeS₈) and to chalcocite (Cu₂S). Even cuprite (Cu₂O) and native copper are not uncommon. A different array of minerals in a single ore body at least exacts a different reagent regime. Chalcopyrite floats well with dithiophosphate, and chalcopyrite-pyrite middlings are boosted by a fairly selective iso-propyl xanthate. Little native copper and copper oxides will be recovered unless stage additions of long-chained xanthates, such as amyl xanthate, are selected.

It has often proven to be an impossible task for the mine to provide the mill with a homogeneous feed in terms of mineralogy, assay, hardness, size distribution, clay content, degree of oxidation and moisture content. Within a single mine different types of ore frequently occur, each with a different metallurgical response and requiring dissimilar reagent dosage. At Palabora, South Africa, two distinct ore types from the same open pit demand separate treatment to generate different byproducts [10]. An attempt to correlate ore types to recovery was given by Satchwell [11]. Major ore types were defined as lattice
Fig. 2. Flowsheets for simple and complex ores. (A) Ok Tedi flowsheet for a copper-porphyry; (B) Sullivan flowsheet for the lead and zinc flotation circuits.
porphyry, basaltic andesite, intrusive andesite, rhyolite porphyry, and the oxides zone. The alteration of clay was very intense and further distinction was made to degree of alteration. Examples are phyllic, phyllic-propylitic, silicified phyllic, argyllic-phyllic etcetera. Mine geologists realized that such identifications had little meaning for the mill man. Ore became identified by hardness and clay content and as such introduced on a new mine map. The fallacy of the new categorization was that naming was realized on visual inspection only. A supposedly hard ore type actually allowed higher production due to its brittleness. Samples of each of the ten major ore types, therefore, were later submitted to Allis-Chalmers where the crushability and grindability work indices could be determined. Table 1 summarizes how the various ore types were classified before using a relative hardness scale. Actual work indices have also been entered in the Table, and discrepancies are apparent. Following bench-scale grinding, comparing each ore type to a 'normal' ore, and mine testing with a Schmidt compression test, the two departments subsequently agreed to a new relative hardness index, which was more suitable to the mill. Coupled with the original geological classification the combination of the data formed the basis of a final revision of the mine map. Now the concentrator could quickly react to the type of ore expected that day.

<table>
<thead>
<tr>
<th>SAMPLE DESCRIPTION</th>
<th>ORIGINAL CLASSIF.</th>
<th>A-C WORK INDEX</th>
<th>NEW SCHMIDT CL.</th>
<th>COMPR.</th>
</tr>
</thead>
<tbody>
<tr>
<td>1 LP-FIL</td>
<td>H</td>
<td>6.6</td>
<td>S</td>
<td>321</td>
</tr>
<tr>
<td>2 LP-FIL/SIL</td>
<td>H</td>
<td>11.3</td>
<td>H</td>
<td>488</td>
</tr>
<tr>
<td>3 LP-SIL(90)%</td>
<td>VH</td>
<td>12.5</td>
<td>VH</td>
<td>504</td>
</tr>
<tr>
<td>4 IA-PROP/SIL</td>
<td>H</td>
<td>14.1</td>
<td>VH</td>
<td>479</td>
</tr>
<tr>
<td>5 IA-FIL/SIL</td>
<td>H</td>
<td>11.3</td>
<td>H</td>
<td>406</td>
</tr>
<tr>
<td>6 BA-(FRESH)</td>
<td>-</td>
<td>13.6</td>
<td>VH</td>
<td>465</td>
</tr>
<tr>
<td>7 RP-FIL</td>
<td>VH</td>
<td>12.9</td>
<td>VH</td>
<td>509</td>
</tr>
<tr>
<td>8 BRECCIA-FL/SIL</td>
<td>VH</td>
<td>8.7</td>
<td>S</td>
<td>-</td>
</tr>
<tr>
<td>9 BA-ARG</td>
<td>HC</td>
<td>4.8</td>
<td>S</td>
<td>-</td>
</tr>
<tr>
<td>10 IA-ARG</td>
<td>HC-MC</td>
<td>5.9</td>
<td>S</td>
<td>-</td>
</tr>
</tbody>
</table>

| LP-FIL             | LATITE PORPHYRY - PHYLIC ALTERATION |
| LP-FIL/SIL         | SAME, BUT SOME SILIFICATION         |
| LP-SIL(90)%        | LATITE PORPHYRY - INTENSE SILIFICATION |
| IA-PROP/SIL        | INTRUSIVE ANDESITE - PROPYLITIC ALTERATION, SOME SILIFICATION |
| IA-FIL/SIL         | INTRUSIVE ANDESITE - PHYLIC ALTERATION AND SOME SILIFICATION |
| BA-(FRESH)         | BASALTIC ANDESITE                    |
| RP-FIL             | RHYOLITE PORPHYRY - PHYLIC ALTERATION |
| BA-ARG             | BASALTIC ANDESITE - ARGLYLIC ALTERATION |
| IA-ARG             | INTRUSIVE ANDESITE - ARGLYLIC ALTERATION |
| HC-MC              | HIGH TO MODERATE CLAY                |

Table 1. Classification by hardness of the various types of ore in the Cuajone ore body.
The first step in the upgrading process usually is comminution, i.e. crushing and grinding. The ratio of size reduction needed between feed and finished product determines equipment type and method. Other factors to play a role are moisture and top feed size [12]. Typical distinctions wielded in comminution are large, intermediate, and small equipment, and application for hard, intermediate and soft material, and for wet or dry material [1]. Hard ore is crushed in a jaw crusher; a smooth roll crusher is applied to medium, hard and abrasive ore having a feed size a factor of 10 smaller than that for a jaw crusher. The resulting product also is a factor of 10 smaller. Likewise, a rod mill is used where overgrinding must be avoided or to produce a coarser grained product, whereas a ball mill is employed to obtain finer grinds. Some ball mills discharge through a grate to produce fine products from minerals that display a tendency to overgrind due to a high specific density (galena). Presently, economics dictate increase usage of autogenous mills, which, however, can not be applied indiscriminately. Tower mills are in vogue for a variety of reasons, ranging from lower installation costs to higher efficiency [13].

The size of grinding media is directly proportional to the work index and feed size distribution of the ore, and inversely proportional to the kinetic energy of the mill media at the time of impact [14]. For each ore, therefore, a top ball size and ball size distribution must be calculated. At Cuajone the mills were charged with a 60/40 mixture of balls measuring 51 and 76 mm in diameter respectively. A calculation showed that a charge composed of 30% balls of 89 mm, 42% of 51 mm, and 28% of 76 mm would be preferable. Indeed the throughput was increased by about 4.5 tonnes per hour per mill or 1.6%. Bond [23] derived a similar relationship to calculate the top ball size. For mill feed containing significant fractions of 5 to 50 mm material, however, a top size of 160 mm may be the calculated answer. In milling practice it is quite unusual to use such large balls as these may cause liner damage.

Steps subsequent to comminution are chosen to make optimum use of the differences in physical characteristics of minerals and host rock. If liberation can be attained at coarse size fractions, gravity separation may be considered. This is applicable where the difference in specific gravity is at least 1.5 and the size range is narrow. Heavy medium separation works on particles larger than 2.4 mm, but as the liquid is expensive practice is limited to applications at a spectrum of lower density. The smallest particles must be at minimum 20 to 30 microns in size. Separation by susceptibility to contact polarization in an electrostatic field is presently reserved for specialty applications such as the removal of rutile from zirconium or vermiculite from gangue. Magnetic separation is ubiquitous and can be employed for fine ground material. Froth flotation is common for complex and low grade ores.

It was mentioned earlier that the complexity of ore, in terms of mineral content, dictates the number of unit processes. Moreover, if minerals are sufficiently intergrown that regrind becomes necessary, then more cleaning capacity should be allowed for. On the other hand, some concentrator flowsheet designs may be too complex and could be simplified. A good understanding of the chemistry and process may allow elimination of some unit process
steps. For example, the middlings thickener collecting rougher concentrate at various porphyry copper operations (Cuajone, Mission in Arizona, Balkash in Russia) may be superfluous or even detrimental. At Cuajone conditioning to pH of about 11.5 took place in the thickener to depress pyrite. Rougher concentrate was routed directly to the cleaner circuit after conditioning at that pH for such a long time was found to partially depress native copper and molybdenite.

Dewatering normally is the last stage in the concentrating process. The destination of the final product prescribes the method or equipment to be selected. A concentrate to be shipped overseas or to a custom smelter has imposed upon strict guidelines of moisture content. Recently, a drum filter followed by a dryer is more and more being replaced by a pressure filter. Other minerals are dried to similar moisture content by belt filters. Thickeners precede filters, but are also used within the flowsheet to increase pulp density whenever required. Centrifuges dewater solids of low slime content. Screens are seen in coarse solid applications. Cyclones are of service in dewatering tailings, apart from their usual place within a grinding circuit.

The interaction between Grinding and Flotation.

To effect adequate separation minerals need to be liberated. If the concentration process selected were based on separation by gravity, then it would be desired to proceed with size reduction to the point that mineral grains are entirely free. It is practically impossible to achieve this, and unnecessary, because a more suitable concentration process such as flotation could be chosen. For adequate response in flotation the extent of area exposure of the floatable mineral is important, not its volume content in the particle. Mineral grains fifty microns in size do not require to be ground to that dimension. Smaller sized grains would ensue that could not even be recovered by flotation. Also fines would impede downstream process steps such as filtration.

It is desirable to employ multiple stages of grinding and flotation. Mineral separation can then be effected as coarse as possible, minimizing overgrinding and costs; middlings streams can be prepared for subsequent re-treatment, and the overall selectivity can be enhanced. In addition, the individual recovery of different valuable minerals is optimized, because each may respond best in a certain size range [15]. Liberation has thus been a topic of much interest. Recent models extend population balance size reduction models to include two minerals with an arbitrary number of locked particle fractions [16]. The major reservations, held against these models, concern the volumetric estimations of grade distributions from linear grade measurements using stereological techniques. Therefore, predictions made from this model must be, and were, tested against pilot plant data. Another, more simplistic approach is to assume that the position, shape and size of a fragment of an original grain is independent of its position in the uncommingled ore [17]. This theory leads to the simple conclusion that, starting with a material in which 60 % of the sulphide is liberated, grinding this material twice as fine would only liberate half of the remaining unliberated portion (20 %). Energy consumed in grinding is roughly proportional to the new surface area created: the limited increase in liberation has been attained at high energy costs. Again, the point is being made to separate particles as coarse as possible.
Liberation may be relevant but the mineral distribution-by-size should be combined with recovery-by-size curves. A good illustration is given in Figure 3 [18]. Table 2 summarizes some relevant data from Figure 3. It shows that the plant size-by-size recovery curves basically remained unchanged despite of wide fluctuations of feed size and rate.

![Graphs showing % Cu recovery for Coarse, Standard, and Fine Flotation Feeds.](image)

**Fig.3.** Rougher bank recovery performance as a function of particle size and frother type.

<table>
<thead>
<tr>
<th>FEED</th>
<th>GRIND SIZE</th>
<th>THROUGHPUT</th>
<th>OPTIMUM FLOTATION RANGE (ie CuRec&gt;95%)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>$K_{80}$ (µm)</td>
<td>(mtph)</td>
<td>MIBC (µm)</td>
</tr>
<tr>
<td>COARSE</td>
<td>130</td>
<td>95</td>
<td>25-40</td>
</tr>
<tr>
<td>STANDARD</td>
<td>100</td>
<td>75</td>
<td>24-45</td>
</tr>
<tr>
<td>FINE</td>
<td>74</td>
<td>53</td>
<td>30-40</td>
</tr>
</tbody>
</table>

**Table 2.** Optimum flotation ranges for two frother types as a function of feed size distribution [18]. The data in this Table were taken from Fig.3.
This Figure also manifests that coarser particles may be floated by selecting an appropriate frother type, which may possibly allow a slightly coarser grind.

No models take into account that, when the model is needed, the ore body has been found and the mineral content, playing an important role, should be introduced as a fixed parameter. Another shortcoming of liberation models is that these do not account for preferential detachment along grain boundaries to which the nature of a comminution process may contribute [15]. Preferential grinding occurs at the Strathcona mill of Falconbridge Limited (Central Ontario, Canada) as Figure 4 attests. The curves were obtained by classifying an overflow sample by size fraction and have each fraction analyzed on metal content.

Evidence of preferential grinding was also noted elsewhere [19]. When analyzing the breakage of a base metal ore from the Bathurst area in New Brunswick, Canada, the mode of breakage was found to be strongly influenced by the behaviour of pyrite. Initial grinding breaks pyrite along grain boundaries and the process continues until most of the poorly to moderately bounded pyrite grains have been separated. Further grinding breaks pyrite across grain boundaries (random breakage) and chips edges of particles.

Liberation requirements and recovery by size can be coupled by the term 'Release Analysis' [20]. It is a technique that separates a ground ore sample into a number of fractions perfectly by flotation, so that the degree of liberation or 'state of release' of the sample may be seen. Consider it analogous to the float-and-sink analysis in dense medium separation operations. With a simple curve-fitting technique the liberation of an ore can be modelled and plotted as a function of a grinding parameter, for example the overflow particle size. A knowledge of metallurgical response is required as a mineralogical investigation does not allow accurate measurements on fine particles and does not provide information on the flotation behaviour of locked particles. Selective breakage effects and other relevant ore texture parameters cannot be quantified by microscopy alone.

In secondary ball milling, usage of the proper ball size is important to effect adequate liberation. At least at two concentrators, that I am aware of, has a smaller ball diameter resulted in a finer product size. Mt. Lyell in Australia reverted to a 30 mm diameter ball, from 50 mm; a steady-state mathematical model proved to them that the recovery in the subsequent cleaning stage was improved by 2% as a direct result of the finer grind [21]. Cuajone, Peru, hoped to attain a better gangue-chalcopyrite separation from a finer grind, and started to use a 25 mm diameter ball in place of 38 mm. After four months, the regrind product of the test mill was reduced in size by a relative 4 %, compared to standard operation. At the time, insufficient data was available to evaluate whether the objectives were met. The decision was taken, however, to proceed with the smaller ball in all regrind mills.

Even for secondary mills overgrinding should be watched for. Table 3 below summarizes laboratory test work results with respect to the recovery of slime pentlandite liberated in a regrind step. Here, the magnetic fraction of a scavenger concentrate is reground to improve pyrrhotite-pentlandite separation in a subsequent flotation step. Currently,
Fig. 4. Preferential grinding at Strathcona Mill, Falconbridge Limited.

<table>
<thead>
<tr>
<th>Recovery</th>
<th>MAG SEP</th>
<th>FLOTATION</th>
</tr>
</thead>
<tbody>
<tr>
<td>-10 micron pentlandite</td>
<td>&gt;86 %</td>
<td>&lt;79.2 %</td>
</tr>
<tr>
<td>10 to 38 micron pentlandite</td>
<td>&lt;71.1 %</td>
<td>&gt;80.4 %</td>
</tr>
</tbody>
</table>

Table 3. The recovery of fine pentlandite by magnetic separation and by flotation.
Falconbridge ponders the addition of magnetic separation after regrind to recover slime pentlandite in a non-magnetic fraction rather than have it end up in the flotation tailings.

Throughput and recovery can be correlated but differs for each ore type. A typical example is given in Figure 5. A grinding circuit controlled on size was sampled once equilibrium was reached. Several flotation tests were conducted on each overflow sample.

![Throughput - Recovery Relationship (Cuajone)](image)

It is not just liberation and recovery by size flotation response that result in optimal metallurgy. Too many concentrators assume their grinding circuit is stable. Grinding stability will further be discussed in Chapter 1.4. It suffices here to demonstrate effects of instability in the grinding circuit on flotation results, although variable performance in the flotation circuit does not solely depend on grinding performance. Figure 6 depicts fluctuations in overflow density and particle size noted for a typical practice in grinding operation: throughput is set and overflow density is supposedly regulated by pump box water addition. No level control on the pump box is apparent. Figure 7 shows daily variations in plant performance.
Fig. 6. Grinding Circuit Operation - Tonnage Control

1.2 Grinding and Classification

1.2.1. Fundamentals of Grinding

The purpose of size reduction is three-fold: to liberate minerals for subsequent processes, to increase the surface area available for chemical reaction, or to produce mineral particles of required size and shape.

Batch grinding a sample in a laboratory mill produces a distinct pattern of curves when plotting the cumulative percent undersize against the particle size on a log-log scale. Up to a certain size limit the lines are parallel. Furthermore, a small change in size (dx) was found to be proportional to the energy expended per unit weight of particles (dE), and the energy required to bring about the same relative change, was found to be inversely proportional to some function of the initial particle size (x). The relationship between energy and breakage may be expressed as [22]:

$$dE = -K \cdot \frac{dx}{x^n}$$

where K is a constant and n an exponent to indicate the degree by which the energy expended is a function of the original particle size.
Various workers have interpreted this relationship differently and established a series of empirical laws.

Modern day grinding circuits are being designed from scale-up simulation programs, which have their input from laboratory bench-scale test results. Simulation techniques and modelling find their roots in the old empirical grinding laws, which were mostly formed even before the concept of fracture mechanics was born. Modifications to the first two laws were made between 1940 and 1960, but the rules were empirical yet. Some attempts had been made towards simulation. The advent of the computer, however, made it possible to solve the difficult mathematical equations.
The old grinding laws

Estimations of the effects on the grinding time required and on the specific energy of grinding (energy per unit mass) when the specifications of feed and product sizes changed, were derived from empirical laws. These predict it takes longer (and more specific energy) to grind a given mass of coarse feed to a fine product than a less coarse feed to a less fine product. The general expression of these laws can also be written in the form [23]:

$$F_r = K_r \cdot \text{(new surface produced per unit mass)} \quad \text{(2)}$$

where $F_r$ is the specific grinding energy and $K_r$ is often used as an index for grinding efficiency. Rittinger interpreted this as the work to break a cube into smaller cubes is proportional to the newly created surface. Communion, however, can only take place by supply of elastic energy, which always goes into volume, never into just the surface [24]. Bonds that previously existed across the new surfaces formed require a minimum energy. The pseudo-Rittinger equation hints at the realization that only a very small portion of the total energy supplied to milling is applied to particle surface energy. It establishes a correlation between mill energy input and feed and product sizes. Table 4 summarizes the various equations alluded to in the text. Kick rightfully claimed that the absorbed work is proportional to the volume when geometrically similar, and otherwise physically identical, bodies are similarly deformed. However, his claims that strength is independent of particle size and that fractures progress similarly under such conditions are unfounded. As particles become smaller their strength increase due to lower flaw density. The similarity in a fracture pattern is disputed by the nonlinear behaviour of the "energy investment", i.e. the volume-related energy over a distance, with the volume related specific surface over that distance. The further the crack propagates, the more invested energy converts to kinetic energy [24]. On the same basis as for Rittinger can the Kick law be reformulated to the pseudo-Kick equation (see Table 4).

Bond formulated the third law and has modified the pseudo-Rittinger equation and related the mill power to the square roots of the 80%-passing sizes of feed and product. He maintained that such inclusion conformed with the Griffith crack propagation theory. Griffith considered stress considerations factors in the regions of the tips of the flaws as first. He showed that cracks originate from stress. These propagate upon release of energy by relaxation of strain in the surrounding material if it is larger than the energy required to create new surface. This law required correction as it neglected plastic deformation energy, crack geometry, transformation induced plasticity and crack deflection due to crystal structure [25].

Charles took as basis the Schuhmann size distribution, a log-log plot of cumulative weight percent undersize $P(x)$ versus size $x$. The point he made was that in the region of parallel straight lines the size $k_x$, at which $P(x) = 1$, may vary with specific grinding energy. This is illustrated in Figure 8. A slight modification of the $\alpha$ value, suggested as a result of an empirical experimental observation, produced a slightly better fit [23]. The Rosin-Rammler equation is directly derived from the grinding data by plotting results on Rosin-Rammler paper. Another concept is the Zero Order Rule, best explained from a plot of the rate of production of fine material against grinding time. Up to a protracted period of grinding the
TOTAL NET ENERGIES (KWh/t) FROM INFINITE FEED SIZE TO A PRODUCT OF SIZE $x$ ARE AS FOLLOWS:

- **Rittinger**
  \[ E = C \cdot \frac{1}{x} \]

- **Kick**
  \[ E = -C \cdot \ln(x/\infty) \]

- **Bond**
  \[ E = 2C \cdot \frac{1}{(x^{1/2})} \]

CORRESPONDING NET ENERGIES REQUIRED TO REDUCE A FEED $x_f$ TO A PRODUCT $x_p$ ARE AS FOLLOWS:

- **Rittinger**
  \[ E_r = C_r \cdot \left( \frac{1}{x_p} - \frac{1}{x_f} \right) \]  
  (pseudo-Rittinger)

- **Kick**
  \[ E_k = C_k \cdot \left[ \ln \left( \frac{1}{x_p} \right) - \ln \left( \frac{1}{x_f} \right) \right] \]  
  (pseudo-Kick)

- **Batch**
  \[ E_b = C_b \cdot \left[ \frac{10}{(x_p^{0.5})} - 10/\left(\frac{x_f^{0.5}}{x_f^{0.5}}\right) \right] \]  
  (Bond-Batch)

- **Charles (Form.1)**
  \[ E_c = C_c \cdot \left[ \frac{\alpha_s}{C_2} - \frac{1}{(x_f^{\alpha_s})} \right] \]  
  \[ \frac{\alpha_s}{C_2} \frac{1}{(x_f^{\alpha_s})} \]  
  (Charles (Form.1))

- **Charles (Form.2)**
  \[ E_c = C_c \cdot \left[ \frac{\alpha_c}{C_1} - \frac{1}{(x_f^{\alpha_c})} \right] \]  
  \[ \frac{\alpha_c}{C_1} \frac{1}{(x_f^{\alpha_c})} \]  
  (Charles (Form.2))

- **Rosin-Rammler**
  \[ 1 - P(x,t) = R(x,t) = \exp[-(x/x_0)^{\alpha_r}] \]  
  \[ \alpha_r \]  
  (Rosin-Rammler)

- **Zero Order**
  \[ E_z = C_z \cdot \left[ \phi \cdot (x^*) - F(x^*) \right] \]  
  (Zero Order)

- **Hukki**
  \[ dE = -K \cdot \frac{dx}{x^{(x)}} \]  
  (Hukki)

Table 4. Summary of Grinding Laws and Equations
rate is constant, but finally tends to zero when grinding proceeds. The rule states that this rate of production is independent of the relative amount or size of material larger than this size, within appropriate limits.

Hukki [26] tried to define the areas of applicability of the old laws to the basic grinding curve. He extended the curve (Figure 9) to imaginary ranges and stated that the concepts of proportionality for the Von Rittinger, Kick, and Bond equations also varied with particle size. Therefore, size x is a function of itself. Table 4 includes Hukki’s formula.

The laws discussed are purely empirical, based on data fitting and not on physical phenomena. A distinction must be made between the usefulness of relations for design and their so-called theoretical basis. A proper comprehension will prevent application of the empirical laws outside their true range of validity. The Bond grindability test has been used traditionally to determine the size of mill to be used. This method is based on the mean empirical fit of data from a number of mills and materials run under normal conditions. Because it is based on data fitting, there will be a range of error for any specific mill, material, and set of operating conditions.
Fig. 9. Relationship between energy input and particle size in comminution.

Several correction factors apply. These include:

- an empirical equation to convert test results to the observed results in a 2.44 m i.d. wet overflow ball mill, operated in closed circuit with a circulating load of 350%.
- a scale-up factor to convert the work index to a standard 2.44 m internal diameter mill.
- a correction for oversized feed
- a fineness-of-grind correction if $P_{80}$ is below 75 microns.
- a correction for a low reduction ratio if the quotient of 80% passing sizes of circuit feed and product is below 6.

The latter three factors are based on experience and attempt to deal with milling conditions different from those maintained by Bond.

By making an empirical match of the standard results to actual plant data, the Bond sizing method avoids the differences in specific grinding energy required for a batch test, a locked-cycle test, or for a steady-state continuous mill. Not included are recycle ratio, classifier efficiency, ball size mixture, variations in retention time distribution with slurry density, mill geometry and rheology to name just a few.
Batch grinding equations

The concepts of specific rate of breakage and primary progeny fragment distribution in a complete size-mass balance allow a more detailed analysis of the basic pattern of grinding curves.

When material of size \( j \) breaks, a suite of smaller particles ensue. Before further breakage occurs such suite is called the primary progeny distribution. In cumulative form it is symbolized by \( B_{i,j} \) and is defined as the cumulative weight fraction of material broken from interval \( j \), which appears below size \( x_i \), the upper size of interval \( i \). The fraction of primary breakage products from interval \( j \) that goes into interval \( i \) is denoted by \( b_{1,j} \). This symbol is referred to as the breakage (distribution) function [28].

\[
b_{1,j} = B_{1,j} - B_{i+1,j}
\]  

(3)

The process of primary breakage is conveniently expressed in algebraic matrix symbols.

\[
p = Bf
\]  

(4)

where \( f \) and \( p \) are vectors for feed and product.

To take account of breakage rates, which change from stage to stage of further breakage, the specific rate of breakage function \( S \) was introduced. \( S \), also named the selection function, is given in kg\cdot time^{-1} per kg. The product size distribution of equation (4) is then more correctly expressed as

\[
p = SBf + (I-S) \cdot f
\]  

(5)

The first part on the right hand side in this equation calculates the actual quantity of material of size \( f_i \), which is broken; the second part relates to the residue that passes unbroken.

Most likely the products generated during the breakage process are subjected to separation by size. This may eventuate within comminution equipment or in a subsequent classification stage. Pieces larger than the discharge gap of a jaw crusher stay behind until broken further. When including the classification function into equation (5), the product size distribution becomes more exact:

\[
p = [I-C] \cdot [BS+I-S] \cdot [I-C \cdot (BS+I-S)]^{-1} \cdot f
\]  

(6)

The annotation used in (6) is explained in Figure 10.
Fig. 10. Annotation for the classification function

The fourth law of grinding incorporates the classification function but only the part that pertains to the action within a classifier [27]. The law states that the slurry density in the cyclone overflow should be kept at maximum, while that in the cyclone underflow should be kept at minimum. Separation on size also takes place within comminution. Fines need to be removed from a mill as overgrinding may occur and valuable space is occupied [29]. Their removal relies on transport of material to the classifier and is a function of mill throughput and rheology. Under normal conditions, the charge in a wet, overflow ball mill consists of two distinct regions: a ball charge region that contains all of the grinding media and part of the slurry, and a pool that contains the remainder of the slurry. In a model for mass transport in such mills, it is assumed that by fluid drag a layer of slurry flows transversely from the pool to the ball charge region at the rotating mill shell [30]. Settling of solid particles in the pool leads to a preferential transfer of solids out of the pool and to accumulation in the ball charge region. Here, the axial velocity is practically zero, compared to an axial velocity $u_x$ in the pool. Larger particles spend relatively less time in the pool because of their greater settling velocities. Consequently, these have lower mean axial velocities than the finer material. An expression for the size-dependent axial velocity $u_x(x)$ is given by

$$u_x(x) = u_0/[1 + K_b \cdot (v_s(x)/v)]$$

where $u_0$ is the axial velocity of an infinitely small particle that doesn’t settle, $K_b$ is the fraction of slurry contained in the ball charge region, $v_s(x)$ is the settling velocity of particles of size $x$, and $v$ the characteristic velocity that describes the convective transport of slurry into the ball charge as a result of the relation of the mill:

$$v = w \cdot \delta / k$$

where $w$ denotes the angular velocity, $\delta$ the effective slurry thickness in the pool near the shell, and $k$ a geometric factor. The medium residence time for water is about 10 to 15% less than for solids, which favours a faster movement of fine particles through the mill, as became apparent from the equation above.
Particle breakage is more popularly described by the kinetic population balance or size-mass rate balance. The net rate of production of size i material equals the sum rate of appearance from breakage of all larger sizes, minus the rate of its disappearance by breakage

\[
df_i(t)/dt = -S_i f_i(t) + \sum_{i \geq j \geq n \geq 1} b_{i,j} S_j f_j(t) \tag{7}
\]

Note that this balance is the fundamental size-mass rate balance. The basic form does not incorporate any classification action, but can be resolved in different ways depending on whether size and time factors are taken as discrete or continuous [28]. To indicate its agreement with the old laws the 'Simple Solution' applies. Assumed is that the specific rate of breakage of size interval j ($S_j$), and the fraction of breakage of size interval j to less than size i, are independent of the original size of j. This statement infers that these two effects exactly compensate ('the compensation condition'). In reality, $S_j$ is larger for larger particles, and the fraction of breakage to less than size i is smaller with smaller initial size j. Under these conditions equation (7) evolves into (8)[23].

\[
P(x_i,t) = 1 - \prod_{i = 1}^{n \geq i \geq 1} [1 - P(x_i,0)] \exp(-S_{i-1},t) \tag{8}
\]

The Rosin-Rammler equation results from this solution upon inserting $S(x) = ax^a$ into (8). $S(x)$ is a power function of size. The equation is directly comparable with the one listed in Table 4, where $x_0 = (1/\alpha)^{1/a}$.

Model solutions are related to the equipment applied. For ease of understanding again the algebraic matrix notation is used for a rod mill. Rod milling can be considered as a series of grinding stages in which both breakage and classification occurs. The pronounced preferential breakage of the coarser particles is effected by the rods. These prevent a particle in the maximum size range of the feed to leave a certain stage before it is broken to a size range it is currently in. This repetitive process may be described as follows [22]:

\[
p = \sum_{j=0}^{i \geq i-i} x_j \cdot f \tag{9}
\]

where

\[
x_j = [I-C] \cdot [BS+I-S] \cdot [I-C \cdot (BS-S+I-S)^{-1}] \cdot f \tag{10}
\]

**Limitations in model application**

Several limitations for effective use of a model or simulator should be taken into account. At first, each model is based on results from samples taken from a laboratory batch grind or from a plant sampling survey. Particulate systems are mixtures of particles of varying size, shape and composition. Sample size becomes important when large particles are present (mill feed and discharge); a sizing method is characterized by its upper and lower
measurement and detection limits [31]. Particles smaller than the lower detection limit are ignored, resulting in an apparent size distribution. Especially if this fraction is large the results are quite questionable. For on-stream particle analyzers entrained air is seen as particles. If not removed the results again are questionable.

Both the Bond grindability tests and the Population Model Balancing technique have inherent advantages and disadvantages. The former method is simple to use, relates to costs, measures the overall efficiency, and is a world standard. It is valid as a gross method for mills operating under normal conditions, and relates to overall economy. The calculation, however, does not incorporate a number of second order effects. For example, the empirical oversize correction factor applied to a feed with \( F_{80} \) of 10 mm, gives a very large reduction in mill capacity at low reduction ratios, especially for material with a high work index. Furthermore, according to calculation procedure, the mill capacity is nearly independent of \( x_p \), the product size. A mill operating at a fixed flow rate would give large changes in product fineness when fluctuations in feed rate or material grindability are minor, which contradicts plant practice. And finally, to convert a Bond work index, arrived at for a wet closed circuit, to a value valid for a wet open circuit, the mill capacity is reduced (or work index is increased). The conversion factor depends on the percentage less than the test screen size (P1) desired in the open-circuit method. This factor can be found in tables. For some materials (galena, limestone, and to some degree quartz) the change in test work index with P1 is insignificant. It is thus perceivable that the size of mill selected on the basis of the P1 value may differ, which is illogical.

The Population Model Balance allows controlled experiments and is more detailed due to its complex nature. Precisely the complexity makes the Model unattractive to the average engineer. To design a new grinding system modelling, however, is the most exact method. When a detailed analysis of an existing circuit is warranted a good simulator can be built. Predictions may be uncertain as it includes some 25 factors of each the rod mill and the ball mill. An effectual change of a parameter produces an overall result, but cannot be isolated to either breakage or classification. In such situations a functional performance analysis as described in Figure 11 may define better how individual components within the grinding circuit are affected [32]. The methodology is based on the realization that the relative solids fraction, coarser than some defined cut-off size for the circuit, represents the effective fraction of the total mill grinding volume and applied power. Simultaneously, it defines the efficiency of the overall classification system of the circuit. The relative grinding efficiency of the plant mill environment can be quantified by comparing the ratio of its grinding rate to that of a standard bench test, using a sample of the same ore. Careful measurement of milling parameters must be done parallel to a detailed sampling of the circuit before and after a change. From the observations and the sampling results, the impact on both the breaking and classification efficiencies can be calculated individually, instead of obtaining one value to describe the overall system.

A third consideration not mentioned in the literature is that a grinding circuit should be in dynamic equilibrium before samples are taken. This is expounded further in Chapter 1.4.1. In short, before test work is conducted to optimize a circuit, a degree of stability should be reached. Results of sampling campaigns become more meaningful once the sensitive parameters can be monitored and maintained near their setpoint.
Fig 11.
Development of the Mill Circuit Functional Performance Analysis

- Production rate of new product size material in the mill
- Circuit production rate of new product size material
- Circuit production rate of new product size material

Effective mill power draw (applied to coarse particles)
\[ \times \text{Grinding Rate of Coarse Particles (per unit of effective power applied)} \]

\[ \text{Total mill x Fraction of mill coarse solids inventory} \]
\[ \times \text{Plant energy specific grinding rate} \]

\[ \text{Total mill x Classification power draw system efficiency} \]
\[ \times \text{Ore laboratory x Grinding rate ratio grinding rate} \]

\[ \text{Total mill x Classification power draw system efficiency} \]
\[ \times \text{Ore laboratory x Grinding environment grinding rate efficiency} \]
Another missing component from the size-mass rate balance equation is slurry rheology. Einstein [33] proposed the hydrodynamic consideration that the increased viscosity of a suspension, as compared to that of a pure solvent, is a consequence of the perturbation of flow of the solvent around the suspended particles. This leads to an increased rate of heat dissipation. To derive at his equation he made the following assumptions:

- the suspended particles are rigid, neutrally buoyant spheres whose diameters are large compared to molecular dimensions, but small compared to the characteristic length of the measuring instrument, which eliminates wall effects.
- the flow is purely laminar and slow enough to neglect inertial effects.
- the solvent adheres perfectly to the spheres
- the concentration of particles are so dilute that particles do not exert interaction forces on each other.

For laminar flows the Stokes equation applies, which then can be solved for the solvent flow perturbed by the presence of a singular particle. The equation is restricted by a boundary condition at the surface of the particle, and at infinite distance from the particle surface. Einstein's equation can then be expressed as

\[ \mu_r = \frac{\mu_s}{\mu} = 1 + 2.5 \cdot C \]

where \( \mu_r \) denotes the relative viscosity of the suspension, \( \mu_s \) the effective suspension viscosity, \( \mu \) the viscosity of the suspending fluid, and C the volume fraction of solids. This formula is valid up to a solid concentration of 10%. Many other researchers have proposed empirical equations for suspensions with higher solids content. Each may be deduced to Einstein's relationship, the constant of which varies with the degree of aggregation, particle size, velocity gradient etc.

Newton's law for ideal viscous fluids relate shear stress (\( \tau \)) to shear rate or velocity gradient (\( dv/dy \)). A simple laminar flow of a liquid between two plates obeys the equation

\[ \tau = -\mu \cdot dv/dy \]

where \( \mu \) is the coefficient of viscosity.

When the coefficient of viscosity is constant the fluid is labelled Newtonian. For a great number of suspensions, however, the viscosity may vary by many orders of magnitude with changing rate of shear. These materials are named non-Newtonian: a linear relationship between shear rate and shear stress does not exist. In such case the coefficient of viscosity is defined as apparent viscosity, \( \mu_a \), and obtained by connecting the particular point on the curve with the origin. For pseudoplastic fluids, the apparent viscosity has a larger value than the viscosity coefficient for a Newtonian fluid at the same shear rate. Clay suspensions complicate matters as these typically require a certain yield stress value to be exceeded before flow commences. They are referred to as Bingham plastics.
Rheology is important in both the rate of production of fines in the mill as in the subsequent classification stage. Initially, grinding aids were thought to improve comminution through its effects on crack propagation. Later was demonstrated that propagation velocities are in the order of 1000 m/s, well above the maximum speed at which water molecules can enter cracks by capillary action [24]. Grinding aids find their usefulness in reducing agglomeration, which is promoted by the electric charges on the freshly formed surfaces. They further reduce the adhesion between balls and mill wall. Breakage interactions between grinding media and particles are thus allowed to be more efficient, hence the rate of size reduction is increased. Adequate equipment to measure rheology on-line is not yet available. As a substitute, models include solids density, which permits an approximation for rheology. Viscosity data of several investigators [33] indicate to a Newtonian flow behaviour of suspensions of spheres in the concentration range up to 25 volume % solids, and in some cases up to 45 %. For more concentrated suspensions, the viscosity was found to depend more on the shear rate to which it was subjected. The main factors contributing to non-Newtonian behaviour have been ascribed to particle interactions, particle aggregation and particle orientation. Errors are introduced with a change of ore type, especially when the clay content varies. For a similar pulp density the apparent viscosity of a slurry will increase dramatically with higher clay content.

1.2.2. Fundamentals of Classification

By definition, classification is the separation of particles according to their different settling velocities. A particle settles in a fluid under the influence of a force field, which may be either gravitational or centrifugal in nature. Particles with a relatively high settling velocity separate from the main field and may be removed. By far the most important device is the hydrocyclone, now in use in most grinding circuits. Its common usage and importance in a grinding circuit demands a more detailed discussion here.

In older plants the cyclone replaced the rake classifier and screw classifier. Requiring not only less space and allowing a higher unit throughput per floor area, it is also simpler to install and operate, and produces a higher overflow density. For a throughput of about 320 tonnes/h, and generating a product containing about 20 % solids at a grind of 8% + 210 microns, the minimum pool area required for a spiral classifier is about 16 m² [34]. A similar product is obtained by a cluster of four 0.66 m diameter hydrocyclones. These cyclones would cover a maximum of 25 % of comparable floor space. To its disadvantage is its sensitivity towards fluctuations in feed flow rate and density. In addition it may be argued that a cyclone produces less sharpness of cut. Consider a concentrator treating a lead-zinc ore. Due to its high specific gravity, galena particles of same diameter as sphalerite particles will experience a higher settling velocity (refer to (11)). The higher g-forces in the cyclone, compared to a mechanical classifier, aggravate this difference. Galena tends to report to the cyclone underflow and is returned to the mill.
Cut size

Particle settling is best described by the Stokes Law, which is applicable to laminar flow:

\[ v = \frac{d^2 \cdot (f - \rho_l)}{\rho_s \cdot (1 - \rho_s)} \cdot g/(18 \cdot \mu) \]  \hspace{1cm} (11)

where \( v \) denotes the terminal settling velocity, \( d \) the particle size, \( f \) the specific gravity of solids and liquid, \( g \) the gravitational acceleration, and \( \mu \) the slurry viscosity for Newtonian fluids. Researchers have used the Stokes Law for hydrocyclones despite the fact that the flow is turbulent. Their argument is that the flow surrounding the particle is laminar [35], but \( g \) in (11) is replaced by the centrifugal acceleration \( b \).

The tangential inlet of a cyclone creates a downward vortex along the cyclone wall. Resulting centrifugal forces separate the larger and coarser particles into the boundary layer on the wall, which then leave the cyclone through the apex. Due to the throttling effect of the apex an upward spinning vortex surrounding the cyclone core is created. Taking into account frictional forces, the vortex action can be described by the Helmholtz equation

\[ v \cdot r^n = v_x \cdot x^n = \text{constant} \]  \hspace{1cm} (with \( n = 0.5 \) for water) \hspace{1cm} (12)

where \( v \) is the tangential speed in radial direction, \( r \) is the radius of the cyclone, and \( n \) the exponent, smaller than 1, to describe the effect of solids suspensions.

The significance of this equation is the increase in tangential speed generated in the radial direction with decreasing radius \( x \). Once the radius approaches the vortex diameter, exponent \( n \) assumes the value 1. As the circumferential speed escalates finer particles separate as a second separating stage. These settle horizontally through a conical envelop of zero velocity [36], which divides the flow of fluid towards the apex and the flow towards the vortex, and join the primary vortex. In relating velocity \( v_x \) to the centrifugal acceleration \( b \), it becomes clear that it is indeed the inner vortex that dictates the cut size \( d_{50} \). The parameters \( b \) and \( v \) are defined by

\[ b_x = x \cdot w^2 \]  \hspace{1cm} (13)

\[ v = w \cdot r \]  \hspace{1cm} (14)

where \( w \) represents the angular velocity, \( r \) the cyclone radius, \( x \) any radial position for \( 0 < x < r \), and \( b_x \) the centrifugal acceleration at distance \( x \). Equations (13) and (14) inserted in (12) results in [35]:

\[ b_x = b_o \cdot (r/x)^2 \]  \hspace{1cm} (15)

where \( b_o \) denotes the centrifugal acceleration for \( x = r \). Thus, at any radial distance \( x \) the centrifugal action is higher than at the wall.

How Stokes Law can be applied to a cyclone may be explained by the Equilibrium Orbit Hypothesis [37]. A particle of dimension \( d_{50} \) is in equilibrium with two principal forces
acting on it. Fluid drag force $F_d$ pulls it towards the centre, while the centrifugal force $F_c$ exerts the counter force towards the outer periphery of the cyclone. In equilibrium the cut size follows from (11), in which $g$ is replaced by $b$.

$$d_{50} = \left[18\mu/(\rho_s - \rho_l)\right]^{0.5} \cdot [v/b]^{0.5}$$

(16)

where $d_{50}$ is the cut size, $\mu$ the Newtonian viscosity, $\rho_s$ and $\rho_l$ the specific gravities for solids and liquid, $v$ the tangential velocity and $b$ the tangential acceleration.

The cyclone volume flow capacity can be expressed as a function of cyclone diameter and pressure height $H$ [35]:

$$Q = X \cdot D^2 \cdot H^{0.5}$$

(17)

where $X$ stands for a correction factor to cope with the geometry of the particular cyclone (cylinder length, cone angle, nozzle diameter, etc.).

The volume flow is also restricted by the cyclone area.

$$Q = v \cdot A \quad \text{with} \quad A = 2/3 \cdot \pi \cdot L_e \quad \approx \beta \cdot D^2$$

(18)

The area $A$ is related to the cyclone diameter $D$ of power factor 2.0, moderated by the "slenderness" coefficient $\beta$, which is a ratio of effective cyclone length $L_e$ and diameter $D$. Again, $v$ is the tangential velocity.

A rearrangement of (16) using equations (13),(14),(17), and (18), finally leads to a function that portrays the correlation of cut size $d_{50}$ to the various undependable variables:

$$d_{50} \approx \left[\mu/(\rho_s - \rho_l)\right]^{0.5} \cdot (\beta / X)^{0.5} \cdot (D^{0.5} / H^{0.25})$$

(19)

The first part in the equation relates the cut size to slurry density. It should drop with density. This effect, however, is counteracted by an increase in viscosity, because the movement of coarse particles is hindered by a zone of fines.

The right hand term correlates cut size to cyclone geometry. Specifics become more visible upon analyzing the empirical models by Lynch and Platt, to be discussed in the following subchapter. Equation (19) also explains that the cut size should be finer when applying a higher cyclone pressure.

**Cyclone models**

To date the cyclone action can best be described by empirical models. These models are based on observations that, for a relatively homogeneous ore, the probability of a particle reporting to one of the product streams depends on particle settling characteristics or on size and specific gravity. Their application is restricted to the region in which the measurements were made. Attempts to simulate cycloning with phenomenological models
are in a preliminary stage. These follow the concept of fluid flow dynamics. It has proven
to be difficult to formulate the relative movement of particles with respect to water.
Solutions to Navier-Stokes equations exist for single phase flow [38] and for solid
suspensions.

An empirical model consists of a set of four relations. The Plitt model [37] is the most
widely used today. Table 5 summarizes his equations. It should be noted that the
coefficients are application dependent and additional terms may be considered to
incorporate effects of temperature and viscosity. Lynch [22] defines the cut size $d_{50}$ by a
non-linear regression model

$$\log_{10} d_{50} = K_{1}\phi - K_{2}D_u + K_{3}D_o + K_{4}Q + K_{5}, \quad K_i > 0$$ (20)

where $\phi$ is the vol.% of solids in the cyclone feed, $Q$ is the feed flow, and $D_u$ and $D_o$ are
diameters of apex and vortex respectively. $K_i$ are constants.

Cyclone efficiency

The Tromp efficiency curve [35] is the traditional curve fitting technique to express the
sharpness of separation within a cyclone. Lynch produced the reduced efficiency curve in
which the sharpness factor $\alpha$ can be found by iteration. Plitt [39] developed a model which
calculates the sharpness factor $m$ from a log-log plot.

$$E_c = 1 - \exp[-0.6931 \times (d/d_{50})^m]$$ (21)

Note that the efficiency is expressed as the corrected efficiency. A certain quantity of fines
directly bypass the classification action in a cyclone and report to the underflow. To
calculate the corrected efficiency, therefore, a correction for the bypass should be made.
According to Lynch the bypass may simply be defined as the fraction of water in the feed
that reports to the underflow. Other researchers [23] maintain this is incorrect. Specific
gravity and viscosity effects should be included in its calculation. Later work [40] revealed
that the bypass is independent of the mineral specific gravity and indicate that the Lynch
relationship is valid.

Horizontal cyclones

It is appropriate to mention cyclone action in horizontal position, especially as this may be
of consideration for future technology. Contrary to common belief, it matters how a cyclone
is positioned when it pertains a large cyclone. Naturally, the major force in a cyclone is
exerted by the centrifugal action. The vortex flow increases the static pressure radially
outward. This "centrifugal static head" is primarily determined by the distribution of the
tangential fluid velocities within the flow. In large units gravity has a significant effect on
the removal of solids to the apex, due to the tall liquid column [41]. Horizontal mounting
HORIZONTAL CYCLONES

\[ d_{50e} = \frac{5.683 \cdot 10^3 \cdot D_o^{0.36} \cdot \exp[-0.020\phi]}{D_u^{0.49} \cdot Q^{0.40} \cdot (f_s - f_f)^{0.5}} \]

\[ S = 0.0206 \cdot (D_u/D_o)^{0.82} \cdot [D_u^2 + D_o^2]^{0.36} \cdot \exp[0.070\phi] \]

\[ P = 114.4 \cdot \exp[-0.0182\phi] \cdot [D_u^2 + D_o^2] \cdot Q^{-0.11} \]

\[ Q = \frac{7.9 \cdot 10^3 \cdot \exp[0.013\phi]}{[D_u^2 + D_o^2]^{0.50} \cdot P^{0.11}} \]

\[ m = \exp[0.82 - 0.41 \cdot R_v] \cdot Q^{-0.09} \]

\-----------

PLITT'S EQUATION

\[ d_{50e} = \frac{35D_c^{0.46} \cdot D_i^{0.6} \cdot D_o^{1.21} \cdot \exp[0.063\phi]}{D_u^{0.71} \cdot h^{0.38} \cdot Q^{0.45} \cdot (f_s - f_f)^{0.5}} \]

\[ S = \frac{2.9 \cdot (D_u/D_o)^{3.31} \cdot h^{0.54} \cdot (D_u^2 + D_o^2)^{0.36} \cdot \exp[0.0054\phi]}{H^{0.26} \cdot D_c^{1.11}} \]

\[ P = \frac{4.7 \cdot Q^{1.78} \cdot \exp[0.0055\phi]}{D_c^{0.37} \cdot D_i^{0.94} \cdot h^{0.28} \cdot (D_u^2 + D_o^2)^{0.87}} \]

\[ Q = \frac{0.21 \cdot P^{0.56} \cdot D_c^{0.21} \cdot D_i^{0.53} \cdot h^{0.16} \cdot (D_u^2 + D_o^2)^{0.49}}{\exp[0.0031\phi]} \]

\[ m = \exp[0.58 - 1.58 \cdot R_v] \cdot [(D_c^2 \cdot h)/Q]^{0.15} \]

Table 5. Equations for vertical and horizontal cyclones (after Plitt).
could be considered for D20B units (Krebs 0.51m cyclone) or larger only to benefit from the possible reduction in required feed pressure. At Cuajone, the D26B cyclones, which are 2.3 m long were mounted at an inclination of 13.5°. The effective vertical length of the cyclone was, therefore, reduced to 0.54 m. To obtain the same centrifugal force within the cyclone, the feed pressure could be reduced by 2.7 m water column, assuming 50% solids by weight. This is equivalent to a drop from 62 kPa to 37 kPa.

Classification can thus be attained at lower feed pressure than required before. At several plants (Mexicana de Cobre (Mexico), Lithium Corp.(USA), Bougainville, and Cuajone (Peru)) the circulating load was reduced by at least 200%, primarily due to less fines reporting to the underflow. It allowed up to 10% higher throughput rates at a similar grind. Even a slightly coarser grind might be permitted, because of the sharper separation achieved. Less circulating load translates into a diminished cyclone feed flow. Combined with a reduced operating pressure (from about 50-60 kPa to 25-30 kPa), the pump can be slowed down. As the wear rate is related to the (particle velocity)^m, where m varies between 2.5 and 4 [42], maintenance costs are cut dramatically. Furthermore, from an operators point of view, it is interesting to note that in horizontal position cyclones are much less prone to plug up because of roping.

In yet unpublished research and development work, conducted at Cuajone, attempts were made to establish an empirical model for horizontal cyclones. Results are summarized in Table 5 where a direct comparison can be made with the Plitt model. In this work about 45 data points were correlated. The model produced a good fit for about 35 data points taken at a later date.

K constants are expected to vary with the application, but one comment about the model should be made. In the work with horizontal cyclones the cut size seems to be inversely related to the solids content in the cyclone feed. This needs to be investigated further, but indications are it may have been caused by the mode in which the cyclones operated. A horizontal cyclone does not display a spray discharge for a pressure below about 28 kPa. Underflow density is higher at 85% than the 78% attained at near roping conditions for a vertical cyclone. A possible explanation may be found in Figure 12, where cut size is plotted against cyclone inlet pressure. The data in this Figure were generated from the Cuajone test work. Beyond about 32 kPa, it may be argued that cyclone operation became 'normal': the cyclone started to display a spray discharge and the cut size appears to become finer with increase in pressure, following the Blasius' relationship for smooth hydraulic flow, P (:) Q^1.75. In that region the pressure drop, caused by centrifugal forces, may be sufficiently large that the absence of gravitational forces would have no impact at all. Such pressures can be attained by replacing solids in the feed flow by water. Roping conditions do not occur due to the relatively lower solid volume fraction fed to the cyclone. At about 27 kPa, roping occurs due to the relatively high volume fraction of solids in the feed. When reducing the pressure the roping action is being relieved in accordance with the SPOC correlation [43]. This constraint infers that the solids content in the overflow controls that in the underflow. It predicts the onset of roping. As a consequence, the cut size becomes finer. Mathematically, the SPOC correlation was expressed as
\[ \phi_\text{L} = \phi_{\text{L20}} + 0.2(\phi_\text{f} - 20) \]

where
- \( \phi_\text{f} \) is the \% solids by volume in the underflow at which roping is initiated for a feed concentration of \( \phi_\text{f} \) \% solids by volume.
- \( \phi_{\text{L20}} \) is the user specified value for the limiting \% solids by volume in the underflow at a feed solids concentration of 20 \% by volume.
- \( \phi_\text{f} \) is the feed volume \% solids.

Fig. 12. Cut size versus pressure for horizontal cyclones with 16.5 cm diameter apex and 22.9 cm diameter vortex.

1.2.3. Traditional operating practices of a grinding circuit.

Whether a grinding circuit consists of a one-stage ball mill or a two-stage rod and ball mill configuration, the traditional approach to operation is quite similar. In older plants rod and ball mills are over designed by virtue of the engineering techniques employed. These incorporate inadequacies of the Bond Grinding Tests and unexpected changes in rheology due to temperature or ore constitution effects. During start-up, and in the years a mill operates at design capacity, little concern for limits in mill capacity prevail. Reliable instrumentation was not available and a grinding circuit became operator intensive.

A typical rod-ball mill circuit is depicted in Figure 13. Two or more feeders provide the feed to the primary mill. At least one of these is equipped with a variable speed drive.
motor. Mechanical belt scales measure the tonnage on the feed belt, which is recorded in the control room. The control room operator sets the mill throughput on an analog controller. This regulates the feed to the mill through a series of relays that control the speed of the variable drive. Of all the water flows that enter a grinding circuit only the water to the primary mill is controlled automatically. Water is rationed to feed to ensure a certain density is maintained. Sporadically, manual field checks on mill discharge density are made by way of calibrating the feed to water ratio.

The only other control is on cyclone overflow density. Sometimes the density is measured and controlled by water addition to the cyclone feed pump box. Mostly, an operator periodically checks this density manually. He then asks the control room operator to change the valve setting on the pump box make-up water line accordingly. In either case, to avoid spillage the pump box normally runs very low, if not dry, and the pump surges.

Ball mill power is recorded and consulted to charge the mill with grinding media once it drops below a specified point. Once a grinding circuit is being operated beyond design
capacity, chances of mill overload increase. The control room operator is forced to watch more closely the power recorder chart. If an overload occurs, it is remedied by temporarily shutting off the feed until the power is restored. Invariably, the operator will reduce the tonnage setpoint to avoid reoccurrence.

Until recently, cyclones were installed in series off a main header. Although each subsequent cyclone is being fed by a feed that consequently changes in composition, cyclone geometry was not adjusted for ease of maintenance. Such arrangements are now widely replaced by cyclone clusters.

1.3. Flotation

1.3.1. Theory of flotation

**Electrical double layer**

Solid particles in contact with an aqueous environment carry an electrical charge. Its magnitude depends on both the surface chemistry of the solid and on the solution chemistry of the aqueous environment. The former is a result of lattice sites occupied by ions that may be released into the solution. Also, size reduction, performed to liberate the valuable minerals, can produce an appreciable number of ionic (hydrophillic) sites due to bond breakage. The surface charge of a certain sign thus created will be subjected to the attraction of counter-ions present in hydrated forms in the aqueous solution. Between the solid surface, specifically the first layer of counter-ions, and the solution an equilibrium will be established through the formation of an electrical double layer.

The electrical double layer structure was most adequately described by Stern [44]. He introduced the concept of specific adsorption of ions from solution, which is caused by a field of molecular adsorption forces in the immediate vicinity of the solid surface. According to Stern's theory, the first layer of counter ions is attracted to the surface under the action of both electrostatic and adsorption forces. Surfaces may become polarized whereby metal ions go into solution. Liquid ions of opposite charge adhere to the surface to assume an equal counter potential [45]. The sign and magnitude of the surface charge controls the adsorption of physically adsorbing agents. It results in the formation of a plane condenser of thickness $\delta$, as envisaged by the Helmholtz theory. In this layer the electrical potential decreases linearly from $\Phi_0$ to $\Phi_\delta$, and is variously known as the "Stern layer", the "Helmholtz layer", or the "adsorption layer". Ions of opposite charge to the mineral surface are firmly attached and do not move with changes in temperature. Figure 14 gives an illustration of the electrical double layer.
Fig. 14. Graphic presentation of the electrical double layer.

How reagents affect the double layer may be noted from the Nernst equation

\[ \text{d}E = (\text{RT}/nF) \cdot \text{d} \ln(a^*_n) \]

where \( \text{d}E \) denotes the change in potential between the solid surface and the solution, \( R \) the gas constant, \( T \) the absolute temperature, \( n \) the ion valence, \( F \) the Faraday's constant and \( a^*_n \) the activity of the metal ion.

Integration yields

\[ E(a^*_n) - E(a^+_n = a^0^+_n) = \Phi_0 = (\text{RT}/nF) \cdot \ln(a^*_n/a^0^+_n) \]  

(22)

Activity \( a^0^+_n \) is defined as the activity of the potential determining ions at which the surface charge is zero, the Zero Point of Charge. Measurements, however, produce the potential \( \Phi_\delta \) not \( \Phi_0 \).
Shifting of the liquid film occurs at the slipping plane, also called the outer Helmholtz layer, and not at the surface itself [46]. This plane is roughly \(2\delta\) away from the surface; it is here where the zeta potential arises. The difference in potential between the Stern plane and the slipping plane is small compared to the total potential difference across the double layer. Therefore, \(\Phi_s\) is often said to be equal to the zeta potential. In actual fact, only when \(\Phi_s\) equals zero the two coincide, for there zeta must be zero.

Equation (22) relays directly that the activity of the ions in solution affects the potential in sign and magnitude. In the absence of specific adsorption, the potential drop, \(\Phi_x\), at a distance \(x\) away from distance \(\delta\) for small potentials \((nF\Phi_o \ll RT)\) is described by [47]

\[
\Phi_x = \Phi_\delta \cdot \exp^{-kx}, \quad \text{with } k = \left[\frac{(8\pi z^2 I)}{(\epsilon kT)}\right]^k
\]

and surface charge density \(\sigma_d\) by

\[
\sigma_d = \Phi_o \cdot \left(\frac{\epsilon \kappa}{4\pi}\right)
\]

where \(x\) is the distance from the interface to the solution. The thickness of the Gouy-Chapman layer equals \(1/k\), \(k\) being a function of ion strength \(I\), solution dielectric constant \(\epsilon\), temperature \(T\), and ion charge \(z\). It represents the distance of the centre of gravity of the electrical charge in the diffuse layer from the solid surface. In this equation \(k\) is named the Boltzman constant.

Three-phase contact

In froth flotation mineral particles float to the top of the cell, where they get skimmed off. This is possible because a dynamically stable three-phase contact has been established between the particle, the liquid phase and the gas phase. Schematically, such three-phase contact is presented in Figure 15. Vectors represent surface tensions on contact planes between the mineral surface and liquid (\(\sigma_{SL}\)), mineral surface and gas (\(\sigma_{SG}\)), and between the liquid and gas phases (\(\sigma_{LG}\)). Angle \(\Theta\) is referred to as the contact angle between air bubble and solid surface.

The condition for three-phase equilibrium is given by Young's equation [48]

\[
\sigma_{SG} = \sigma_{SL} + \sigma_{LG} \cos \Theta
\]

The last term on the right hand side in this equation expresses the wettability, also referred to as the adhesion.

Other wetting terms include the spreading coefficient, \(S\), defined by Harkins [49] as

\[
S = \sigma_{SG} - (\sigma_{SL} + \sigma_{LG})
\]
Fig. 15. Schematic representation of the equilibrium contact between an air bubble and a solid immersed in a liquid.

The liquid spreads out partially on the solid surface and creates a floatable condition for $S < 0$. This conclusion may be drawn when considering the work of adhesion of liquid to the solid, $W_A$, and the work of cohesion of the liquid, $W_C$, which are given by

$$W_A = \sigma_{SG} + \sigma_{LG} - \sigma_{SL}$$  \hspace{1cm} (26)

$$W_C = 2\sigma_{LG}$$  \hspace{1cm} (27)

Combining equations (25), (26), and (27) yields

$$S = W_A - W_C$$  \hspace{1cm} (28)

For partial spreading, $S < 0$, cohesion within the liquid is greater than the adhesion of the liquid to the solid surface. Theoretically, maximum hydrophobicity implies a complete absence of interaction of solid with water, i.e. $W_A = 0$. The work of adhesion for a solid-liquid pair can be represented by the sum of terms for the different kinds of interaction [50]:

$$W_A = W_{d_A} + W_{h_A} + W_{\pi_A} + W_{p_A} + W_{e_A}$$

where superscripts $d$, $h$, $\pi$, $p$, and $e$ represent respectively London dispersion forces, hydrogen bonding, interfacial pi-bonds, other polar interactions, and electrical double layer effects.

When the solid and gas phases are mutually saturated as the equilibrium requirement, the solid surface is often covered with a film of adsorbed gas phase. This lowers the surface-free energy of solid by an amount of $\pi^*$, which is known as the surface pressure of the adsorbed film.

$$\sigma_{SG} = \sigma_s - \pi^*$$  \hspace{1cm} (29)
where \( \sigma_s \) is the surface-free energy or surface tension of solid in the absence of any adsorbate (i.e. vacuum). Equations (24) and (29) combined result in

\[
\sigma_{LG} \cos \theta = \sigma_s - \pi^\gamma - \sigma_{SL}
\]  \hspace{1cm} (30)

This equation indicates that an increase in \( \pi^\gamma \) results in a larger contact angle \( \theta \).

It is often difficult to experimentally determine the true contact angle as described by Young's equation (24). At a surface over which a liquid is advancing the angle is greater than at a surface from which it is receding. Sulman [51] referred to the difference as the hysteresis of the contact angle. Wenzel [52] explained the hysteresis through surface roughness. He defined the roughness factor \( r \) as the ratio of the actual surface area to the geometric plane area, and used it as a correction factor in the following expression

\[
\cos \theta = \frac{1}{r} \cdot \cos \theta'
\]

where \( \theta' \) is the apparent or measured contact angle, and \( \theta \) the true contact angle.

**Critical surface tension of wetting**

A relationship between the wettability of a solid and the surface tension of a wetting liquid was found by Zisman [53]:

\[
\cos \theta = 1 - b(\sigma_{LG} - \sigma_c)
\]

where \( b \) is the slope of the line, \( \theta \) the wetting angle, \( \sigma_{LG} \) the surface tension between the liquid and gas phases, and \( \sigma_c \) the critical surface tension of wetting. A surface tension below the value \( \sigma_c \) allows the liquid to spontaneously spread the surface completely. Especially for strongly hydrophobic solids the above relationship is linear as solid-liquid interaction is thought to result from London dispersion forces only. In some cases, deviations from linearity are apparent for \( \sigma_{LG} \) values above 50 mN/m. According to Zisman the curvature results from weak hydrogen bond formation between the molecules in the liquid and in the solid surface.

Coals are solid hydrocarbons containing various quantities of oxygen, nitrogen, sulphur, and some other elements depending on the rank of coals. Their \( \sigma_c \) values appeared nearly constant at 45 mN/m, but \( b \) values varied and were indicative to the level of hydrophobicity [54]. In such case, a high \( b \) value means a larger contact angle for liquids with \( \sigma_{LG} > \sigma_c \). A correlation between the rank of coals and the \( b \) value of their respective Zisman plot is, therefore, expected. The \( \sigma_c \) values vary with the packing of atom groups: closed packed -CF\(_3\) groups exhibit the lowest surface energy (\( \sigma_c = 6\text{-}28 \text{ mN/m} \)), but increases in the order of hydrocarbons (\( \sigma_c = 22\text{-}35 \text{ mN/m} \)), chlorocarbons (\( \sigma_c = 39\text{-}43 \text{ mN/m} \)), and nitrated hydrocarbons (\( \sigma_c = 40\text{-}45 \text{ mN/m} \)) [55].

The Zisman parameters (\( \sigma_c \) and \( b \)) are useful in characterizing the degree of non-polarity of low-surface-energy solids. These solids, however, invariably possess some polar sites,
which could be due to impurities like surfactant molecules used in emulsion polymerization, as in the production of Teflon. The surface concentration of the polar sites can vary enormously. Teflon 6 can accept highly polar adsorbates such as water [56], as it contains about one polar site per 200 surface sites. An increase in polar sites is accompanied by an increase in $\sigma_c$, and a simultaneous decrease in $b$.

<table>
<thead>
<tr>
<th>MINERAL</th>
<th>$\sigma_c$ (mN/m)</th>
<th>$b$ ($10^5$)</th>
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</thead>
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<td>Talc</td>
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</tr>
<tr>
<td>Stibnite</td>
<td>43</td>
<td>2.86</td>
</tr>
<tr>
<td>Chalcocite</td>
<td>44</td>
<td>3.00</td>
</tr>
</tbody>
</table>

Table 6. Zisman parameter values for various mineral samples [57]. The critical surface tension of wetting below which liquid spreads spontaneously is given by $\sigma_c$, and $b$ is the slope in the Zisman relationship.

Kelebek [57] published Zisman parameter values for various mineral samples, which are reproduced in Table 6. The critical surface tension value $\sigma_c$ for galena is nearly the same as those reported by Olsen [58] for both allotropic forms of sulphur. This may suggest the presence of elemental sulphur on the galena due to oxidation. The values for the other mineral samples approach those of various coals ($\sigma_c = 47.5-49$ mN/m) [54]. This would place the minerals and the coals as the low surface energy solids near the nitrated hydrocarbons, which feature a similar range of $\sigma_c$ [55]. This might imply additional support for the inherently hydrophobic character of common sulphides such as pyrite and chalcopyrite, as is detailed in Chapter 1.4.2.

**Adsorption density**

A bubble can only adhere to a surface if the change in free energy, $\Delta G$, as given by Dupré [48], is negative.

$$\Delta G_s = \sigma_{sg} - (\sigma_{sl} + \sigma_{lg})$$  \hspace{1cm} (31)
This condition is satisfied for \( \cos \theta < 1 \), or \( \theta > 0 \), as follows from (24) and (31). Surface tensions between a solid-gas and solid-liquid interfaces are usually large, however, and must be reduced before bubble attachment can take place. Only for the contact plane between a gas and a liquid equilibrium may exist. When the radius \( R \) of a bubble increases the rise in surface energy must be equal to the pressure \( P \) exerted on that bubble

\[
(4\pi(R+dR)^2 - 4\pi R^2)\sigma = (4/3 \cdot \pi (R+dR)^3 - 4/3 \cdot \pi R^3)\Delta P
\]

When terms of higher order are ignored the equation reduces to

\[
\Delta P = 2\sigma / R
\]  

(32)

For a large bubble the internal pressure becomes too small and renders the bubble unstable.

In a solid the mobility of atoms is restricted. They may take a long time to rearrange in order to create equilibrium, by which the surface tensions between solid-gas and solid-liquid interfaces can be lowered. As a solution to this problem, chemical reactions may be forced upon the surfaces of particles. An example of such mechanism is the adsorption of reagents, like collectors. In thermodynamic terms, adsorption is defined as a change in surface concentration. It can be quantified by the adsorption density. The adsorption density, \( \Gamma_i \), is defined as the excess number of molecules per unit area in a system, over the number of molecules that would be contained in that system if the density of each phase were to remain uniform up to the dividing plane between the two bulk phases [59]. Gibbs-Duhem [59] deduced a correlation for the adsorption isotherm from the first and second laws of thermodynamics:

\[
\Gamma_i = -(1/RT)\ln[d\sigma/d\ln a_i]_t
\]  

(33)

where \( \Gamma_i \) is the adsorption density of species \( i \), \( \sigma \) the surface tension, and \( a_i \) the chemical activity of species \( i \). The adsorption density will be positive if the term \( (d\sigma/d\ln a_i) \) is negative, as occurs when the concentration of the chemical increases and the surface tension decreases. To float a particle the adsorption density of collector \( R \) on the solid surface in contact with air should be greater than on the part that is in contact with the liquid:

\[
\Gamma^S_R > \Gamma^SL_R
\]

Again, this probability can be enhanced through chemisorption when a mineral surface oxidizes, or through reducing the solid-liquid surface tension by influencing the electrical double layer.

Equation (23) illustrates that the addition of an electrolyte may compress the potential in the double layer. The effective thickness of the diffuse double layer \((1/\varepsilon)\) varies inversely with the square root of ionic concentration of the solution. The sign can even be reversed upon which the collector used will be prevented from adsorption onto that particle. Stern quantified this effect on the adsorption density.
Assume a steady state for the concentration profile at a charged surface. In the diffuse layer, the movement of ions towards the charged surface due to electrostatic forces is in equilibrium with the flux away from an area with high ion concentration, due to the Brownian motion. In the form of a differential equation this equilibrium can be formulated as

\[
dC_i / dx = -(C_i z_i F) / RT \cdot d\phi / dx
\]

where \( C_i \) refers to the concentration of species \( i \), \( x \) the distance from the interface to the solution, \( z \) the ion charge, \( F \) the Faraday constant, \( R \) the gas constant, \( T \) the absolute temperature, and \( \phi \) the potential. This equation integrates to

\[
C_i = C^b_i \cdot \exp(-z_i F\phi / RT)
\]

where \( C^b_i \) denotes the concentration in the bulk of the solution. At the Slipping-plane \( C_i = C^b_i \) and \( \phi = \phi_{2\delta} \) (i.e. the zeta potential). Furthermore, the adsorption density, \( \Gamma_i \), is given by \( n_i / A \) or \( 2r_i C^b_i \), with \( r_i \) being the radius of the adsorbed ion. Substitution into (34) then results in the Stern-Grahame expression [46]

\[
\Gamma_i = 2r_i C^b_i \cdot \exp(-z_i F\phi_{2\delta} / RT)
\]

The exponential term signifies the electrostatic energy of adsorption. When chemisorption takes place this energy term is augmented with a factor \( \phi \), the specific adsorption potential. The Stern-Grahame expression then reverts to

\[
\Gamma_i = 2r_i C^b_i \cdot \exp(-z_i F\phi_{2\delta} / RT + \phi)
\]

Such occurrence may be exemplified by the adsorption of the copper ion onto sphalerite. At the surface of sphalerite an ion exchange adsorption happens and copper ions are adsorbed to a monolayer saturation. On these bridges xanthogen is adsorbed, again to a monolayer [60]. For \( \phi \to -z_i \phi_{2\delta} \) the surface charge may change sign.

Collectors

Collectors of importance in flotation are hetero-polair in nature: these consist of a polair part that adsorbs onto a surface, and a hetero-polair part that sticks into the solution. The hetero-polair carbon chain effects the hydrophobicity of the mineral through reducing the surface energy. Once a mineral is hydrophobic, attachment of air bubbles becomes possible. Adsorption often is a combination of an electro-chemical and a chemical reaction.

In the case of pyrite, it has been established that xanthate oxidizes to dixanthogen in the presence of pyrite and oxygen, but not in the presence of oxygen alone [61]. Evidence exists that the negative xanthate anion does not adsorb onto the surface of pyrite, although this could be possible below pH 6. Figure 16 presents the pyrite surface charge as a function of pH in the presence and absence of potassium ethyl xanthate. The zeta potential of pyrite does not appear to be influenced by the presence of xanthate. Fuerstenau [62] established,
however, that below pH 9 the concentration of the xanthate anion is insignificant compared to the dixanthogen concentration. Redox potential measurements of xanthate/dixanthogen systems were found to coincide with the rest potential on the surface of pyrite [63]. Taking into consideration that ordinary pyrite is covered by a surface film from either adsorbed oxygen or an oxygen-sulphur compound [64], Majima and Takeda proposed the following mechanism

\[ 2X^- + X_2 + 2e^- \]  \hspace{1cm} (35) 

In this system oxygen is reduced to hydroxyl ion. The cathodic reduction reaction is given by

\[ \frac{1}{2}O_2(\text{ads.}) + H_2O + 2e^- \rightarrow 2OH^- \]  \hspace{1cm} (36) 

This reaction is similar to one proposed by Fuerstenau [62], but considers the equilibrium between adsorbed species and bulk solution species. The reaction further suggests that the pyrite surface acts as a catalyst in allowing one redox system (eq. (36)) to equilibrate with another (eq. (35)).

![Graph showing zeta potential as a function of pH](image)

**Fig. 16.** Zeta potential of pyrite as a function of pH in the absence and presence of potassium ethyl xanthate [62]

In the galena-xanthate system, the reaction of xanthate with oxygen to form dixanthogen is catalyzed by the presence of ions of several transition metals. The rate of catalyzed formation, however, appeared too low to be significant for galena flotation [65]. Fuerstenau [61] supported the theory that galena flotation fulfils according to a chemical model. First,
the xanthate ion reacts chemically with surface lead ion to form a 1:1 lead-to-xanthate insoluble complex. For a charged species to chemisorb on a surface, an ion of similar charge, assumed to be the hydroxyl ion, must be replaced. The second step in xanthate adsorption on galena occurs following sulphide ion oxidation to thiosulphate and sulphate. Multiple layers of lead xanthate form by metathetic replacement of these oxidized products on the surface. This theory seems to be confirmed by the similarity in heats of reaction of ethyl-xanthate with galena, open to the air, to the thermo-chemical measurements on metathetic formation of lead xanthate from lead thiosulphate, lead sulphate and lead carbonate. Also Finkelstein’s inability to detect elemental sulphur on the surface of galena [108] seems in accordance with this theory. As will be discussed in Chapter 1.4.3, however, it will become clear that most recent investigations support the electrochemical model of flotation, which incorporates the formation of $S^+$, or a metal-deficient sulphide layer on the surfaces of sulphide minerals.

Pyrite is easily depressed by cyanide or hydroxides. Depressing reagents interfere with cathodic reactions, thereby preventing the anodic production of dixanthogen and effecting depression. Figure 17 presents data on the interrelationship between pH and collector concentration necessary for flotation using sodium diethyl dithio-phosphate. This Figure was originally produced by Wark and Cox [61]. Together with Figure 16, Figure 17 shows that, for this collector, hydroxyl depression of pyrite may simply be due to the creation of a negative surface charge that prevents the adsorption of the dithiophosphate anion.

![Graph showing the relationship between pH and concentration of collector for pyrite, galena, and chalcopyrite.](image)

**Fig.17.** Relationship between concentration of sodium diethyl dithiophosphate and critical pH (Wark and Cox)

In a xanthate system, depression of pyrite does not occur until the pH is much higher. The effect of pH depends on the ratio of adsorption densities of collector R and hydroxyl ion. It is a function of the respective concentrations
\[ \Gamma_{X^-} / \Gamma_{OH^-} = f\{[X^-]/[OH^-]\} \]

For a constant adsorption density, a tenfold increase in hydroxyl ion concentration required a tenfold increase in collector concentration [66]. Regardless how pH is increased (using KOH, NaOH, K₂CO₃, or CaO), complete depression will occur, probably due to driving reaction (36) to the left. At a pH of about 6, there is depression of pyrite, but not because of a lack of dixanthogen. The exact mechanism is not well understood, but formation of hydrophillic Fe(OH)₃ will occupy surface sites, thus preventing adsorption of dixanthogen. Above pH 10.5, pyrite is not floatable in the xanthate system. The mixed potential created in an aerated system appears to be higher than the redox potential of the xanthate/dixanthogen couple [63]. It was concluded, therefore, that because of the cathodic effect of oxygen, the electrode potential of pyrite above pH 10.5 would be lower than that necessary to produce dixanthogen.

![Graph showing zeta potential of pyrite as a function of pH upon addition of KCN](image)

**Fig.18. Zeta potential of pyrite as a function of pH upon addition of KCN [67]**

In the presence of the cyanide ion the zeta potential decreases as shown in Figure 18 [67]. The zeta potential changes of sign and renders the mineral repulsive to adsorption of the particular collector. In this work it was proposed that the Fe(CN)₆⁴⁻ complex was the species responsible for depression. A calculation was reported to indicate that almost all iron above pH 6.4 exists in this form. Experimental flotation data corresponded well with the regions of depression predicted by the Pourbaix diagram shown in Figure 19. It should be noted that this diagram represents thermodynamic data of bulk species only. Further divulgence is offered in Chapter 1.4.3.
Above pH 11, galena is depressed due to the formation of plumbite [Pb(OH)$_3^-$], which is the stable species of the lead ion under those conditions (Figure 34). Domains of stability of metal ions in the presence of complexing ions of for example cyanide, sulphide, etc.,

![Graph showing pH vs. Eh for Fe$^{+3}$, Fe$^{+2}$, HCN, FeOH$_3$ at pH 0-14.](image)

**Fig. 19.** Stability fields of Fe$_4$[(CN)$_6$]$_3$ and Fe(OH)$_3$ at 3 · 10$^{-4}$ M dissolved sulphur, 5 · 10$^{-5}$ M dissolved iron [67]

--- 6 · 10$^{-3}$ M cyanide

----- 6 · 10$^{-4}$ M cyanide

...can be visually represented in Eh - pH diagrams. Examples are Figures 19 and 34. These provide a window to the metallurgical engineer to which metal complexes prevail where, what the possible surface charge might be, and indicate a direction to how he could chemically influence the outcome of test or plant results. Such diagrams are essential to the hydrometallurgist, as they show regions of corrosion, immunity, and passivation of metals. As an example may serve zinc plating, where the solution must be purified carefully to avoid the thermodynamically more favourable hydrogen discharge. Reaction is dependent on H$_2$ overpotential. They are not much in vogue in the mineral industry except for some laboratories.
Dynamic flotation conditions

Thus far, it has been assumed that particles will float after adsorption of a collector. Flotation, however, is a process of probabilities. Trahar [68] states the probability of flotation \( (p_f) \) as

\[
p_f (:) = p_c \cdot p_a \cdot (1-p_d), \quad 1-p_d = p_s
\]

where \( p_d \) denotes the probability of detachment. The probability of collision \( (p_c) \) is a complex relation to particle size and increases with size; the product of the probabilities of attachment \( (p_a) \) and floatability \( (p_s) \) is a function of hydrophobicity and size. With increase in hydrophobicity and decrease in size its yield becomes larger.

Under dynamic flotation conditions, the events leading to the recovery of particles are usually considered to occur in three main stages:

- **a** bubble-particle collision
- **b** thinning and rupture of the disjoining film which separates the colliding bubble and particle to allow bubble-particle attachment to take place
- **c** formation of the stable bubble-particle aggregate capable to withstand considerable disruptive forces operating in a flotation cell

Collision

Many researchers have investigated the particle-bubble collision process. Each based their analysis on the equation of motion of a spherical particle relative to a spherical bubble (that is large compared to the particle) rising in an infinite pool of liquid. Hydrodynamic drag will tend to sweep the particle around the bubble, following the fluid streamlines. Particle inertia and gravity act in a combined manner to move the particle out of the fluid streamline and toward the top of the bubble [69]. The equation of motion is

\[
1/6 \cdot \pi d^3 \cdot \delta v_j/\delta t = 1/6 \cdot \pi d^3 \cdot f_d (p^ - f) g_j + 1/8 \cdot \pi C_d \cdot f \cdot d^2 (u_j \cdot v_j)^2
\]

(37)

where \( v_j \) is the particle velocity in the j direction, \( u_j \) is the velocity in the j direction that the liquid around the bubble would have if the particle were not there, \( g_j \) is the component of gravitational acceleration in the j direction, \( d \) the particle diameter, \( p \) and \( f \), the densities of particle and liquid, and \( C_d \) is the drag coefficient. The second term on the right hand side in equation (37) is the steady state drag term; the first term on the right hand side is the particle body force due to relative gravitational acceleration. The term on the left hand side in the equation represents the body force accounting for particle inertia. Figure 20 shows a diagram of a particle approaching a gas bubble.

According to Jameson [70] no term relating to the contact angle appears in the expression for \( p_c \), the probability for particle-bubble collision. This probability is related to particle and bubble diameters, particle and liquid densities, liquid viscosity, and relative particle - bubble velocity. It should, therefore, not be influenced significantly by the chemical nature of the
environment, except indirectly through the number of bubbles generated per unit volume of air and the relative electrical charges of particles and bubbles.

A particle collision with a bubble occurs over a defined time period, the particle-bubble contact time. This period is a function of the chemical nature of the environment, as it is related to the surface tension. It should be lengthened as much as possible to extend the contact time to beyond the induction time. The latter is defined as the time period required for subsequent thinning and rupturing of the water film. To attach a particle to a bubble, it is essential that the induction time is shorter than the contact time.

\[ R_c \]

\[ d_p \]

\[ d_b \]

\[ \theta \]

\[ x \]

\[ y \]

**Fig.20. Illustration of a particle approaching a gas bubble.**

The time elapsed from the moment of collision to maximum bubble deformation can be divided into two intervals. In the first the particle velocity along the gravitational direction is decelerated to zero. In the second interval, the velocity of the particle opposite to the gravitational direction is accelerated by the buoyant bubble until \( v_y = u_y \). At this moment maximum deformation is achieved, and the time elapsed from the beginning of the collision equals half the contact time. As the bubble restores its shape until \( t = \tau \), the velocity of the
particle opposite to the gravitational direction is further accelerated. If thinning and rupture of the disjoining film does not transpire, the particle will bounce off and/or start to slide over the bubble with the restored water film separating the two bodies.

Ye and Miller [71] derived an expression for the particle-bubble contact time for the collision/deformation event. For a touching angle $\theta$ less than 45° between gravitational axis and bubble-particle centre line at the moment of collision, the average value for the contact time, $\tau_c$, is

$$\tau_c = 1.5\left(\frac{\pi m^p_b m_p}{2 \sigma (\cos \theta - \sin \theta)^2}\right)^{1/3}$$

(38)

where $m^p_b$ is the effective mass of the gas bubble (i.e. the bubble mass itself increased by the buoyant force on the bubble during collision), $m_p$ is the mass of the particle, and $\sigma$ is the surface tension. Note that the contact time will be greater as the surface tension is reduced and the particle diameter is larger. At touching angles greater than 45° only sliding occurs. Schulze [72] improved on the concept proposed by Evans, who considered the liquid-gas interface to act similarly to a spring where the elasticity is caused by surface tension. He calculated the contact time $\tau_c$ from equations describing simple harmonic motion:

$$\tau_c = \left(\frac{\pi m_{eff}}{4 \sigma f}\right)^{1/3}$$

where $m_{eff}$ is the effective oscillating mass, $\sigma$ is the surface tension, and $f$ is a complex function of particle diameter, liquid density, surface tension, and touching angle $\theta$.

![Fig.21. Calculated contact time with respect to particle size and bubble size for a particle specific density of 2.6 and surface tension of 72.75 dyne/cm, using equation (39)[71].](image)
A graphical representation of the contact time with respect to particle size and bubble size is given in Figure 21, for a particle density of 2.6 and \( \sigma \) is 72.75 dyne/cm. The curves are calculated from equation (38), assuming an expected touching angle \( E(\theta) = \arcsin(2/3) \). Equation (38) then reduces to

\[
\tau_c = [(\pi m_b^e m_p)/\{\sigma(m_b^e + m_p)\}]^\gamma
\]

(39)

The actual values of the contact time in the plateau region may vary slightly due to the difficulty in calculating the exact effective bubble mass \( m_b^e \).

When the effective mass of a bubble is very large, the buoyant viscosity of the bubble is not affected by collision with a particle of small mass. The contact time with respect to particle size will be a straight line on a log-log plot. From this plot (Fig.21) it can be seen that the contact time increases with particle size, but plateaus where this time becomes independent of the particle size. By enlarging the bubble size, however, the contact time can be increased further. Assume that a particle, 100 \( \mu m \) in size, has an attachment time of 1 ms (point a in Fig.21), it demands a bubble of \( d_b^e = 400 \mu m \) to effect flotation. Only then, the contact time exceeds the attachment time. The Figure further indicates that particles a and b (two different mineral species of same size) may be separated by flotation by properly controlling the bubble size.

**Attachment**

Usually, the attachment time is recognized to consist of two components, the true induction time plus the displacement time. The first component is the time required for the disjoining of the water film between the solid phase and the gas phase to such a thickness that rupture of the water film takes place. The second component is the time needed for the water film to be displaced to such an extent that the attachment is stabilized under the prevailing force field [93].

The disjoining pressure \( P_d \) is a result of the pressure component due to the Van der Waals forces, to the deformation of the electrical double layers \( P_{el} \), and due to hydration steric repulsion forces. Depending on the individual contribution of each component, a particle surface may or may not acquire a stable film in an aqueous suspension. Hydrophillic particles feature a high \( P_d \) because of an indefinitely stable aqueous liquid layer attached to their surface. Its magnitude is a function of the thickness of the film. Blake and Kitchener [73] found that over a range of 60 to 200 nm that stability was due to electrical double layer forces (i.e. high \( P_{el} \)), while hydrogen bonding played a minor role. Below 20 nm, thinning may depend on the capability to form a hydrogen bond, and on an increased Van der Waals attraction \( P_{vv} \). Unoxidized sulphide minerals are not capable to form hydrogen bonds; in that case the magnitude of \( P_{vv} \) can influence considerably the thinning of the aqueous film in the absence of significant electrical double layer forces. A measure of attractive force of water to mineral is expressed in terms of the Hamaker constant \( A_{12} \).
\[ A_{12} = \pi^2 N_1 N_2 C_{12} \]

where \( N_1 \) and \( N_2 \) are the numbers of atoms per unit volume in materials 1 and 2, and \( C_{12} \) is the London dispersion force constant between atoms 1 and 2.

The concept of the induction time is the link between the chemistry of flotation and the physics of flotation, and is a function of the surface features, rate of film displacement, particle size and density, and of captive-bubble velocity.

Various researchers reported that the attachment time for a given particle type is related to the particle size [71]. As the size increases the attachment increases, which produces an almost linear relationship on a log/log plot with different slopes for particles of different density [93]. An illustration of such a plot is given in Figure 22.

![Graph showing comparison of measured attachment time for five coal types with contact time calculated by equation (39)][93]. (open square: sub-bituminous; open circle: high-volatile bituminous; open triangle: anthracite; open hexagon: medium-volatile bituminous; closed triangle: low-volatile bituminous)

The contact time calculated from equation (39) for bubble size 200 and 600 \( \mu \)m, \( \sigma = 72.75 \) dyne/cm and \( \rho_p = 1.3 \) g/cm\(^3\), is also given in Figure 22. The graph suggests that large particles show a poor flotation response because the attachment time required exceeds the contact time. For more hydrophobic particles, the attachment time versus particle size shifts to a lower line. Flotation, however, is only effective for a sufficiently large bubble size. If a fine particle can reach the bubble surface during the bubble-particle motion, then its floatability should equal that of a medium sized particle. Due to their small momentum and
the strong effect of fluid motion around the bubble surface, the probability of collision for
fine particles is much smaller.

To put the effects of hydrophobicity (φ) and particle size into perspective Trahar designed
the relationship which is reproduced in Figure 23. His diagram is based on the experimental
evidence that increased reagent levels are required to float particles of larger size to the
same extent as small particles. An increase in dosage enhances φ. Hence, \( p_f \) increases as
\( p_e \cdot p_s \) will be larger. Collectors preferentially adsorb onto small particles, leaving insufficient
quantity to cover large particles completely. Therefore, hydrophobicity conditions for small
particles are more favourable and the rate of flotation is primarily determined by \( p_c \), which
increases with particle size. This concludes that small particle flotation is determined by
more available flotation time to satisfy \( p_c \).

![Graph showing the relationship between floatability, hydrophobicity, and particle size](image)

Fig.23. Conceptual relationship between floatability, hydrophobicity, and particle size [68]. Floatability can be thought of as recovery.

**Collection**

Dobby developed a collection model for a single particle-bubble system, expressed as the
collection efficiency \( E_c \) [74]. This efficiency is the product of the collision and attachment
efficiencies \( E_c \) respectively \( E_a \):

\[
E_k = E_c \cdot E_a
\]

(40)
The efficiency of collision is defined as the rate at which the particles collide with the bubble, divided by the rate at which particles flow across the projected bubble area. He was able to calculate \( E_c \) from a multiple regression model using 40 sets of conditions, representing realistic combinations of the Stokes number \( S_k \), the bubble Reynolds number \( Re_b \), and the ratio of particle velocity relative to liquid \( \left( u_p / u_b \right) \) and bubble rise velocity \( u_r \).

\[
E_c = E_{c0} \left( 1.627Re_b^{0.06}S_k^{0.54}\left( u_p / u_b \right)^{-0.16} \right)
\]

(41)

\( E_{c0} \) represents the collision efficiency for \( S_k \to 0 \), and was calculated from (41) for \( S_k = 0.05 \). The Stokes number is defined as

\[
S_k = \frac{1}{9}\left( \frac{f_p}{f_l} \right) \left( \frac{d_p}{d_b} \right)^2 Re_b
\]

and \( Re_b \) as

\[
Re_b = \frac{u_b \cdot r \cdot d_p}{\mu}
\]

where \( \mu \) is the viscosity for Newtonian fluids, \( d_p \) and \( d_b \) are the diameters of particle and bubble, and \( f_p \) and \( f_l \) the densities of particle, respectively liquid. Bubble and particle size effects are incorporated, and the collision efficiency increases with particle size. This can also be seen from Figure 21, where a larger particle incurs a longer contact time.

The attachment efficiency \( E_a \), defined as the fraction of particles that attach, is more complex. It requires knowledge of the distribution on the bubble surface, of particle collision angles \( \theta_p \), and of the angle at which fluid streamlines start to carry the particle away from the bubble (maximum angle of contact \( \theta_m \)), and the sliding velocity.

The distribution of particle collision angles \( n_\theta \) is approached by

\[
n_\theta = \frac{\sin^2 \theta_p}{\sin^2 \theta_c}
\]

(42)

representing the fraction of all colliding particles that collide between the front stagnation point (top of bubble) and angle \( \theta_p \). Angle \( \theta_c \) stands for the angle of closest approach of liquid streamlines. For Stokes flow \( \theta_c = 90^\circ \), and thus \( n_\theta = \sin^2 \theta_p \). The maximum angle of contact \( \theta_m \) is defined as the angle at which the radial component of the particle settling velocity towards the bubble, equals the radial component of the liquid velocity directed away from the surface. It is related to \( \theta_c \) and the particle density \( f_p \) according to:

\[
\theta_m = 9 + 8.1f_p + \theta_c(0.9 - 0.09f_p)
\]

From this relationship it can be seen that as the particle density increases \( \theta_m \) increases.

The third important parameter is the particle sliding velocity \( u_{ps} \), which is a function of bubble rise velocity \( u_r \), particle and bubble sizes \( d_p \) respectively \( d_b \), the angle of contact \( \theta \), and the tangential velocity gradient \( \xi \). The latter parameter must be included because on the upper surface of the bubble exists a significant tangential velocity gradient. This term is also referred to as the surface vorticity and is defined as [74].
\[ \Phi = a + b\theta + c\theta^2 + d\theta^3 \]

where a, b, c, and d are third order polynomial regressions as a function of the Reynolds number. Figure 24 shows for Re = 290 the tangential component of the \( u_u/u_b \) ratio, \( u_{\theta}^* \), versus the distance from the sphere surface. Here (r-1) refers to the dimensionless distance from the sphere surface (at \( r_b, r = 1 \)). Seeley’s velocity data is approximated by two linear functions

for \( r - 1 \leq 0.06 \)
\[ u_{\theta}^* = 0.7\Phi(r - 1) \quad (43) \]

for \( r - 1 > 0.06 \)
\[ u_{\theta}^* = 0.7\Phi(0.06) \quad (44) \]

For \( (r - 1) = (d_p/d_b) \leq 0.03 \) the particle tangential velocity then is

\[ u_{\theta} = 0.7\Phi u_u[0.06(d_p - 0.03 d_b)/d_p + 0.03(0.03 d_b)/d_p] + u_p \sin \theta \]

For a particle colliding at angle \( \theta_n^* \) and sliding to an angle \( \theta_n \) (at which it begins to travel away from the bubble surface), the sliding time \( t_s \) is given by

\[ t_s = [(\theta_n - \theta_n^*)/360]\pi d_p/u_{\theta}^* \quad (45) \]

where \( \theta \) is in degrees and \( u_{\theta}^* \) is the particle tangential velocity.

A particle attaches to a bubble when the slipping time \( t_s \) at least equals the induction time \( t_i \). For \( \theta_n^* = \theta_n \) when \( t_s = t_i \), equation (45) reduces to

\[ u_{\theta} \cdot t_i = [(\theta_n - \theta_n^*)/360]\pi d_p \]

and by rearranging

\[ \theta_n^* = \theta_n - (360 \cdot u_{\theta} \cdot t_i)/\pi d_p \quad (46) \]

Replacing \( n_b \) in (42) by the attachment efficiency \( E_A \) can be done for \( \theta_n = \theta_n^* \) at \( t_s = t_i \) to express \( E_A \) as

\[ E_A = \sin^2 \theta_n^* / \sin^2 \theta_c \]

Figure 25 shows a plot of this efficiency versus the induction time. There appears to be a strong correlation with particle size, as is evident from (46).

The two efficiencies can now be combined to arrive at the collection efficiency \( E_c \), see equation (40). A graphical demonstration of the collection efficiency as a function of particle size is given in Figure 26. To construct this Figure, the induction time was assumed independent of bubble and particle size, which is not valid as was observed in Figure 22. The peak in Figure 26, however, corresponds well to the expected maximum in size-by-size recovery data [68]. A smaller bubble implies a larger collection efficiency, because both \( E_c \) and \( E_A \) increase. The latter is greater due to the fractional decrease in particle sliding velocity on a smaller bubble exceeding the fractional decrease in sliding distance.
Fig. 24. Tangential velocity gradient at surface of sphere at Re = 290 and θ = 45°.

The model for the collection efficiency was derived for a single bubble particle collision. Dobby accounted for the interaction of a bubble swarm by changing the surface vorticity \( \Phi \) to incorporate the gas holdup \( \epsilon_g \). The relationship thus acquired is

\[
\Phi_{eg} = \Phi + n \epsilon_g
\]

where \( \Phi_{eg} \) is the surface vorticity of a gas bubble at gas holdup \( \epsilon_g \), and \( n \) is a function of Re, and angle \( \theta \). An increase in holdup results in a larger vorticity, and thus a larger \( E_x \), because of an increase in \( u_b \). However, to produce a larger holdup the gas rate must be
increased, which has a larger $d_b$ as result. The flotation rate constant is related to the bubble diameter as [75]

$$k (:) \frac{d^{1.5}}{d_b}$$

and to the collection efficiency $E_x$ according to

$$k (:) \frac{v_g E_x}{d_b}$$

(47)

where $v_g$ denotes the gas velocity. Consequently, the flotation rate may pass through a maximum as the gas rate is increased. It increases because of a larger holdup and thus larger $E_x$, but is subsequently reduced because a larger $d_b$ will negatively affect $E_x$, as is seen in Figure 25.

In conventional flotation cells the probability mechanism does not entirely explain the recovery of fine particles. Trahar also found that ultrafine particles (<5-10 micron) are mechanically entrained in the froth. Their recovery is linearly related to the recovery of water. Column flotation offers a possible mechanism to minimize this effect, as is discussed in chapter 2.8.
1.3.2. Flotation considered an art.

Flotation was conceived and implemented early this century. The first distinct departure from the old ore dressing practice of amalgamation and concentration by vanners and blankets was the cyanide process. Use of a weak solution of potassium cyanide to dissolve gold was patented first in 1887 by McArthur and Forest [76]. The process was combined with mineral dressing methods as it required solutions to slime separation problems (from sand, and from pregnant liquor). The discovery that recoveries could be enhanced by grinding finer, saw an evolution in separation techniques, including the invention of flotation. Many men who pioneered in flotation came from cyanide plants.

Reagent types were added by trial, not because their fundamental action was understood. From the mid 1960’s onwards, first attempts were made to control the flotation process in a plant. The principle tool used as reference was the on-stream analyzer. Desired grades and calculated recoveries were controlled primarily by changing reagent addition rates. A real optimizing control model lacked, however, and guide lines for control logic were
absent. Nevertheless, operators changed setpoints during shifts and just after shift change. The question is, were these alterations made randomly or by feeling that the process was positively influenced [77].

In reality, one might sum up a hundred variables that prevail upon the flotation process. These include ore effects, grinding and classification conditions, feed grades, mineral specific gravities, liberation, hydrophobicity, pulp density, bubble size, air, reagents, temperature, viscosity, flotation equipment to mention just a few. Laboratory flotation tests are conducted with the objective to maintain most variables constant. Other variables are introduced, specifically froth removal. As it is done by hand, it requires a well experienced and steady person to ensure frequency and paddle depth. Flotation tests are planned "on the basis of experimental results with less consideration to theory than most other metallurgical processes require" [6].

Some researchers made their own cells to produce high concentration ratios. Specific objectives in the research to flotation fundamentals can only be achieved by adapting an apparatus. The Hallimond tube, a glass container equipped with a gas fritter and an outlet to collect particles that floated to the top, is used to study flotation chemistry; The INCO-voltammetric cell [78] was developed to relate the impact of conditions during liberation to mineral hydrophobicity. Mineral electrodes are polished in situ in the test electrode, and allowed to attain their rest potential after aeration. Standard Denver laboratory flotation machines handle mixtures of minerals of varying size. Tests with this type of apparatus are conducted for reproducibility and maximum recovery. Improvements can be made through automation, such as is done for the Leeds laboratory cell, which removes froth continuously.

Results from those tests, however, are difficult, if not impossible, to scale-up to plant scale. Some seasoned flotation engineers add reagents to batch tests on demand, making reproducibility questionable. Batch tests are also done at lower pulp density, due to pulp transfer from grinding mill to cell. It will be shown in chapter 1.4.2. that pulp density may have a dramatic effect on flotation response. Ore preparation for these tests often differ from actual plant practice.

Batch test operators are guided by unquantifiable observations such as froth colour and froth fluidity across the cell lip. The pH is not always recorded continuously, but measured sporadically. It is no wonder that flotation has been frowned upon as an art and was not viewed as a science. Small scale flotation units like the Leeds cell, the Noranda mini-cell and the Outokumpu mini-pilot plant were invented to overcome cell level problems. Other more sophisticated equipment to measure the dissolved oxygen content and the redox potential throughout the test have only been introduced fairly recently. Yet no on-line measurement of bubble size is available for the conventional flotation cell, although it is being tested on small and large flotation columns. As was shown in the previous subchapter, the size of a gas bubble determines the rate of recovery of the various particle sizes.

The philosophy of flotation in a plant often is analogous to that in the laboratory. Traditionally, each operator has his own concept of how a plant should be operated. Conditions are altered right after shift change. On occasions operators artificially attain high concentrate grades and high recoveries in their shift at the expense of a build-up of
circulating load. This practice is most apparent where multiple cleaning stages are concerned. Just before shift change the tailings valve is opened for a few minutes to release the build-up, which goes unnoticed if done between sampling. Also, if on-stream analysis data, or results of assays from samples taken on the same shift, indicate to poor metallurgical results, more often than not the reagent dosage is changed. This should not necessarily be the first step, as the problem may lay in insufficient aeration or pulp level. It is easier, however, to make one change at the reagent distribution system, than manipulating the air valve of many cells.

Steps taken in the middle sixties to optimize flotation were the first attempts to move away from the art. Nevertheless, these control procedures were arrived at by trial and error, and were not based on fundamentals. More sophisticated instrumentation, the computer age, and application of mathematical models, coupled with techniques learned in the chemical and hydrometallurgical fields should help to improve the image of the mineral dressing engineer as a scientist, and not as a black art performer.

1.4. Philosophy and science in mineral dressing

In the last two decennia mineral dressing has undergone a fast evolution in terms of equipment, control and automation. This evolution is strengthened by a better scientific understanding of ores (quantitative mineralogy), the actual processes employed (flotation, magnetic separation) and their control and automation (on-line sensors and computers). The drive to produce metals at lower costs to maintain a competitive edge forces a company to do 'something'. Personal experience has shown that not always the correct approach is taken at first instance to streamline an operation at the most economical way. For example, at Cuajone a hundred million dollar expansion plan was tabled for plant expansion, required to offset falling head grade. It included one additional secondary crusher, two tertiary crushers, a ball mill and replacement of 168 flotation cells, with 8.5 m$^3$ volume each, by 66 cells that were nearly quintuple in volume to raise the throughput from 47 to 55 ktonnes/day. This plan fell through in the light of costs and a flailing Peruvian economy. At Falconbridge's Strathcona Mill the replacement of 252 - 2.8 m$^3$ rougher and scavenger cells with twenty 38.2 m$^3$ cells would provide opportunity for process control to deal with the fluctuations in grade and recovery. Installation of instrumentation for fewer units would make process control more manageable.

Similar philosophies are maintained at various concentrators with which I have been in contact. The bottom line is that quite often a decision is made towards costly capital investments before analyzing a current operation. Instead, a managerial philosophy may be identified, which is suggested to be followed. This procedure is illustrated in Figure 27. It advocates the knowledge of process capacity and maximum utilization before considering high capital investment. This can be achieved through a rigorous research and development program, which considers modern technology and recent knowledge acquired at plants worldwide. The plan must be rational, take an incremental approach, incorporate a cost-benefit analysis for each step, and must view the desired final plant lay-out.
MAXIMUM UTILIZATION OF INSTALLED AVAILABLE EQUIPMENT

START AT THE BEGINNING OF THE PROCESS

Determine physical capacity of unit process

Create stability in unit process

Establish maximum/optimum under controlled conditions

Enhanced physical changes (to optimize circuit geometry)

Once debottlenecked, step to higher level in control hierarchy

Continue to next unit process

Expansion involving high capital costs

OPERATOR

TRAINING AND LEARNING AND

SHIFT FOCUS OF OPERATOR AND INSTRUMENTATION ENGINEER

in House Engineers

Consultants

People

Optimal Use

OF KNOWLEDGEABLE

Fig. 27. Managerial philosophy to maximize the utilization of installed equipment

Description of the managerial philosophy

Start at the beginning of the process. If crushing is not the bottleneck, any modification there may be of lower priority, especially since most concentrators feature fine ore storage before grinding. Improvements in a grinding circuit are likely many: Reliable sensors and instrumentation equipment have been developed in the last decennium. Digital based systems or multi-processors allow custom tailoring of controllers to decouple loops, overcome control variable dynamics, and use higher forms of control including cascade, feedback, adaptive control, and ramping techniques. Successful application to plants worldwide has demonstrated that improvement through control is feasible. The approach to implementation is

- install a few sensors strategically on test circuit
- monitor the process, involve operators
- develop and apply a stabilizing control strategy
- debottleneck test circuit
- install more sensors to complete strategy
- implement strategy and train operators
- expand to cover complete circuit
Debottlenecking a circuit to determine the physical capacity will involve modifications of the smaller material handling equipment. The philosophy here is that the most expensive piece of a circuit should be the restriction.

Once the circuit is stable, i.e. fluctuations are controlled, determine maximum and optimum operating ranges. A subsequent unit process may not be able to handle adequately the maximum operating conditions: a straight increase in grinding capacity will have deleterious effects on the recovery in flotation. Yet a temporary sacrifice should be agreed to in order to establish the boundaries of the operation. After establishing that the throughput of the grinding circuit can be increased, but at the expense of flotation recovery, subsequent studies in the flotation circuit may demonstrate that lost territory can be recouped. This will be the subject of Chapter 1.4.2.

Flotation, usually subsequent to grinding when treating sulphide ores, can be dealt with next. Besides eliminating physical constraints, a much improved process will ensue when attention is given to historical data and when trending of process parameters becomes available. Chapter 1.4.2. stresses the importance of data recording which can result in better reagent distribution and dosage, optimize pulp density, better pulp chemistry, and upgrade regulatory control to stabilizing control.

**Implementation of control**

In house expertise should be complemented with the right outside consultant to provide focused training for the process engineer, and to bring in experience from elsewhere through which goals can be reached faster. On the other hand, there should be a continuous flow of selected information to the operating personnel accompanying each of the implementation steps described in Figure 27. When made part of the team, bottom up involvement and commitment is ensured. They will realize that streamlining a plant through process control incorporates a shift in focus of their work, rather than elimination. A similar shift will affect the work of electrical and instrumentation engineers, who traditionally controlled the instrumentation hardware and software. Their involvement will be more as a service group in troubleshooting and assistance to the process engineer, who will now decide on control strategy and implementation.

Various levels can be distinguished in the control hierarchy. The levels and interpretations considered here may deviate from that found in literature as these better explain the functionality. They are:

- regulatory control - Level 0
- stabilizing control - Level 1
- supervisory control - Level 2
- optimizing control - Level 3
- expert control - Level 4
- management information systems - Level 5
Regulatory control is found in all plants. It consists of a sensor and an actuator that provide control over a single parameter. As an example may serve the signal of a belt scale, monitoring mill feed. This process variable is checked against a manually inputted setpoint on a controller, which in turn sends a signal to the variable drive of a feeder. The great number of 'isolated' controllers would not permit close observation of all parameters all the time.

Stabilizing control ties several regulatory control systems through techniques such as feedback, feedforward, and cascade: The setpoint of a feed controller may be regulated by the output of a size control station.

Supervisory control is attained when several constraints apply in a stabilizing control system. For example, a grinding circuit throughput that is varied automatically to satisfy a desired product size, may be frozen upon an impending mill overload. This signal could originate from a drastic change in measured rate of change in mill power. Chapter 2.2 illustrates various forms of constraints that could be incorporated in a control strategy.

![Diagram](image)

**Fig.28.** Building blocks of stabilizing and supervisory control

Figure 28 illustrates the coherence of basic building blocks to provide stabilizing or supervisory control. Distributed Control (DCS) and Programmable Logic Control (PLC) find their own application, the former aids in process control, the latter in motion control. A mistake sometimes made is not to distinguish between the two. All a DCS should know is whether a pump functions; the PLC will look after individual valves and flushes on the
pump. The philosophy behind Statistical Process Control (SPC) is important for any process control project. A process needs to be controlled within certain limits. When exceeded, an explanation must be found with assurance given to future prevention. Process control is possible only if

- trained people are available
- good housekeeping is practised before equipment installation
- equipment is designed well, and is well maintained
- there is good understanding of the cause of the process

Level 3 control pertains to plant management or optimizing control. At this level the system continuously searches for an optimum, which may be based on economics. The economic efficiency is maximized by manipulating the circuit along the grade-recovery curve. This exacts tight control measures within, for example, the four quadrants that can be identified bordering this relationship [79]. Wherever the metallurgical results are in either of the four areas (Figure 29), the objective of optimizing control is to steer the circuit towards the direction of the optimum. Similar strategies were devised by Outokumpu [80].

Fig.29. Optimizing control of the grade-recovery curve [79].
I consider expert control the highest level in control. The expert system is personal-computer-based. Based on signals received from the field, the computer evaluates the information using rules obtained from field experts. An enhanced version of expert control is a learning classifier system, where the computer develops rules for use in the expert system. A different approach is the use of 'on-line mathematical modelling'. Field signals are processed by comparison with complex mathematical models, using Kalman filters for example, derived from historical data to predict the correct response the control system should give.

Each level of control must be built on the previous level to provide a strong foundation. If not, the failure of control at a particular level may result in complete failure.

1.4.1. Control in the grinding circuit

The previous chapter alluded to the distinction between process control and automation or motion control. Motion control replaces routine jobs with a programmed sequence of signals that emulate physical actions. It creates a convenience and should shift the attention of personnel to higher level tasks. Automation generally is not a big income generator, rather a cost reducing measure. Examples of automation are pump start-up/shutdown and switching cyclones.

Process control, on the other hand, ensures that parameters are maintained within predefined limits. Continuous monitoring, setpoint checking and correctional action dispenses stability in a circuit with opportunity of operation under optimum conditions. Absence of fluctuations may be translatable in a better product for similar input. Process control, therefore, generates income.

An example of a successful implementation of a grinding control strategy is described in Chapter 2.2. Before the project was completed, several bottlenecks in the circuit required attention. The objective was to operate the most expensive piece of equipment, the mill, at its limit. Some of the changes involved are related in Chapter 2.1., and pertained to raising the cyclone cluster several decimeters, enlarging the cyclone overflow bin, and increasing the speed of the cyclone feed pump. Eventually, the cyclone cluster was replaced by a row of horizontal cyclones to accommodate a higher throughput. It becomes evident from the mill power charts when a mill is operated near its limit. At that point, frequent impending overloads impose a freeze on tonnage control through the size loop.

The major contribution of any control is the resulting stability of operation. Compare Figures 8 and 10 of Chapter 2.1. to note the stabilizing effect of pump box level control. For any operation it is relatively inexpensive to have an ultrasonic level measurement device installed, and to link it in with pump box make-up water. A steady level in the pump box reflects in steady cyclone operation. The noticeable drift in Figure 10 in Chapter 2.1., with respect to product size and density, can only be eliminated by additional expense. Improving regulatory feed control and manual overflow density control by integration of size control produces long term stability, as may be seen from Figure 11 in Chapter 2.1. Particle size is the most sensitive parameter: controlling size in most cases implies a controlled
overflow density. Table 7 summarizes data obtained at Falconbridge's Strathcona Mill. The standard deviations of size and density have been reduced from 2.7 to 0.7 and from 2.7 to 1.9 respectively. One may also observe that a slightly higher throughput accompanies these results. Knowledge of recovery as a function of size would provide the economic justification for control.

<table>
<thead>
<tr>
<th></th>
<th>FIXED SIZE VARIABLE THROUGHPUT PUMP BOX LEVEL CONTROL</th>
<th>FIXED THROUGHPUT PUMP BOX LEVEL CONTROL</th>
</tr>
</thead>
<tbody>
<tr>
<td>THROUGHPUT</td>
<td>181.2</td>
<td>180.8</td>
</tr>
<tr>
<td>STD. DEV.</td>
<td>5.2</td>
<td>4.3</td>
</tr>
<tr>
<td>RANGE</td>
<td>174.8 -187.5</td>
<td>176.4 -185.3</td>
</tr>
<tr>
<td>SIZE</td>
<td>54.2</td>
<td>52.5</td>
</tr>
<tr>
<td>STD. DEV.</td>
<td>0.7</td>
<td>2.7</td>
</tr>
<tr>
<td>RANGE</td>
<td>52.3 -56.1</td>
<td>48.9 -56.1</td>
</tr>
<tr>
<td>DENSITY</td>
<td>42.8</td>
<td>44.4</td>
</tr>
<tr>
<td>STD. DEV.</td>
<td>1.9</td>
<td>2.7</td>
</tr>
<tr>
<td>RANGE</td>
<td>40.1 -45.5</td>
<td>40.7 -48.2</td>
</tr>
</tbody>
</table>

to recover tonnes/h  

Table 7. Grinding circuit throughput, product size and density at Strathcona Mill, comparing operation at fixed tonnage with operation at fixed size.

The installation of equipment like a particle size monitor, and the configuration of a control strategy is relatively simple regardless of what forms the control basis. Many examples can be found in the literature. It is the effective manipulation of the constraints where the challenge lies. A few constraints are mentioned in Chapter 2.2. Of these the ball mill power constraint deserves some more attention.

Using the scan rate of the modules in the DCS, the difference in ball mill power is calculable in Kw/min. For a mill in equilibrium this value would always be equal to zero. This is never the case, however, because of the high frequency of the readings. In addition to the rate of change the absolute power also is an important number. One may imagine that, depending on the power value, a temporary negative rate of change is tolerable. For example, if the mill power is 2.1 Mw, then no harm is done if the rate of change of power is -5 Kw/min; for a mill power of 1.8 Mw it would be dangerous if the power would continue to drop. With a function generator in the DCS an admissible rate value can be configured for any actual power value, which can serve as a soft constraint. Once the
measured rate of change of mill power matches the soft constraint, the size station enters a tracking mode and the feed rate is frozen. Two more power rate limits at predetermined lower fixed values exist, the hard and the critical constraint. On reaching the hard constraint the feed rate is reduced at a certain rate, for example 0.25 tons/sec. This value is increased to 1.00 tons/sec once the critical constraint is realized. The gradual drop in throughput guarantees that equilibrium is maintained, i.e. no major upsets occur. When the impending mill overload is cleared, size control is restored only after the power rate of change is positive and the mill power is beyond 2.0 Mw in this case.

A rod mill in a grinding circuit creates the concern of tangling rods. Brenda Mines developed a method based on vibration analysis [81]. The maximum throughput of the circuit is updated continuously, by recalculation from present mill vibration values. This method does not take into consideration a desired product size. If size were important in their circuit, one may perceive a strategy whereby the throughput, which is dictated by size control, is compared to the maximum allowable throughput calculated from vibration analysis. At Brenda, this strategy resulted in a 10 % increase in tonnage.

Fig.30. Potential grinding control strategy for the Strathcona Mill at Falconbridge.

Falconbridge operates a rod mill-ball mill circuit. Mill power and rod tangling are not the only concerns, but changes in mineral content of the ore must also be considered. In most cases a softer ore is accompanied by a higher metal content. If grinding control were based on size control, the flotation cells would need to handle more concentrate because of more
throughput and because of the higher feed grade. Flotation may become unmanageable, especially when operated manually. The additional constraint required pertains to a maximum production rate, or better, a maximum allowable rate of change of production. As soon as an upper limit is reached, the size setpoint may then be automatically lowered gradually. Figure 30 may illustrate how the final control strategy at Falconbridge could look like.

Technology has advanced to the point that measurements of size, density, mass flow, tonnage, and level are known to be reliable. Calibration of sound, vibration and rheology measurements are more difficult. Sufficient information is available, however, that stability and some optimizing control can be attained through direct measurement. Prediction models that utilize either regression analysis or more sophisticated techniques are used at various locations. An example of the latter approach is the multi-variable model-based optimal control method in which the 'states' of the system are estimated, typically by a Kalman filter [82].

In my opinion prediction models should only be used once plant personnel have a basic understanding of the principles of grinding. Such understanding may have been acquired through experience with a distributed control system, which should be available in any modern grinding circuit. At a minimum, sufficient equipment should be available to support stabilizing control. Gradually, the focus on work may become readjusted, especially when size control is adopted. Rather than checking overflow density, which becomes secondary in importance, more attention must be devoted on mill discharge density and cyclone underflow characteristics. The optimum density range in a mill is quite narrow, typically between 3 or 4 %. An apparent impending mill overload often is the result of high pulp viscosity. In practice a lot of training proved necessary to accustom working habits to new technology.

McIvor [32], Chapter 1.2.1., advocates a detailed sampling campaign to conduct a functional performance characteristics study to optimize a grinding circuit. Results of the study may incorporate physical changes in the plant. This method cannot be implemented on a regular circuit as he claims. For composite samples taken over a few hours to be meaningful a grinding circuit must be stable. Instruments can measure that stability. His proposal should be regarded as an optimization step (Level 3), to be performed once stabilizing control has been established. Similar considerations should be given to circuit fine tuning, with respect to optimum ball charge, optimum ball diameter, and viscosity modifiers.

1.4.2. Flotation Circuit Optimization

There is a trend to move away from the black art concept in flotation. Science is making its introduction especially in the design, and to a lesser degree, in the operation of a plant. Simulation techniques and control strategies make their inroad, but do not take into account all parameters adequately. In plant operations the tendency exists to select large, capital intensive equipment before a thorough analysis of its capacity has been conducted. To deal with a possible expansion or improve an existing operation, it requires a managerial philosophy to apply the fundamentals of science. Knowledge of the current circuit may save
the company money. Trending and data logging are to be encouraged in flotation operations.

Optimization in design

The design of flotation circuits so far has been primarily based on past experience. While designing a circuit for a new sulphide ore, the common tendency is to choose from one of the circuits listed by Lynch [83], or a variation thereof, based on experiments. It would certainly be more desirable to base such decision on a quantitative design criterion as none of the traditional circuits may result in the best performance. A flotation circuit is designed optimally when between interconnected banks of cells there is a minimum of circulating loads. In addition, such operating conditions have been determined that the best flotation performance is attained.

Simulation studies now undertaken predict a performance under the assumption that flotation is a first order process. For \( M(k,t) \), the mass of solid species remaining in the cell with overall rate constant \( k \) at time \( t \), the rate of removal of this solid from the pulp can be expressed for a batch test as

\[
dM(k,t)/dt = -k \cdot M(k,t)
\]

(48)

The concept of first order kinetics was expanded to continuous operation where the residence time in the system is distributed. Considering each cell as a perfect mixer [84], generalized mass balance equations may be set for each component in the feed, reporting to concentrate and tailings. The mean residence time is a function of operating conditions such as particle size, pulp density, and agitator speed. Of the many physical parameters involved Agar [85] found the residence time as probably the most critical. He defined three interrelated criteria from which the optimum flotation time can be selected, once the pulp chemistry has been defined. These criteria involve maximizing the difference in recovery between the desired mineral and gangue, maximizing separation efficiency, and prevention of material of lower grade than feed to enter the concentrate. In later work [86] he established a simulation technique based on locked cycle tests.

Other simulation techniques have been summarized elsewhere [87]. With these a flotation plant might be built that consists of an adequate number of cells. One may still have to resort to piloting to determine proper reagent dosage and pulp chemistry. Also neither the effect of pulp density nor of froth behaviour have been considered properly in the simulation techniques listed in this recent reference.

Optimization in operation

Most of the control strategies written for flotation base a supervisory control algorithm directly on questioning proper reagent levels [79,80,88]. These show little reference to pulp
level control [80], and none to air or pulp density control. Fully integrated supervisory control has not yet been implemented with success to date.

Despite this, the use of a computer or a micro processor in flotation should be encouraged. As in the grinding circuit, it will provide stability through individual regulatory control loops that may be linked to a localized supervisory control. Available data can be presented in the form of trends, increasing the understanding of interdependent process variables. Data logging offers the process control engineer a historic data bank to be used in statistical processing of process information. This first step can be followed by defining the real capacity of installed equipment.

Two simple examples may illustrate the use of historical data at Cuajone. After multilinear regression of various parameters, the losses of copper to tailings were found to be strongly dependent on the iron content in the feed. Results of these analyses indicated that especially an increased dosage of the secondary collector (sodium isopropyl xanthate) enhanced the recovery of middlings that could be liberated further in the subsequent regrind stage. A relation between iron content and secondary collector was developed whereby the dosage could be raised two- or threefold, which resulted in at least one percent better recovery of copper. A similar study demonstrated the benefits of splitting frother dosage between head and mid section of rougher flotation.

Pulp density

One parameter that received little attention is pulp density. M in equation (48) can be assumed to constitute the sum of a floatable mass M₁, and a non-floatable mass M₂. Furthermore, assign V as the volume of pulp in the cell, V₁ as the volume of froth delivered in unit time and V₂ the volume of tailings delivered in unit time. Then the concentration of total solids in the cell is [89]

\[
\frac{(M_2 + (1-\alpha)M_1)}{V_2}
\]

Considering all the solids in the cell the rate of recovery R is

\[
-dM(k_2t)/dt = R = \frac{\alpha \cdot M_1}{V} = k_2 \cdot \left[\frac{(M_2 + (1-\alpha)M_1)}{V_2}\right] \tag{49}
\]

where \( \alpha \) represents the fractional yield of floatable material and c equals 1 for a first order reaction. \( k_2 \) is a rate constant. This equation implies that the rate of recovery is a function of pulp density. In a continuous test to alter the percentage of floatable material in the pulp may thus be achieved by altering the feed pulp density. Jowett et al. [89] generated a linear relationship between R and solids concentration in the pulp. As \( \alpha \) remained fairly constant throughout the tests the rate equation was concluded to be of first order with \( \alpha \) depending only on the effective residence time, \( t_e \), of increments of pulp in the cell:

\[
\alpha = 1-e^{-k \cdot t_e}
\]
The conclusion drawn from this work was that, as long as floatable material is present in the pulp in sufficient quantities to impart a good chance of collision and cohesion in the froth, this material can be recovered. No information was given on the size distribution of the coal used in the tests. Imaizumi and Inoue [90] concurred with the conclusion that flotation follows first order kinetics, but found anomalies at high pulp densities. For test work on a group of uniform particles in the system equation (48) may be rewritten as

\[-\frac{dw}{dt} = k_1 \cdot w\]  

(50)

where \(w\) is the total weight of the specified particles with rate constant \(k_1\). Integration of (50) gives

\[y = \frac{w}{w_0} = e^{-k_1 \cdot t}\]  

(51)

\(w_0\) is the weight of the specified particles in the pulp at time zero. The expected straight line plot of log \(y\) against time showed an upward convex tendency, which indicated to a deviation from first order flotation kinetics. The authors explained this phenomenon in terms of over crowding at the surface of the air bubble. In contrast of the findings of Jowett there seems to be an upper limit in pulp density. A more appropriate general equation for the linear flotation was derived:

\[y = \int_0^\infty W(k_1) \cdot e^{-k_1 \cdot t} \, dk_1\]  

(52)

\(W(k_1)\) represents the distribution function of the flotation rate constant \(k_1\) for a group of particles uniform in size.

Tomlinson and Fleming [91] characterized two types of flotation: free flotation, which is described by equation (48), and "inhibited" flotation. The latter occurs at very high density of floatable material and the rate of flotation becomes dependent only on the rate of aeration. This implies that all air bubbles are saturated with particles with a load determined by surface capacity or lifting power. Equation (48) effectively becomes

\[-\frac{dM}{dt} = k_3 \cdot N = \text{constant at unchanging air rate}\]

\(N\) is the number of bubbles per unit volume at time \(t\). Heavily mineralized bubble-particle aggregates become immobilized, and merge with other aggregates. Through coalescence the new total surface area will be less and the overall buoyancy increases due to a net loss in lifted weight. The coarsest size fractions are mostly affected. It is theorized that when collision occurs the particles attached on the colliding faces of the bubbles experience a free fall between the bubbles. The surface tension of the bubbles are insufficient to reattach the larger grains. Indeed, inhibited flotation of coarse particles was observed to occur at dilute suspensions.

Frew [92] related rate constants \((k)\) upon pulp density \((p)\) after a two year survey in a plant. He described the behaviour with the following exponential relationship

\[k = \alpha \cdot e^{-\beta \cdot p}\]

\(\alpha\) and \(\beta\) are constants

(53)
If this were true for each size fraction $W(k)$ in equation (52) could be replaced by (53) and the equation would revert to

$$y = \int_{0}^{\alpha} e^{-k \cdot t} \cdot e^{k \cdot t} \, dk$$

(54)

Test work at Cuajone confirmed the dependency on pulp density. As Figure 31 attests, a reduction by a few percent solids in the pulp has a significant impact on the floatability of coarse size fractions. The recovery of grains coarser than 208 microns improved by at least 10%. In this case it more than compensated for a loss in retention time as a result of treating a more dilute pulp. The Figure includes observations in a large 38.2 m$^3$ flotation unit as well as in a small 2.3 m$^3$ Galligher unit.

Fig.31. Copper recovery as function of particle size and pulp density. (Cuajone data)

This discussion stresses the importance of testing an existing circuit thoroughly before making a expensive decision to replace equipment. More dilute pulps could imply more cells than originally planned. One more factor to consider is the redox potential. For better selectivity or better grade-recovery relationships it may become important to incorporate conditioning time in a circuit. A separate chapter (Chapter 1.4.3.) is devoted to redox potential.
1.4.3. Redox potential

Traditionally, pH has always been one of the main flotation modifiers to ensure selective separation of minerals. Control of pH at a specific value has given adequate results for ores of relatively simple and homogeneous structure. In 1934! Wark and Cox (Figure 17) illustrated the dependence of flotation of pyrite, galena, and chalcopyrite as a function of pH. Researchers have investigated the relation between pH and zeta-potential. Ye et al. [93] found that the IEP (iso-electric point) pH values for several coals were close to the pH at which the smallest induction time is observed for the corresponding coal sample. The induction time is defined as that time required for the disjoining water film between a hydrophobic surface and a gas bubble to drain to such thickness that rupture of the water film takes place. The smaller the induction time, the faster the flotation response. For oxide minerals, whose surfaces are probably hydroxylated in the presence of water, there is considerable experimental evidence that surface potentials are controlled by the activation of $\text{H}^+$ and $\text{OH}^-$.

Flotation may be relatively effective at a certain pH value, or within a certain range, but results may not be optimal. Despite controlled pH conditions, redox values may alter due to oxygen uptake or increased collector dosage. Sipila [94] illustrated observations at Outokumpu in Figure 32.

![Diagram](image)

Fig.32. Changes in redox potential in a pH-controlled flotation test.
Fig. 33. Shift averages of Cu recovery and potential measurements at Cuajone, taken every hour.

Potential measurements at the head of rougher flotation at Cuajone showed that values may change hourly (Figure 33b). When averaged over the duration of a shift there appeared to be an inverse relationship between measured potential and overall plant copper recovery (Figure 33a).

The first to suggest that in flotation an electrochemical mechanism takes place on the surface of minerals were Salamy and Nixon [95]. This mechanism involves the simultaneous anodic reaction of the collector and cathodic reduction of oxygen:

$$2\text{ROCS}_2^- \rightarrow (\text{ROCS}_2)_2 + 2e^- \quad (35)$$

$$\frac{1}{2}\text{O}_2 + 2\text{H}^+ + 2e^- \rightarrow \text{H}_2\text{O} \quad (36)$$

In Chapter 1.3.1, ROCS$_2^-$ was symbolized as X$^-$ for simplicity. Reaction (35) describes the oxidation of xanthate to dixanthogen, but the electrochemical approach is equally applicable
to chemisorption reactions [96] and to reactions in which the surface itself is a reactant. The anodic reaction may then be described by for example

\[ \text{PbS} + 2\text{ROCS}_2^- \rightarrow \text{Pb}(\text{ROCS}_2)_2 + \text{S} + 2e^- \]  \hspace{1cm} (55)

In the field of corrosion the concept of surface reactions where conductive materials are involved are well established to proceed in separate steps. Pourbaix [97] created an atlas in which stability regions are displayed for various metals and their compounds as function of pH and Eh. Garrels and Christ [98] extended these diagrams to minerals, but regions refer to compounds that are thermodynamically stable in the geological time sequence. Other investigators have adapted these diagrams to stability domains of various species in a solution comprising of collectors and other anions. The position of phase boundaries in an Eh-pH diagram, i.e. the size of a stability domain, is a function of concentration of various species. Such diagrams for several flotation systems are readily available in the literature [99,100,101,102,103]. Figure 34 illustrates the diagram for a galena-xanthate system for a xanthate concentration of 10^{-4} M. Equilibrium lines are calculated using thermodynamic data for bulk solutions. The PbS to PbX_2 line involves thiosulphate formation according to the reaction:

\[ 2\text{PbS} + 4\text{ROCS}_2^- + 3\text{H}_2\text{O} \rightarrow 2\text{Pb}(\text{ROCS}_2)_2 + \text{S}_2\text{O}_3^{2-} + 6\text{H}^+ + 8e^- \]  \hspace{1cm} (56)

This reaction is favoured on thermodynamic grounds at pH values relevant to flotation. Toperi and Tolun [104], however, measured rest potentials following along the equilibrium line for reaction (55), as indicated in Figure 34. These results suggest that the reaction producing thiosulphate proceeds too slow to be significant; The formation of a hydrophobic surface seems to be effected by (55).

Indiscriminate use of Eh-pH diagrams has thus been shown to lead to erroneous assumptions regarding the stable species. Diagrams must be drawn for a preassigned value of concentration of various solution species. In an actual flotation system such concentrations are not ordinarily known. Another drawback is that prior knowledge of the metastable species in a system is required. The original diagrams were calculated for bulk thermodynamic data and did not take into account the formation of submonolayer quantities on mineral surfaces. Their thermodynamic properties can differ significantly from those of the bulk phases.

To gain better understanding of the reactions involved, flotation kinetics and equilibria were studied. Changes in composition and in the hydrophobic or hydrophilic character of the mineral surface are caused by the anodic reactions. The potential across the solid/solution interface determines the identity and rates of these reactions. It becomes important to determine which are the hydrophobic entities through which flotation takes place. Metastable species must be taken into consideration because the mineral concentration process eventuates within a short period, in contrast to the geological time frame.

Recent Eh-pH diagrams are composed by computer, and include metastable species at various concentrations of reaction product. Figure 35 [105] shows examples for stable
copper sulphides (covellite (A), djurleite (B), and chalcocite (C)) and for chalcocite with metastable copper sulphides (D). The dashed lines in the Figures represent reaction product concentrations of $10^7$, $10^5$, and $10^3$ mol. An evenly distributed concentration of $10^5$ mol would constitute a monolayer coverage across the surface. Woods compared these diagrams with the results he obtained in linear potential sweep voltammetry. This method increases or decreases the potential systematically in linear steps at a fixed scan rate. Despite several disadvantages associated with this method, such as lack of identification of reaction product and mass transfer dependency on scan rate, studying anodic and cathodic currents that occur at a potential value may lead to significant information. For example, in the presence of some djurleite in a chalcocite electrode the mineral started to reduce at about -0.35 V at pH 9.2 on the negative-going scan. A peak was reached at -0.5 V. With reference to either Figure 35(C) or 35(D), reduction should not commence before about -0.55 V, peaking at -0.75 V. At these conditions, however, djurleite starts to reduce as may be observed in Figure 35(B).

Early theories on the mechanism for the interaction of xanthates with sulphide minerals lead to the adsorption theory by Wark [106]. Xanthate was considered to adsorb in the double layer at the mineral-solution interface. Woods [107] concurred that xanthate can be in the double layer of platinum, but maintained that it would not affect the hydrophobicity. He theorized that the strong attraction of water molecules by the charged polar head groups
Fig. 35. Eh-pH diagrams for (meta)stable copper sulphides [105]
influenced neighbouring CH$_2$ groups. As a result, their contribution to the hydrophobic effect is diminished. Appraising the electrochemical reactions as suggested by Salamy [95], the total potential drop from a surface to solution, E, may be viewed as the work done in transporting a unit charge across the double layer into the electrode. This is an important parameter for electrode reactions, since such processes involve transfer of electrons between dissolved species and the electrode. The zeta-potential is the potential difference across the diffuse double layer. For the same value of E, the magnitude and sign of the zeta-potential can vary, as was discussed in Chapter 1.3.1. It becomes unnecessary to measure the zeta-potential as a quantity; more important is to know the general way in which a particular solution parameter changes the surface charge from, say, negative to zero to positive.

The first step in flotation of PbS with xanthate is chemisorption of the collector [107], which renders the surface slightly hydrophobic. This process involves the transfer of an electron to the mineral. Once a near monolayer has formed, further oxidation to dixanthogen takes place and is adsorbed physically to make the surface hydrophobic. A mixed potential mechanism is required to produce a rest potential on the electrode above the xanthate/dixanthogen equilibrium line. As Salamy and Nixon inferred, this can only happen in the presence of oxygen to allow oxygen reduction at the electrode surface (reaction (36)). The presence of oxygen is a condition for flotation to take effect.

The role of oxygen has also been explained in terms of purely chemical reactions by an adaptation of Taggart’s mechanism. The mineral reacts with oxygen first for proper surface conditioning before reacting with xanthate.

$$2\text{PbS} + 2\text{O}_2 + \text{H}_2\text{O} \rightarrow \text{Pb}_2\text{S}_2\text{O}_3 + \text{Pb(OH)}_2$$

$$\text{Pb}_2\text{S}_2\text{O}_3 + 2\text{ROCS}_2^- \rightarrow \text{Pb(ROCS)}_2 + \text{S}_3\text{O}_3^{2-}$$

With regard to bulk chemistry, these reactions are thermodynamically stable. It has been shown, however, that sulphur does not oxidize to thiosulfate immediately, but only after extended polarization.

**Collectorless flotation**

A study by Finkelstein [108] sparked interest in researching the mechanism responsible for collectorless flotation. His work showed that a variety of minerals was amenable to flotation without use of a collector. Strong wetting forces, however, do not take place at a clean sulphide mineral-water interface because of the inability of sulphides to participate in hydrogen bonding with water. Heyes and Trahar [109] contested stating that natural floatability of clean particles is observed in a slightly oxidized environment. Chalcopyrite does not float under reducing conditions, but flotation response may be restored entirely when returning to an oxidizing environment, with the exception of coarse particles. The mineral surface produces a hydrophobic entity by superficial oxidation. It seems reasonable to assume that this is the elemental sulphur formed according to the following general reactions:
\[
\text{MS} \rightarrow \text{M}^{n+} + S + n\text{e}^- \quad (57)
\]
\[
\text{MS} + n\text{H}_2\text{O} \rightarrow \text{M(OH)}_n + S + n\text{H}^+ + n\text{e}^- \quad (58)
\]

Reaction (57) would occur under acidic conditions, while reaction (58) would take place in neutral or alkaline solutions. Techniques available at the time were deficient in correlating the amount of surface S\textsuperscript{+} to the floatability of for example chalcopyrite [108], nor could S\textsuperscript{+} be extracted from the surface by acetone [109]. Results of detailed electrochemical studies combined with photovoltage measurements [110] were consistent with the formation of elemental sulphur.

For galena reaction (57) would produce a soluble lead ion below pH 7. Cyclic potential sweep voltammograms confirmed a shift in the anodic current at the potential where reaction (57) should take place upon stirring the solution. Excess sulphur on the electrode is expected to react with hydrogen ions to H\textsubscript{2}S, which again was confirmed as voltammograms showed a higher current in a stirred solution. Chalcopyrite was found to behave in a similar manner [111] with respect to an initial oxidation reaction that involves the removal of iron from the mineral lattice. The voltammetric behaviour was interpreted in terms of copper and sulphur products of surface oxidation being CuS and elemental sulphur:

\[
\text{CuFeS}_2 + 3\text{H}_2\text{O} \rightarrow \text{CuS} + \text{Fe(OH)}_2 + S + 3\text{H}^+ + 3\text{e}^- \quad (59)
\]

Equation (59) is identical to (58), which is written in general notation. This reaction proceeds as a function of pH. A change in pH should cause a shift of the flotation edge, with formation of S at -0.102 V, rather than 0.547 V which corresponds to the above reaction [112].

Voltammetric studies of pyrite and pyrrhotite also gave evidence to the formation of S\textsuperscript{+} [113]. Oxidation to higher oxy-sulphur ions succeed at higher anodic potential.

Yoon's investigations [114] suggested that hydrophobicity can be provoked in a reducing environment without the need to form elemental sulphur. The addition of strongly reducing sodium sulphide did not depress chalcopyrite. Rather, it removed hydrophilic oxidation products from the mineral surfaces such as sulphates and hydroxides. As a reducing agent, Na\textsubscript{2}S lowers the redox potential on the mineral surface, but as long as the potential is above the flotation edge potential of about -100 mV no depression should be apparent. His conclusions were not entirely correct because the flotation tests were conducted with air. Despite the presence of Na\textsubscript{2}S, air will raise the potential sufficiently high to allow sulphur formation on the surface, which induces flotation. In later work [115] Yoon confirmed that slightly oxidizing conditions are required indeed. Fuerstenau [116] provided experimental evidence that all common base metal sulphides are naturally floatable in oxygen-free environment. One of the main reasons leading to this apparent disagreement among researchers may be the difference in surface state, which includes the degree of surface oxidation, surface roughness, and surface defects. This surface state has not been given adequate attention. Not only chemicals in the pulp, but also conditions of preparation may alter the state.
Electron spectroscopy for chemical assays (ESCA), also called X-Ray photo electron spectroscopy (XPS), failed to detect elemental sulphur on the surfaces. Based on peaks in the $S_{2s}$ and $S_{2p}$ spectra, indicating binding energies that corresponded to a sulphide sulphur, a copper polysulphide was proposed as the hydrophobic entity. Others [117] confirmed this observation. For each mineral investigated (galena, bornite, chalcopyrite, and pyrrhotite) the initial oxidation reaction involved the removal of a metal atom from the surface, which then formed a hydroxy-oxide when exposed to air or in alkaline solutions. In acid solutions the metal atoms formed soluble ions. The result is a metal-deficient sulphide layer with a composition outside the range of the established stable phases used in the original $Eh$-$pH$ diagrams.

Analogous to this thinking the reaction for chalcopyrite under basic conditions differs from (59)

$$\text{CuFeS}_2 + 3\text{H}_2\text{O} \rightarrow \text{Fe(OH)}_3 + \text{CuS}_2 + 3\text{H}^+ + 3\text{e}^-$$

The metastable $\text{CuS}_2$ contains less negative free energy than the two stable phases $\text{CuS}$ and $\text{S}$. Formation of the metastable product will occur at a higher potential than when calculating the reaction on the basis of the two stable compounds. This reveals the relative insensitivity of voltammograms: the potential at which a reaction begins to occur cannot be defined precisely, and a difference of 1 kJ/mol in $\Delta G$ relates to only a 3 mV shift in reversible potential. It falls beyond the resolution of the voltammogram to discern the reaction potentials of $S^+$, polysulphides or metal-deficient sulphides.

Under acidic conditions, the oxidation reaction of chalcopyrite originally proposed resembles (57)

$$\text{CuFeS}_2 \rightarrow \text{CuS} + \text{S} + \text{Fe}^{2+} + 2\text{e}^-$$

XPS data point to a metastable surface layer of approximate composition $\text{Cu}_{0.8}\text{S}_2$. The initial oxidation step involves partial dissolution of copper as well as removal of iron from the surface. Iron forms hydrophilic ferric hydroxides in situ which inhibits flotation. However, matters are more complicated when chalcopyrite was observed to be floatable at much more cathodic potential values through oxidation of a sulphide rather than by oxidation of the surface [118]. The exact identity of the deposited layer is not known.

The natural floatability of pyrite could be induced by the addition of sodium sulphide [119] and was pronounced at concentrations of $\text{Na}_2\text{S}$ of $10^{-3}$ to $10^{-2}$ M. Sulphide ion was suspected to hydrolyse to $\text{HS}^-$ which then reacted with the mineral surface to form $S^-$ according to

$$\text{HS}^- \rightarrow \text{H}^+ + S^- + 2\text{e}^-$$

Cyclohexane was used to leach the pyrite surface and the presence of sulphur was detected in a corresponding UV spectrum. This presented sufficient evidence for the researchers to state that $S^-$ is the hydrophobic entity. XPS studies [120] refuted this work and the
formation of a metal-deficient sulphur layer was restated as the hydrophobic entity. Reaction (58) for pyrite was written as [113]

\[ \text{FeS}_2 + 3\text{H}_2\text{O} \rightarrow \text{Fe(OH)}_3 + 2\text{S} + 3\text{H}^+ + 3\text{e}^- \]

but should be reformulated to

\[ \text{FeS}_2 + 3x\text{H}_2\text{O} \rightarrow x\text{Fe(OH)}_3 + \text{Fe}_{1-x}\text{S}_2 + 2\text{S} + 3x\text{H}^+ + 3xe^- \]

Thermodynamic data for the metastable phases of pyrite are not available in the literature. Reformulation of (58) is defended on the basis of complete reduction of the oxidation products on the cathodic sweep. Each product of this reaction is insoluble. Besides, when a metastable compound is formed the surface structure remains unaltered. If oxidation products indeed were \( \text{Fe(OH)}_3 \) and \( S^+ \), then on reduction an iron monosulphide of pyrrhotite properties would be formed

\[ \text{Fe(OH)}_3 + S^+ + 3\text{H}^+ + 3\text{e}^- \rightarrow \text{FeS} + 3\text{H}_2\text{O} \]

The second anodic potential sweep appeared unaltered from the first, giving reason to belief that the pyrite structure remained intact.

**Metal-deficient sulphur layer or polysulphide formation?**

At this point it should be remarked that Buckley et al. [120] observed a metal-deficient sulphur layer on the minerals, using XPS techniques. Elemental sulphur was noticed after exposure of the minerals to strongly oxidizing conditions [117]. Luttrel and Yoon [115] explain the sulphur species on chalcopyrite as a polysulphide \( (\text{S}_x^{2-}) \), where \( x \) represents the number of sulphur atoms within the chain structure. Polysulphide ions may be held strongly to the surface by the following mechanism

\[ 2\text{Cu}^+ + S_x^{2-} \rightarrow \text{Cu-S}_x\text{-Cu} \]

This mechanism suggests that adsorption is considered to take place on surface copper sites. The two sulphur ions at both ends of the \( S_x^{2-} \) chain would tend to approach a valence of -1, whereas those in the middle would assume an elemental form \( (S^+) \). There is no significant difference in the resultant product layer formed as a metal polysulphide or metal-deficient sulphide. At least two types of sulphur bonding would be expected for the polysulphide model, however. These are sulphur to metal and sulphur to sulphur bonds. For air-exposed galena XPS studies [117] detected just a single type sulphur species. On metal electrodes polysulphides are likely to form when treated with Na₂S. Evidence of polysulphides was found on gold electrodes in alkaline solutions [121]. Deposition of elemental sulphur on pyrite and chalcopyrite in the presence of Na₂S is possible [119]. But again, XPS spectra of these layers discerned a structure slightly different from a \( S_8 \)-layer [122].
On the assumption that sulphur oxidation does not proceed beyond the formation of polysulphides, sulphur distribution diagrams were constructed for alkaline solutions of pH 8, 10, and 12 [123]. Figure 36 reproduces these diagrams, in which elemental sulphur, thiosulphate, polythionate and sulphate are absent. It appears that the onset of collectorless flotation corresponds closely to the critical Eh where polysulphides are formed. The higher the potential, the longer the chain of this molecule. Consequently, the mineral becomes more hydrophobic as the number of elemental sulphur atoms in the middle of the chain outweigh the effect of the charged sulphur ion at its end. This would help explain the increase in floatability with increase in potential in alkaline solutions.

![S.H.E. Potential Diagrams](image)

Fig.36. Sulphur distribution diagram as function of Eh and pH [123].

With this concept in mind, the Eh-pH diagrams of Figure 35 were constructed not considering sulphur species with oxidation states above sulphur(0). Polysulphides were included because their metastable domains under alkaline conditions possess a state between sulphur(-2) and sulphur(0).

Peters [124] approaches the oxidation of a mineral from a leaching point of view. Under leaching conditions, a sulphide mineral is presumed to relinquish electrons to an oxidant. The electrochemical leaching rate, as well as the mixed potential, is determined at the cross-over point (Figure 37) of the anodic and cathodic processes. This infers that the leaching rate is determined entirely by the physical and structural nature of the mineral surface, and by the electrochemical potential of that surface. When copper sulphides leach, the solid reaction products on the mineral surface are fractured and fissured due to a decrease in volume. When pyrite, with a molar volume of 12 cm³ per mole of contained sulphur, oxidizes anodically, there will be a volume increase to 15.5 for rhombic or 16.3 cm³ for monoclinic sulphur formed at the surface [125]. The result is an unfractured sulphur coated pyrite surface, which passivates the mineral. Consequently, pyrite requires an overpotential to leach, as has been observed by several investigators. This also concurs with
XPS spectra that reveal elemental sulphur [120], although evidence of a metal-deficient sulphur layer in the initial stages of oxidation was apparent.

In a later paper [126], Peters suggested that the postulation of a rhombic sulphur layer on the surface of pyrite may be too naive. The reaction analogous to (57) would stop immediately as sulphur forms a protective layer. Such layer would be too thin to impede passage of electrons because the cathodic reduction of oxygen on pyrite continues despite the layer. This can be deduced from the fact that the presence of pyrite promotes galvanic reactions on less "noble" minerals. Detection of the layer is impossible by spectroscopic methods. Creation of the necessary vacuum, common in surface investigations, would destroy the layer. Yet, after a pressure leach, pyrite may be floated from the pulp due to a coating of sulphur. If such layer would not exist, he concludes, then at least a metal-deficient sulphide layer must have developed.

![Diagram](image)

Fig. 37. The cross-over point of the cathodic and anodic reactions determines both the electrochemical leaching rate as well as the mixed potential [124].

For the minerals chalcopyrite and pyrrhotite, their molar volumes were larger than that of sulphur [125]. About 85% or more of the mineral sulphur was noted to have converted to the elemental form. Peters concluded that S²⁻ cannot be oxidized unless the overpotential applied is beyond the value required for the Fe²⁺/Fe³⁺ redox couple.

To create Eh-pH diagrams that include the destabilization of the sulphate equilibrium lines, the standard free energies were shifted by 75 kcal/mol. Figure 38 illustrates resulting diagrams for Cu₂S and CuS, which may be compared directly with those in Figure 35(C) and 35(A). The only difference is the sulphate formation at high potential values above pH 5. In an alkaline medium copper sulphides oxidize to species of lower sulphur content at potentials below which elemental sulphur would be formed during the oxidation process of
some mineral sulphide ions to sulphate. Perhaps these diagrams should exhibit lower oxy-sulphur species than sulphate. Yet an overpotential would be required regardless.

Fig.38. Eh-pH diagrams obtained by shifting standard free energies of the sulphate equilibrium lines by 75 kcal/mol [125].

Use of xanthate as collector

Collectorless flotation proved feasible for almost any mineral. When placed in an ore environment, however, mutual interference of minerals may effect flotation results negatively. To elicit adequate flotation response a collector like xanthate must be added to the pulp [127]. Doubt remains whether it contributes to the hydrophobicity of the mineral surface through, for example, the formation of a collector coating. One possibility may be that xanthate counteracts the hydrophillic effects of the metal hydroxides that are formed on the surface. A recent study [128] also demonstrated that collectors are required to enhance the recovery of copper. Table 8 addresses the fact that collectorless flotation, including the effects of sulphidization, frother and particle size, could account for only 50 to 80 % of sulphide copper recovery.

Collectorless flotation of chalcopyrite from a pulp also containing pentlandite and pyrrhotite is possible without loss in recovery. In fact, the resulting grade and recovery curve compared favourably to standard flotation conditions. Figure 39 shows the comparison and includes several repeat runs. Except for the lack of the xanthate collector, test conditions were similar to those maintained in the standard tests. Collectorless flotation involved a pre- aeration period when necessary to raise the potential to a suitable level (265 mV, SHE). The potential in the case of flotation with collector was in the same range, about 285 mV.
For a pH between 8 and 9.5 the response in collectorless flotation was identical, but dropped off to the curve obtained for flotation with collector when the pH was raised above 10.5. Collectorless flotability of pentlandite (Pn) is substantially lower (recovery < 30%), than that of chalcopyrite under similar conditions with respect to redox potential. Grinding under nitrogen nearly doubled Pn recovery, while treatment with sodium sulphide after grinding under air tripled it to 78 %.

**Copper Recovery %**

<table>
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<tr>
<th>ORE</th>
<th>SULPHIDE</th>
<th>OXIDE</th>
<th>TOTAL</th>
<th>SULPHIDE</th>
<th>OXIDE</th>
<th>TOTAL</th>
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<td>71.2</td>
<td>84.1</td>
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Table 8. Flotation with and without collector at -500 mV (SIE) [128].

Fig. 39. Collectorless flotation of chalcopyrite from a pulp containing nickel minerals. Tests conducted at Falconbridge.
Richardson and Maust [110] considered the drop in potential, effected by xanthate addition, could be contributed to the formation of insoluble lead xanthate. This reduces the lead ion concentration in solution, displacing the equilibrium potential of the reaction

$$\text{PbS}^- \rightarrow \text{Pb}^{2+} + \text{S}^+ + 2e^-$$

to more cathodic values. Pritzker and Yoon [129] offered an alternative for this reaction: At a potential of -50 mV, PbS is oxidized to Pb(OH)$_3$ and S$^+$ at pH 10.5, which is beyond the critical pH for flotation. At this pH, S$^+$ is formed at a faster rate than for any lower pH value. They believe possibly elemental sulphur, rather than xanthate, renders the surface hydrophobic.

The chemical adsorption of xanthate is generally considered to attach to a metal atom on the sulphide surface. Reference was made to xanthate bonding with surface sulphur atoms [96]. Compounds, such as ROCS$_2$S-S$_2$COR are quite stable and cannot be distinguished from dixanthogens by spectroscopic analysis.

It may be perceived that the hydrophobicity of a mineral is a complex mechanism, involving the formation of a hydrophobic metal-deficient surface. The hydrophobicity may be enhanced by the presence of xanthate possibly adsorbing to the sulphide compounds on the surface. Xanthate renders flotation possible at lower cathodic potential values than is likely without collector. Formation of hydrophobic metal xanthates occurs on first instance, followed by dixanthogen formation. These may partly prevent the formation of hydrophillic hydrides, either because of lowering the potential or because less sites will be available for hydroxide formation.

**Interaction of minerals in flotation**

Laboratory flotation of clean minerals individually suggest selective flotation can be attained using proper redox potential ranges. Figure 40 compares the recovery of pyrite within a certain Eh range to the recoveries of several copper minerals [130]. At +200 mV it should be possible to float all the chalcopyrite away from free pyrite. Figure 41 is a similar graph pertaining to chalcopyrite and galena in the absence or presence of a collector [112]. Between +400 and 600 mV chalcopyrite should float preferentially, and below -100 mV the galena. Unfortunately, these minerals interact when ground together as is made illustrative in Figures 42(A) and 42(B). The flotation of galena is modified by interaction with copper ions produced when chalcopyrite oxidizes:

$$\text{Cu}^{2+} + \text{PbS}^- \rightarrow \text{Pb}^{2+} + \text{CuS}$$

This reaction is driven by favourable kinetics. Vice versa, lead ions are considered to have a detrimental effect on chalcopyrite by replacing iron in the chalcopyrite lattice:

$$\text{CuFeS}_2 + \text{Pb}^{2+} \rightarrow \text{CuPbS}_2 + \text{Fe}^{2+}$$
Fig. 40. Flotation responses of copper minerals as function of Eh [130].

Fig. 41. Floatability of chalcopyrite and galena at pH 8 in presence and absence of collector [112].
A similar effect was noticed for a mixture of pyrite, chalcopyrite and quartz when floated in the absence of a collector. A highly oxidized pyrite sample provided a reservoir of dissolved ferrous ions that caused precipitation of hydrophilic hydrous ferric oxide on the surface of chalcopyrite. If these cations are complexed, then depression may be prevented.

Fig. 42. Flotation response of galena/chalcopyrite/quartz mixtures in reducing and oxidizing conditions [112]

Interaction with grinding

Fundamental studies on single minerals in the sulphide system may be misleading. Flotation responses not only depend on the interaction between minerals due to hydroxide precipitation or possible lattice replacement. Galvanic interactions may influence the potential on the surface of minerals, such as is the case for pyrite present with other minerals. In acidic solutions galvanic coupling with pyrite increases the rate of formation of elemental sulphur on the surfaces of galena, sphalerite and chalcopyrite [131].

Galvanic coupling was also demonstrated between pyrrhotite and steel balls in a grinding mill. The intimate contact between mineral particles and balls results in currents of a magnitude related to the contacting ball material [132]. Rest potential measurements, summarized in Table 9, show that due to the galvanic coupling pyrrhotite will act as the more noble electrode when in contact with mild steel or stainless steel balls. The more active the steel, the higher the current. Mild steel was most active. Bubbling oxygen through the pulp causes the current to rise by an order of magnitude. It accelerates the electrochem-
ical reactions on the surface of pyrrhotite, which results in lower floatability. After a 20 minute contact with austenitic stainless steel and mild steel without abrasion, the recovery

<table>
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<th>$E_{ASS}$ (mV SCE)</th>
<th>$E_{MSS}$ (mV SCE)</th>
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<tbody>
<tr>
<td>0.5 M NaCl</td>
<td>6.7</td>
<td>-174</td>
<td>-156</td>
<td>-300</td>
<td>-640</td>
</tr>
<tr>
<td>air exposed</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Distilled water</td>
<td>6.8</td>
<td>-190</td>
<td>-126</td>
<td>-542</td>
<td>-760</td>
</tr>
<tr>
<td>($N_2$ bubbling)</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Distilled water</td>
<td>4.0</td>
<td>-30</td>
<td>-94</td>
<td>-350</td>
<td>-618</td>
</tr>
<tr>
<td>(air exposed)</td>
<td>6.8</td>
<td>-85</td>
<td>-38</td>
<td>-146</td>
<td>-580</td>
</tr>
<tr>
<td></td>
<td>10.0</td>
<td>-130</td>
<td>-234</td>
<td>-332</td>
<td>-486</td>
</tr>
<tr>
<td>Distilled water</td>
<td>6.6</td>
<td>-72</td>
<td>+28</td>
<td>-178</td>
<td>-420</td>
</tr>
<tr>
<td>($O_2$ bubbling)</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

Table 9. Rest potentials of pyrrhotite and steel electrodes under non-abrasive conditions.[132]

of pyrrhotite dropped from 94 % to 85% respectively 74 %. In situ precipitation of hydrophillic iron hydroxides competed with the formation of a polysulphide or metal-deficient sulphide. Equation (58) describes the reaction that took place. Even after breaking contact between the mineral and the metal the floatability could not be restored.

Rest potentials on pyrrhotite in contact with different steel balls under abrasive conditions were measured as well. In contact with mild steel the rest potential for pyrrhotite was the most cathodic as is shown in Table 10 [133]. Balls made from this material produce more debris, which is thought to deposit on the mineral as it assumes a low cathodic potential value that approaches that of mild steel itself. When both pyrite and pyrrhotite are brought in contact with mild steel simultaneously, the situation becomes more complex. Flotation results in a Hallimond tube, after contact with mild steel for 120 minutes, showed that the recovery of pyrrhotite dropped from 90 to 23 %. But in the presence of pyrite, the recovery dropped to 70% [133]. Apparently, when short circuited in a three-electrode combination, the cathodic roles of the sulphides are reduced.
Rao et al. [134] measured the mixed potentials developed by different sulphide mineral electrodes in xanthate solutions in the presence and absence of a metallic iron electrode. When there was no galvanic contact between the sulphide and the iron, the mixed potential values reached the xanthate/dixanthogen redox potential. In contact with iron, however, the mixed potential became more cathodic and too reducing for a xanthate collector to be adsorbed effectively. This would bring up the question whether collectors should be added to the grinding circuit. To restore the floatability completely, removal of iron particles and hydroxide precipitate must precede. Heyes and Trahar ground chalcopyrite in an inert mill under strongly reducing conditions and could restore the floatability after raising the potential, except for the coarser size ranges [109]. Metallic debris were not produced under these circumstances.

### Table 10. Rest potentials for pyrrhotite in contact with steel under abrasive conditions [133]

<table>
<thead>
<tr>
<th>ELECTRODE</th>
<th>GRINDING BALLS</th>
<th>REST POTENTIALS V (VS SCE)</th>
</tr>
</thead>
<tbody>
<tr>
<td>PORCELAIN</td>
<td></td>
<td>- 0.05</td>
</tr>
<tr>
<td>ASS</td>
<td></td>
<td>- 0.24</td>
</tr>
<tr>
<td>PYRRHOTITE</td>
<td></td>
<td></td>
</tr>
<tr>
<td>HCLA</td>
<td></td>
<td>- 0.32</td>
</tr>
<tr>
<td>MILD STEEL</td>
<td></td>
<td>- 0.69</td>
</tr>
<tr>
<td>ASS</td>
<td>PORCELAIN</td>
<td>- 0.26</td>
</tr>
<tr>
<td>HCLA</td>
<td>PORCELAIN</td>
<td>- 0.55</td>
</tr>
<tr>
<td>MILD STEEL</td>
<td>PORCELAIN</td>
<td>- 0.70</td>
</tr>
</tbody>
</table>

ASS: austenitic stainless steel
HCLA: high carbon low alloy steel

Plant experience

Most grinding circuits exhibit very cathodic redox potentials, due to the very nature of the grinding balls used. Oxygen uptake after grinding is very rapid and the potential rises. More than likely the potential to be measured in the flotation section is not ideal, but could be optimized. This can be effected by use of reducing or oxidizing reagents, or by maintaining a reducing gas atmosphere in flotation.
The presence of Basaltic Andesite in the Cuajone ore body traditionally posed problems to plant metallurgy. When floated separately, the maximum recovery that could be attained ranged from 60 to 65%. It could constitute up to 20% of the mill feed. The major difficulty with this ore type appeared to be the intricate intergrowth between pyrite and chalcopyrite. Conditioning with Na₂S improved flotation results by removing hydrophillic hydroxide precipitates, but danger of overdosage persisted. To relate the dosage of Na₂S to a measurable parameter, measurements of redox potential values were commenced.

A preliminary plant survey was conducted to monitor current potential values in various areas of the Cuajone concentrator. Measurements were made in both the sands and the slimes rougher flotation sections, in the combined grinding overflow stream, in the fresh and reclaim water sources, and in the copper thickener overflow. This thickener treats the molybdenite plant tailings, which is the final copper concentrate. The results are summarized in Figure 43. The presence of NaCN and Na₂S in the copper thickener overflow, both added during the flotation of molybdenite, resulted in a low potential value. Except for the fresh water source, most streams show similar tendencies in their fluctuations. Shift averages of total plant copper recovery are illustrated separately in an attempt to correlate recovery to potential. In general, a high recovery seems to correlate with a low potential value. Figure 33 already showed follow-up results of measurements

<table>
<thead>
<tr>
<th>SAMPLE</th>
<th>FLOTATION MAKE-UP WATER FROM THICKENER OVERFLOW (%)</th>
<th>E (MV vs SCE)</th>
<th>RECOVERY (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td></td>
<td>Cu</td>
</tr>
<tr>
<td>DAY 1</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Rougher₁ standard</td>
<td>0.0</td>
<td>30</td>
<td>84.5</td>
</tr>
<tr>
<td>Rougher₂</td>
<td>0.6</td>
<td>10</td>
<td>85.6</td>
</tr>
<tr>
<td>Rougher₃</td>
<td>1.5</td>
<td>0</td>
<td>88.5</td>
</tr>
<tr>
<td>Rougher₄</td>
<td>3.0</td>
<td>-15</td>
<td>85.7</td>
</tr>
<tr>
<td>Rougher₅</td>
<td>4.5</td>
<td>-40</td>
<td>85.5</td>
</tr>
<tr>
<td>DAY 2</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Rougher₁ standard</td>
<td>0.0</td>
<td>65</td>
<td>82.4</td>
</tr>
<tr>
<td>Rougher₂</td>
<td>0.6</td>
<td>60</td>
<td>82.3</td>
</tr>
<tr>
<td>Rougher₃</td>
<td>1.5</td>
<td>55</td>
<td>82.4</td>
</tr>
<tr>
<td>Rougher₄</td>
<td>3.0</td>
<td>45</td>
<td>82.7</td>
</tr>
<tr>
<td>Rougher₅</td>
<td>4.5</td>
<td>30</td>
<td>83.2</td>
</tr>
</tbody>
</table>

Table 11. Pulp potential controlled by copper thickener overflow. Summary of recovery results of bench tests at Cuajone.
made hourly. Controlled laboratory tests, using thickener overflow to maintain the redox

![Graph](image-url)

**Fig. 43.** Daily potential measurements of various process streams and copper recovery in rougher flotation and for the plant (Cuajone).
PORTION OF THICKENER OVERFLOW DIRECTED TO:

<table>
<thead>
<tr>
<th></th>
<th>HEAD OF ROUGHER BANK</th>
<th>MIDDLE OF ROUGHER BANK</th>
</tr>
</thead>
<tbody>
<tr>
<td>E (mV vs SCE)</td>
<td>COPPER RECOVERY</td>
<td>MOLY RECOVERY</td>
</tr>
<tr>
<td>60</td>
<td>85.6</td>
<td>87.5</td>
</tr>
<tr>
<td>30</td>
<td>84.2</td>
<td>87.9</td>
</tr>
<tr>
<td>0</td>
<td>88.1</td>
<td>88.9</td>
</tr>
<tr>
<td>-30</td>
<td>87.4</td>
<td>----</td>
</tr>
<tr>
<td>-60</td>
<td>85.7</td>
<td>86.9</td>
</tr>
</tbody>
</table>

Table 12. Results of Cuajone plant tests, controlling pulp potential with copper thickener overflow.

Fig. 44. Eh-pH diagram for the Fe-Cu-Mo-S-O-H system [98]
potential, supported these observations. The results summarized in Table 11 indicate to an optimum of 0 mV. Actual plant results are given in Table 12, and were obtained by pumping part of the overflow to either the head or midsection of a rougher flotation bank.

In a simple Pourbaix diagram for the chalcopyrite-pyrite-molybdenite system, a work area can be indicated for the Cuajone process (Figure 44). Although the diagram does not refer to metastable phases, a metallurgist should be aware when to expect oxidation of copper minerals. This diagram may explain the disadvantage of a too anodic potential, but does not offer a reason for depression at the lower potential range applicable to Cuajone.

This case study shows that effective control of a redox potential may be practical, despite the view of many investigators. The success of potential control at Cuajone was assessed as a four percent increase in recovery, at the expense of little capital.

1.5. Column Flotation versus Conventional Flotation

1.5.1. Column Flotation: Principle and Control

Klassen and Mokrousov [138] probably were the first to report that improved drainage of gangue particles from the froth was possible by sprinkling water on top of the froth. Boutin and Wheeler [139] capitalized on this idea and claimed to have built the first commercial column at Opemiska, Canada. Major problems were encountered with the sparger system. During the 1970's, Wheeler continued his efforts, but the first breakthrough came in 1981. Columns successfully cleaned molybdenite concentrate at Les Mines Gaspé, Canada [140], replacing a row of sixteen 1.4 m³ cleaner cells by a 1.8*1.8*12 m column. One column replaced four stages of cleaning. In March 1983, McGill University in Montreal held a seminar on columns, which brought this technology to the attention of the industry. Especially the successful installation on bulk copper-molybdenite cleaning at Gibraltar, British Columbia [141] had North America convinced. In one stage a final concentrate grade is attained, while the column tailings are scavenged to maximize recovery. Cleaner cells typically have a unit cell volume of 1.4 to 2.8 m³. One column, depending on its size, may possess a volume equivalent to 15 to 40 conventional cleaning cells of 2.8 m³ each. Cuajone, Peru operates two units in series, each 3.1 m in diameter and approximately 9 m high; Curragh Resources in the Yukon, Canada, recently commissioned a 3.7 m unit with a length of 13.5 m. A few cells with a diameter of 4.9 m have come on-stream in Mantos Blancos, Chile, in December 1989. Since then columns were tested and installed in numerous applications, including in the industrial minerals industry [135].

With the interest in columns different versions of the apparatus were conceived. Examples that found industrial application, are the packed column [142], the Jameson high-intensity column [143], and the Bahr cell [144].
Description of the column

A schematic diagram of the column flotation cell is included in Chapter 2.4, Figure 1. Feed slurry enters about one third the way down from the top and descends against a rising swarm of bubbles generated by the sparger arrangement. The counter-current flow pattern offers better attachment probabilities than the concurrent mode of operation as employed in conventional cells. Floatable particles are collected by the rising bubbles, hence this zone is defined as the collection zone. At the pulp-froth interface the particles are transferred into the froth, conveniently called the cleaning zone.

Washwater is the key feature in column operation and is added from the top. Care should be taken to have the water enter the froth as a sprinkle. A jet spray would break up the desired plug flow rise of the bubbles. Two methods of water distribution are in vogue. Most installations chose for the submerged water distributor; others have an overhead tank with perforated bottom, located just above the froth. The latter provides better surface coverage, visual inspection on-line, and avoids solids build-up from concentrate. To its disadvantage is water short circuiting to the overflow, as may be appreciated viewing Figure 9, Chapter 2.8. For a 0.91 m diameter column, the quantity of excess water was calculated at about 40 to 60 L/min. A second concern is that water may not penetrate the froth sufficiently in case of a well loaded froth. Short circuiting is prevented with a submerged distributor design, which must be performed such that a minimum of surface area is blocked. This characteristic is especially important when carrying capacity limits are approached. A further dispute is the depth at which this distributor should be installed. From Figure 8, Chapter 2.8, washing may be seen as becoming less effective the greater the depth. In this case the nickel content in the copper concentrate rose from 0.55 to 1.0 %.

Chemical screen analyses identified the problem as increased entrainment of pentlandite particles smaller than 30 microns.

Spargers are located near the bottom of the column, although some installations are known to feature a 'booster' sparger assembly, just below the feed inlet. Bubble generation mechanisms are of two types, internal spargers and external bubble contactors. Internal spargers are made from perforated pipe clad usually in perforated rubber or fabric. The US Bureau of Mines [135] developed the first external bubble contactor which was improved upon by Cominco Ltd. Other external devices are the one marketed by Deister [136] and the one tested by Virginia Polytechnique [137]. The Bahr cell also mixes air and pulp externally through a series of ports along the bottom periphery of the cell. A detailed description of a cloth-type sparger is given in Chapter 2.9.

Column dimensions

An alternative division of the various zones in the column could be, from top to bottom, the cleaning zone, the recovery zone, and the scavenger zone. The cleaning zone still is the froth zone, as defined earlier. The recovery zone is the volume between the pulp-froth interface and the feed inlet. A need to describe this zone separately has risen in earlier papers, because the action between the pulp-froth interface and the feed inlet is not completely understood. Wheeler [179] referred to this intermediate zone as the washing zone; in Chapter 2.3 this zone was named the transition zone, as its volume appeared to
dictate the overall recovery. The interaction between the collection and cleaning zones was recognized by Falutsu [145,146] to influence overall flotation performance, and was expressed in terms of froth drop back. Froth drop back was defined as that fraction of solids entering the froth but is returned to the collection zone. It should be noted here that when a particle drops back from the froth in a column, it retains 100% of its original retention time to be picked up by a bubble. This improves separation, but decreases column capacity per unit volume. For a bank arrangement of cells, the rejected particle in cell 2 and on is always less than the original retention time.

The flotation column may be visualized as a network of connected well-mixed zones, with provision for flow of the phases involved from one zone to all the adjacent ones. An illustration of the back flow mixing model is given in Figure 45 [147].

**Backflow mixing model**

Representation of a Flotation Column

![Diagram of backflow mixing model](image)

**Fig.45.** Back flow mixing model of the N-zones-in-series for the representation of the flotation column.
The material balance for the tracer in the i-th zone is:

\[ V_i \cdot \frac{dC_i}{dt} = Q_{i-1} C_{i-1} + Q_{i+1} C_{i+1} - (Q_{i,i+1} + Q_{i,i-1}) C_i \]  \hspace{1cm} (60)

where \(i,i+1\) denotes flow from the \(i\)th to the \((i+1)\)th cell, \(V_i\) is the zone volume, \(C_i\) the zone concentration of the tracer, and \(Q\) the symbol for flow. The two parameters that need to be solved are \(N\), the number of zones in a column, and the back flow \(Q_{i,i-1}\). It is not possible to geometrically separate the zones in a column; for convenience the back flow mixing was expressed in terms of the back flow ratio \(\beta\)

\[ \beta = \frac{Q_{i,i-1}}{Q_i} \]

where \(Q_i\) is the net downward liquid flow in a column, and \(Q_{i,i-1}\) is the back flow. These authors fitted the model to an experimental RTD (residence time distribution) test and found \(\beta\) to lie in between 1 and 16 (Figure 46). A large \(\beta\) value denotes a more or less intense recirculation of the liquid contents in the column. It was noted that \(\beta\) seemed to reach a limiting value as the number of zones got closer to 10. This observation was correlated to that by Joshi and Sharma [148], who found that in bubble columns the recirculating liquid establishes a pattern of loops, each one having a height of approximately equal to the column diameter \(D\). The limiting height of transfer unit (HTU) being equal to \(D\) then dictates a column height to diameter ratio of close to 10.

Fig. 46. Fitting of the back mixing model to the experimental RTD curve [147].
A similar conclusion was reached by Yianatos for feed flow rates up to 600 L/min [149]. The two parameters N and B are combined into the single dimensionless Peclet number, $P_{e}$, which traditionally is used to describe the mixing process through axial dispersion coefficient $E_{p}$

$$P_{e} = 2N/(1 + 2B) \quad \{P_{e} = 1/N_{p} = (u_{p} + u_{l})H/E_{p}\} \quad (61)$$

where $u_{p}$ represents the particle settling velocity, $u_{l}$ the interstitial liquid velocity, H the height of the column, and $N_{p}$ the particle dispersion number. The dispersion coefficient $E_{p}$ for large diameter columns is related to the column diameter D by [150]

$$E_{p} = 0.063D(J_{g}/1.6)^{0.3} \quad (62)$$

where $J_{g}$ is the superficial gas velocity. The last term in equation (62) may be ignored for $J_{g} < 0.03$ m/s. Combination of (61) and (62) yields

$$N_{p} = [0.063/(u_{p} + u_{l})] \cdot D/H \cdot (J_{g}/1.6)^{0.3} \quad (63)$$

The effect of various D/H ratios on the grade recovery curve is demonstrated in Figure 47.

![Figure 47. Effect of H/D ratios and feed flowrate on column performance [149]](image-url)
The maximum height to diameter ratio is determined by the gas carrying capacity, since the volumetric gas rate decreases with increase in H/D. The concept of carrying capacity will be discussed later.

It should be noted here that the dispersion coefficient \( E_p \) becomes larger when the column, or the baffles, are vertically misaligned. For a tilt over angle \( \alpha \) the modified dispersion coefficient \( E_\alpha \) is given by [151]

\[
E_\alpha = E_p (1 + 1100\alpha D)^2
\]

(64)

where \( E_p \) is calculated from equation (62), and \( D \) is the column diameter.

**Column performance**

Important variables that have a bearing on the performance of the column are the gas and water rates and the gas holdup. These parameters are described in the following subchapters.

**Wash water rate**

The wash water rate \( J_w \) comprises of two terms, the bias rate \( J_g \), and the split of water to the overflow \( J_{cw} \). Each is expressed in terms of superficial rates, which is the volume flow \( Q \) divided by the column area \( A \)

\[
e.g. J_w = \frac{Q_w}{A}
\]

The split of wash water between bias and overflow depends on several factors and is difficult to predict. An equation for the superficial flow rate of water to the overflow \( J_{cw} \) is [152]

\[
J_{cw} = J_g \cdot \left\{ \frac{(1 - \varepsilon_{g0})}{\varepsilon_{g0}} \right\}
\]

(65)

where \( J_g \) is the superficial gas rate, and \( \varepsilon_{g0} \) the air gas holdup on the top of the froth. A lower gas holdup in the froth zone would result in more water reporting to the overflow.

The superficial bias rate \( J_g \) is defined as the difference between tailings water flow rate and feed water flow rate. When this difference is positive, usually referred to as positive bias, there exists a net downward flow of water. Tracer techniques have shown that when a positive bias prevails, less than 1 % of the feed water will report to the concentrate [150]. Reduction or elimination of feed water entering the froth zone would prevent entrainment of fine particles into the concentrate. Conventional cells operate under negative bias, since the water in the feed is split between concentrate and tailings. Table 1 in Chapter 2.8 attests that especially the minus 37 micron particles flow with the water to the concentrate. Entrainment has been studied elsewhere [68].
On the other hand, an increase in the bias rate results in a drop in gas holdup in the upper part of the froth, as may be noted in Figure 48 [153]. Holdup values of collection and cleaner zones approach, making the profile flatter. Increased mixing and channelling will short circuit feed water, entrained as far as the well mixed upper portion of the froth, to the overflow. In this case this occurred at $J_g > 0.4 \text{ cm/s}$, at $J_g > 2 \text{ cm/s}$.

![Diagram](image)

Fig.48. Schematic illustration of the effect of increasing bias water on the gas holdup profile [153]

Espinoza [154] studied the effect of bias rate on the copper concentrate grade, retreating copper at Mt. Isa, Australia. Over the range studied ($0 < J_g < 0.35 \text{ cm/s}$), he did not observe a change in grade with increasing $J_g$. A plot of copper grade in the concentrate versus the bias rate is shown in Figure 49. In fact, an increase in bias will reduce the residence time through an increase in $u_t$ considering the following relationship [149]

$$u_t = Q_t H / \{V_c (1 - \epsilon_g)\} + J_g / (1 - \epsilon_g)$$
where \( Q_f \) is the feed volume flow, \( H \) the column height, \( V_c \) the volume of the collection zone, \( \epsilon_g \) the gas holdup, and \( J_b \) the superficial bias rate. Figure 49a may even suggest that the bias should be kept near zero. Experimentation at Falconbridge showed that no detrimental effect can be observed as long as the bias rate is kept above about -0.1 cm/s, as may be noted in Figure 49b.

In conclusion, \( J_u \) increases not just equal to the additional rate at which \( J_b \) enlarges, but also through a higher \( J_u \) as became apparent from equation (65) and Figure 48.

**Fig.49a.** Effect of bias rate on copper concentrate grade (0.05 m diameter column, copper retreatment concentrate at Mt. Isa) [154]

**Fig.49b.** Effect of bias rate on copper concentrate grade and on the nickel content in this concentrate (0.91m diameter column, Cu-Ni separation at Falconbridge)
Effect of gas rate

The bias rate can be altered two ways: through enlarging the difference between tailings and feed water flows, and through in-creasing the gas rate, as is evident from Figure 50 [155]. A control strategy that employs the tailings flow to maintain the pulp-froth interface, runs the risk to enter a mode of operation under negative bias conditions. This is brought about by increased mixing, which carries additional water from the collection zone into the froth zone. Mavros et al. [147] quantified the higher degree of mixing. When they raised the gas rate from 200 to 300 cm³/min, an attempt to fit the experimental RTD to the back mixing model yielded β values in excess of 40, even for a larger number of zones. A high β value produces a low Peclet number, or high vessel dispersion number (eq. (49)). Pal and Masliyah [156] estimated the bias rate Jb as a function of the gas rate and proposed the following model

\[ J_b = (1 - \epsilon_{gf}) \cdot \exp(0.1 - 2.5\epsilon_{gf}) \cdot u_r - \{(1 - \epsilon_{gf})/\epsilon_{gf}\} \cdot J_g \]

where \( \epsilon_{gf} \) is the gas holdup in the froth zone, \( u_r \) is the terminal rise velocity of a bubble, and \( J_g \) the superficial gas velocity. Bias rates thus calculated only agreed with measured rates for gas rates in excess of 3.5 cm/s. This model is limited as it does not take into account differences in froth gas holdup with location. Work at McGill is being undertaken

![Graph showing the relationship between bias rate and superficial gas rate](image.png)

Fig.50. Bias rate versus superficial gas rate, measured in a laboratory scale column. Tests were done for two pulp-froth interface levels [155].
to model radial gas holdup variations and is exemplified in Figure 51 [157]. In general, bubbles are not uniform in size, and in principle, larger bubbles rise at the centre of the column cross section at a larger velocity. The resulting radial variation of gas holdup is the cause of liquid circulating in a flotation column. Yasunishi [158] measured the gas holdup at various radial distances, using a dual electro resistivity probe. This probe consisted of two needle point probes of Pt-wire at 3 mm distance, and a stainless steel sheet reference probe attached to the inside wall. Measuring the gas holdup this way bears resemblance to the work conducted with the conductivity probe, as described in Chapter 2.5. His results correspond to the model being developed at McGill.

\[ \varepsilon_g (r/R) = \varepsilon_{gc} \left[ 1 - \left( \frac{r}{R} \right)^p \right] \]

Fig.51. A model for the radial variation of gas holdup [157].

The volume content of gas in a column is referred to as gas holdup, symbolized by \( \varepsilon_g \). It is a function of several parameters [159]

\[ \varepsilon_g = fn(d_{p0}, \rho_g, \eta, \eta_{st}, \rho_{st}, h_{st}) \]
where \( d_p \) is the bubble diameter, \( \rho_b \) and \( \rho_{sl} \) are the densities of bubble and slurry, \( J_g \) and \( J_{sl} \) are the superficial velocities of gas and slurry, and \( \mu_{sl} \) the slurry viscosity. Gas holdup can be calculated from the pressure differential \( \Delta P \) from two pressure transducers located in the collection zone, at a distance \( h \). It is calculated from

\[
\varepsilon_g = 1 - \Delta P(f_{sl}gh) \tag{66}
\]

where \( g \) is the gravitational acceleration. For an air-water system Kato et al. [161] developed a relationship for gas holdup as a function of gas rate

\[
\varepsilon_g = \frac{J_g}{[31 + 4.5J_g^{0.8}(1 - \exp(-0.040J_g^{1.8}))]} \tag{67}
\]

This relationship was tested at Cuajone in a 1.25 m column. From Figure 52 it becomes apparent that the measured values were higher than those calculated by eq. (67). In addition, the Cuajone data were confirmed in a later publication [155], which are reproduced in Figure 53.

Figures 52 and 53 exhibit a region of fairly linear relationship between gas holdup and gas rate. In this region, bubbles fairly uniform in size rise at a fairly uniform velocity. Above a transition gas velocity, holdup becomes unstable and the flow becomes characterized by large bubbles rising rapidly, displacing water and smaller bubbles downward. The extent of this range depends on factors such as frother type and addition rate, and sparger surface area. Dobby and Finch fitted gas holdup to gas velocity using the drift flux model [160].

![Graph](image)

**Fig.52.** Gas holdup in an air-water system as a function of gas rate and frother dosage (Cuajone data). Included are values calculated from eq. (67).
Fig. 53. Gas holdup versus gas rate in an air-water system. Effect of frother dosage \((J_f = 0.5\ \text{cm/s})\)

The slip velocity \(u_{sg}\) is the velocity of the gas phase relative to the liquid phase

\[
u_{sg} = J_g / \epsilon_g + J_l / (1 - \epsilon_g)
\]

(68)

where \(J_g\) and \(J_l\) are superficial velocities of respectively gas and liquid. The slip velocity of gas is related to the terminal rise velocity of a bubble following

\[
u_{sg} = u_t \cdot (1 - \epsilon_g)^{m-1}
\]

where \(m\) is a function of \(Re_b\), the Reynolds number for a gas bubble. Recently, modelling has indicated that using \(m=2\) allows a proper fit of the model. Yianatos [162] incorporated system parameters into the model, including the term for hindered settling. The equation for the slip velocity rendered to

\[
u_{sg} = \frac{[g d_b^2 / \rho_l \cdot (1 - \epsilon_g)^{m-1}]/[18 \mu_l (1 + 0.15 Re_b^{0.687})]}
\]

(69)

where \(d_b\) is the bubble diameter, \(\rho_l\) is the liquid density, \(\mu_l\) is the liquid viscosity, and \(Re_b\) the Reynolds number for a single bubble, and \(m\) a function of \(Re_b\).

The bubble diameter \(d_b\) changes with the gas rate according to

\[
d_b = C J_g^n
\]

(70)
where a value of 0.25 for \( n \) was found suitable for the range of interest in this system [163]. The value \( C \) corresponds to the mean bubble diameter for \( J_g = 1.0 \) cm/s, and depends principally on frother concentration. Figure 54 illustrates that the model fits the experimental data reasonably well, up to a maximum gas rate. Model violation at higher gas rates is explained by transition from bubbly flow to churn-turbulent flow. The formation of large bubbles that rise rapidly cause liquid circulation, as mentioned before. Gas rates at these values, however, are not of interest for column flotation.

The addition of surfactant has significant impact on gas holdup, due to changes in surface tension. The value \( C \) in equation (70) qualifies this change. Flint [164] correlated frother addition rate \( f \) to bubble size, for an external bubble contactor. Compared to a porous-media sparger this contactor would produce smaller to equally sized bubbles [135]. The relation he established, therefore, only is indicative of the effect

\[
d_b = 0.04(f)^{-0.14}J_g^{0.20}J_i^{-0.17}J_T^{0.26}
\]  

(71)

where \( J_g, J_i, \) and \( J_T \) stand for superficial velocities of respectively gas, at the tip of the sparger tube, and of the tailings. A higher frother addition would produce a smaller bubble. Hence, the bubble rise velocity becomes smaller (eq. (69)), which implies a larger holdup as equation (68) would predict. This was noticed in Figures 52 and 53. In a mineral system the bubble size produced may not be entirely relevant. Figure 2, Chapter 2.9 suggests that metallurgical results follow the same grade-recovery curve, in spite of the type of sparger employed. Apart from mechanical advantages, the only advantage of an external bubble contactor may be a possible reduced gas consumption to attain the same results.

Fig.54. Gas holdup versus superficial gas rate. Experimental data points are fitted with the drift flux model in which system parameters are incorporated [155].
Effect of solids on holdup

The role of solids on gas holdup is more difficult to describe. As a general consideration, small particles may promote or retard coalescence depending on their surface properties. Solids attached to a bubble increase the density of a bubble-particle aggregate. This density is equivalent to the ratio of mass of solids and the sum of volumes of solids and bubble. It is given by [165]

\[ \rho_b = \left( \frac{1}{\rho_p} + \frac{J_s}{C_s kh} \right)^{-1} \]

where \(\rho_b\) is the bubble-particle aggregate density, \(\rho_p\) the particle density, \(J_s\) the superficial gas velocity, \(C_s\) the concentration of floatable particles or bubble loading, \(h\) the height of bubble rise, and \(k\) the rate constant.

An increase in bubble density will decrease the slip velocity of the gas phase relative to the liquid phase. This can be appreciated by viewing equation (69), which was written for a gas-liquid system. With solids present, the liquid density term reverts to

\[ (\rho_{sl} - \rho) \]

which relates to the difference in density between slurry and bubble. Gas holdup will increase as predicted by equation (68). Simultaneously, slurry viscosity increases which would reduce the slip velocity further, but as the viscosity goes up, opposing effects are placed on the bubble rise velocity.

Effect of gas rate on recovery

The effect of gas rate on recovery is documented in Chapter 2.8. For a copper-nickel separation circuit, Figure 3a, Chapter 2.8, illustrates a rise in recovery up to a gas rate of 1 cm/s. More gas in the pulp enhances the probability of collision, as was explained in Chapter 1.3. Equation (47) relates the flotation rate constant to gas velocity and bubble diameter

\[ k (:) v_g E_s / d_b \]  

(47)

Gas rate affects bubble size (eq. (70)), but further gas rate increase will reduce the collection efficiency as was shown in Figure 26.

It becomes clear that several factors play a role in determining gas holdup. To state recovery as function of gas holdup may, therefore, make more sense. The recovery of copper in the Cuaione column as function of frother type and addition rate is depicted in Figure 55. It follows from this Figure that, although bubble sizes will change with the frother rate, as determined by Flint [164], the spread in results is quite narrow. This is useful in a control strategy where it is desired to maximize recovery as a function of a single parameter. The frothers used were Dowfroth 250 and Oreprep F507, both mixtures of
polyglycols, but the varying molecular weights produced different results in terms of frothing power.

![Graph showing recovery as function of gas holdup.](image)

Fig.55. Recovery as function of gas holdup. Results in the 1.25 m column at Cuajone for frothers Dow 250 and Oreprep F507, varying dosage rates between 0 and 25 ml/min at constant feed rate.

**Control philosophy**

In dealing with a single large unit compared to a multi-unit bank arrangement, the measurement of variables not only becomes feasible, but is also commanded. As a rule of thumb a higher value of the metal content in the concentrate demands a higher degree of process control to maximize the metallurgical performance. Examples of variability in an uncontrolled circuit were given in Figure 7, Chapter 1.1. Measuring control parameters assists in accomplishing stabilized control. Changes in variables are contained and upstream upsets may be dealt with by anticipation (adaptive control) and by setpoint ramping. This feature allows changes in setpoint levels at a controlled, predetermined rate, which incidentally may change depending on the severity of the upset. The objective is to maintain equilibrium in the system at all costs. Its importance has been dealt with in Chapter 1.4.

A simplified instrumentation diagram for the column is given in Figure 1, Chapter 2.3. Two control loops are absolutely essential, one to control the bias, another to maintain level [144]. Different philosophies for level control exist. The cheapest and about the simplest is
where level is kept constant through manipulation of the tailings control valve. But there is no direct control over bias, and wash water addition is manual [166]. This strategy will work provided the gas rate is kept sufficiently low not to drive enough water into the froth to effect a negative bias [155]. Cominco [167] sets the water rate in accordance with a predetermined displacement wash ratio. One displacement wash is given by a wash water flow equal to the volume of water recovered with the concentrate.

The more control loops are added the more sophisticated the installation. A simple installation features manually operated gas and spray water valves. Newer outfits incorporate setpoint selection by the operator on a controller. Attempts were made to control recovery to gas rate, but with reference to Figure 55 it appeared easier to relate recovery to gas holdup. Gas holdup is calculated on-line; the output of its controller is hooked up to a control valve in the air line, as shown in Figure 1, Chapter 2.3.

The control scheme becomes optimized when variables are managed to control grade and recovery, which was discussed in Figure 29. Assays usually are available on-line or through frequent sampling, because columns produce a final product. The recovery may be obtained through a calculation routine and may be displayed in a trend of for example a distributed control system (DCS), as is done at the concentrators of Gibraltar Mines, Cuajone, and Falconbridge. Immediate scrutiny on metallurgical results by the operator himself after manipulating variables is available. Hence, there is reassurance for management that an optimum performance can be attained throughout shifts by means of the window into operation provided by the DCS. As the floor operator is enabled to better understand the intricacies of the process he no longer has to rely on sole visual inspection and "gut" feeling.

Limitations in operation

At first thought, a control strategy based on holdup becomes attractive. Flaws of an interactive holdup - wash water control system are dispelled in Chapter 2.3. It became evident from tests at Cuajone that, especially when a deep froth is maintained, the lack of froth buoyancy appears to prevent concentrate carry-over, which results in a drop in recovery as is illustrated in Figure 56.

This drop off in recovery can be avoided by raising the pulp-froth interface to "push" the concentrate across the lip. This action was adopted in the control strategy. A build-up at the froth-pulp interface, as spoken of in Chapter 2.3, has been referred to as "impending holdup". Mineral recovery is hampered as a result of two phenomena probably occurring at once: coalescence of bubbles and increase in froth viscosity.
Fig. 56. Grade-recovery relationship for different froth levels. At a level of 0.9 m from the top froth buoyancy problems hampered recovery.

At this point the term carrying capacity should be introduced. The carrying capacity quantifies the physical limitation on the production of concentrate of a column. This concept becomes very important when the concentrate consists of heavily loaded froth. At this point, the available surface area restricts the production rate or recovery, not the available retention time. The height of the column can thus be limited to the point at which carrying capacity is reached; extending the column beyond this height would not increase production, only increasing the diameter could. Carrying capacity ($C_p$) is defined as the maximum solids mass rate that can be carried across the lip of a column per unit of time, and is expressed in $g/cm^2/min$ [168].

$$C_p = K_1 \pi d_p \cdot \rho_p \cdot J_g / d_b$$  \hspace{1cm} (72)

where $K_1$ is a constant, $d_p$ refers to particle size, $\rho_p$ to particle density, $J_g$ to superficial gas velocity and $d_b$ to bubble size. Up to a $d_{80}$ of 44 microns, and a $\rho_p$ of 5.8, equation (72) can be reduced to [168]

$$C_p = 0.068 \cdot d_{80} \cdot \rho_p$$
Contini et al. [169] established that the carrying capacity reaches a maximum value at a certain feed mass rate, and drops with a further increase in feed mass rate. This drop corresponds to a decrease in mass of particles in the concentrate that are larger than 37 microns, as is illustrated in Figure 57 [170].

Fig. 57. Mass recovery to concentrate of particles larger than 37 microns as a function of feed rate under limitations of carrying capacity. Concentrate becomes finer for higher feed rate.

For an increase in feed grade, at unchanged volume flow, the content of floatable species in the froth zone will rise if an equivalent recovery is to be maintained. If the column would already be operating at the limit of carrying capacity, then the overall recovery would actually drop as stated earlier. This drop was contributed to further loss of coarse particles in particular, due to either froth drop back or bubble coalescence. According to Moys [171] the predominant mechanism in the upper part of the froth is coalescence, while at the bottom of the froth pulp drainage is more important. The froth recovery is found to virtually remain the same, or a slight decline may perhaps be recorded [145]. Espinoza and other researchers [168], however, have not considered any viscosity effects within the froth (view equation (72)). Froth drop back, i.e. coarse particles falling off as rising bubbles collide with the bottom of the froth layer, may be the reason for a drop in recovery, but not the cause. A higher quantity of solids in an equivalent volume of froth would increase the static weight above the bottom part of the froth. Interstitial water is squeezed out, the relative froth density goes up, and the overall mobility of the particles, now more closely packed, is reduced. This concurs with Szatkowski [172], who assumed that one of the three most important phenomena that govern the flotation rate was the lack of buoyancy of the mineral laden bubbles. The other two were particle collection efficiency, and the rate of bubble coalescence in the froth. As a result, the increasing lack of buoyancy prevents the natural
flow of solids across the lip, a phenomenon often observed in flotation in general. A concentrate requires a certain fluidity to pour over the lip.

Barrera-Godinez [173] recently reported that the froth thickness must be taken into account when assessing flotation kinetics. A deeper froth layer increased the solids content of the concentrate, and resulted in lower rate constants. Also the effect of a scraper and its speed was lessened with increasing froth depth, which was attributed to a limitation in froth production rate above the bottom of the scraper as a result of bubble coalescence.

Up to a point, the concentrate mass flow should have increased corresponding with the increase in feed mass, or at least have stayed even. C_s would have remained constant and would not drop as illustrated in Figure 58. It is thus proposed that carrying capacity is directly related to buoyancy effects of the froth. A solution to this problem would be to raise the level, as has been indicated in Chapter 2.3. Carefully controlled laboratory scale test work should be performed to establish the validity of this reasoning. The concentrate density has indeed been observed to rise when such carrying capacity was reached.

The decrease in concentrate particle size, and consequent decrease in concentrate rate, appears to have its origin in the preferential drop back from the froth of coarse particles [145]. Drop back is the fraction of solids entering the froth that are rejected back into the collection zone. Studies in a laboratory column concluded that drop back occurs at the froth-pulp interface [146]. The interaction of the two zones is illustrated in Figure 59. For R_c as the collection zone recovery, and R_f as the recovery in the froth zone (drop back then is 1-R_f), the overall recovery is given by

\[
R_{fe} = \frac{R_c R_f}{R_c R_f + (1 - R_f)}
\]  

![Figure 58. Concentrate solids recovery versus concentrate feed rate [169].](image)
The recycling interaction through drop back from the froth and recollection within the collection zone is critical to the separation performance. Large size installations experience an $R_c$ of approximately 30%, compared to small sized laboratory scale columns that have $R_c$ reach anywhere from 50 to 80%. As bubbles rising in the collection zone hit the lower bubble layer of the froth, the sudden deceleration creates a mechanical shock that dislodges the larger, more weakly, attached particles. Beyond that point in the froth coalescence occurs. Bubbles collapse but particles may be picked up in the lower part of the froth zone. Coalescence also occurs in the upper part of the collection zone, because as bubbles rise they grow in size, become weaker, and probability of collision increases. When bubbles collide, one larger bubble is formed. Larger particles will shake loose as less surface area becomes available. Excessive coalescence is noticeable in the form of "burping" at the froth surface.

**Level, bias and holdup inferred from conductivity**

Chapters 2.4 to 2.7 relate the development of the conductivity probe to measure the froth-pulp interface in a column. A detailed description and diagrams of the stationary conductivity probe is given in Chapter 2.6, specifically in Figure 4. Basically, the probe consists of 25 rings, of which 13 in alternate position are connected to ground. Conductance is measured by each active ring in the cell bordered by its two adjacent grounded rings. A
computer routine imports the signal of each active ring in sequential manner. The point where the jump in conductance is largest corresponds with the pulp-froth interface; the higher gas holdup in the froth contributes to a lower conductance. A typical example of a resulting profile can be found in Figure 5, Chapter 2.5.

An alternative, and reliable method to measure the pulp-froth interface for a deep froth is the use of three pressure sensors. Two should be placed in the collection zone, and one in the froth zone, but near the interface. This requirement evolves from the fact that the froth bulk density is not linear with depth, as may be noted from Figures 3 and 7, Chapter 2.7.

The pressure exerted on a sensor placed in the collection zone equals the sum of hydrostatic pressures created by the froth layer and that part of the collection zone above the pressure transducer. With reference to Figure 1, Chapter 2.7, for one pressure transducer level can be calculated from

\[ h_f = \frac{f_c g h_b - P_B}{g(f_c - f_f)} \]  

(74)

where \( h_f \) denotes the pulp-froth interface measured from the top of the column, \( h_B \) the distance from the top of the column to the pressure sensor at location B, and \( f_c \) and \( f_f \) the densities of the froth and collection zones. No measurements of these densities are available and estimations lead to large discrepancies between calculated and actual level. Only with the addition of two more pressure sensors, one to measure the froth density, the other to measure the collection zone density, can level be calculated more accurately, using [Chapter 2.7]

\[ h_f = \frac{h_A[(P_c-P_B)h_c-P_c(h_c-h_B)]}{[(P_c-P_B)h_A-P_A(h_c-h_B)]} \]

where A, B, and C refer to the locations of the pressure sensors, at distance \( h \) from the top of the column, each measuring pressure \( P \).

Conductance profiles offer additional advantages. At first, conductivity has been used to calculate gas holdup [174]. The relationship between the adimensional conductivity and holdup is given by the Maxwell model [175]

\[ \frac{k_e}{k} = \frac{1 - \epsilon_g}{1 + 0.50\epsilon_g} \]  

(75)

where the adimensional conductivity is expressed as a ratio of \( k_e/k \), and \( k_e \) denotes the effective dispersion conductance at holdup \( \epsilon_g \) in ohm\(^{-1}\), and \( k \) the conductance in the liquid without air, also in ohm\(^{-1}\).
Fig. 60. Location of pressure sensing devices on the Falconbridge pilot column.

In a series of tests, in air-water medium, the gas holdup was simultaneously inferred from pressure measurements along the column, and from data obtained from the conductivity probe. For better reference Figure 60 has been included to indicate the locations of the pressure sensing devices installed on the pilot column at the Strathcona Mill of Falconbridge. The sight-glasses refer to transparent PVC tubes used as manometers. In addition, the column was equipped with three pressure transducers A, B, and C. Table 13 summarizes the pressure in meters water column, inferred at each location as function of gas rate. The gas rate was varied in the range of 0.7 to 2.6 cm/s. This same Table summarizes gas holdups, calculated using equation (66). The gas holdup at the top of the column was calculated from pressure transducer B, at the middle using sight-glasses 1 and 2, at the bottom using sight glasses 2 and 3. Results are shown in Figure 61, depicting gas holdup as function of gas rate.
Fig. 61. Gas holdup in air-water system inferred from pressure and conductance measurements. Tests were conducted in a 0.91 m pilot column at Strathcona Mill.

From the bottom to the top of the 13.7 m tall column gas holdup appears to increase consistently. The magnitude of change, however, is greater than can be accounted for by changes in head pressure. If this would hold true in a three phase system it might have important implications to the design and operation of flotation columns. Alternatively, this discrepancy can be displayed in a plot of pressure (m water column) versus distance from the top of the column. This has been done in Figure 62 for zero gas rate and the maximum gas rate of 2.6 cm/s.

For a three phase system, Figures 2 and 6, Chapter 2.7, however, showed a virtually linear relationship between pressure and distance from the top in the collection zone. The presence of solids apparently compensates for deviation from linearity, the exact mechanism requiring further study.

Holdup can be inferred from conductivity through use of equation (75). Table 14 summarizes gas holdup values calculated for the bottom active ring on the conductivity probe, which was located 1.25 m below the lip. From Figure 61 these values are observed to coincide with those calculated from pressure measurements made in the top zone of the column.
<table>
<thead>
<tr>
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<th>Gas Rate cm/s</th>
<th>P(A)</th>
<th>P(A)</th>
<th>P(B)</th>
<th>P(B)</th>
<th>P(C)</th>
<th>P(C)</th>
<th>SG1</th>
<th>SG1</th>
<th>SG2</th>
<th>SG2</th>
<th>SG3</th>
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<td>3.50</td>
<td>10.47</td>
<td>11.96</td>
<td>12.14</td>
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</tr>
<tr>
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<td>3.32</td>
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<table>
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<th>Test #</th>
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<th>Gas Holdup (%)</th>
<th>Gas Holdup (%)</th>
<th>Gas Holdup (%)</th>
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<td>Bottom</td>
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<td>2.1</td>
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<td>2.23</td>
<td>20.9</td>
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<td>10.4</td>
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Table 13. Summary of pressure measurements made at the various points indicated in Figure 60, and of gas holdup calculations inferred from pressure.
Fig. 62. Pressure plotted versus distance from the top for gas rates of 0.0 and 2.6 cm/s in an air-water system.

Fig. 63. Conductance profiles in the froth as a function of bias rate.
The slope of the conductance profile in the froth may be indicative for the bias in the column, as long as the conductance of the feed liquid is significantly larger than that of the wash water. In that case a negative slope of the profile translates into a positive bias rate, and vice versa. This is apparent from Figure 63.

If that condition is not satisfied than the change in slope becomes unmeasurably small. Conductance profile data still relate to bias rates after normalizing the conductance as is done for Figure 64. The value of the normalized conductivity ratio $\phi$ is calculated as

$$\phi = \frac{(k_c - k_w)}{(k_f - k_w)}$$

where $k$ is the specific conductivity, and attributes $c$, $w$, and $f$ stand for concentrate, wash water, and feed.

When the function $\phi$ approaches zero the system enters an operating regime of positive bias. The disadvantage of this method is that the conductance of additional points must be measured. Furthermore, more development is required to filter the signal for better prediction of the bias value, which could be done with use of the Kalman filtering technique.

<table>
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<tr>
<th>Test #</th>
<th>Gas Rate cm/s</th>
<th>Avg. Conductance for Ring 01 1) mOhm</th>
<th>Gas Holdup 2) %</th>
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<td>12.540</td>
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<td>11.387</td>
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<td>1.67</td>
<td>10.770</td>
<td>13.45</td>
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Notes 1) Ring 01 corresponds with the bottom active ring on the conductivity probe. It was situated 1.25 m below the lip of the column.

2) The gas holdup was calculated using equation (54). For tests 2 to 5, the $k$ value of test 1 was used as the conductance in liquid without air; similarly, for tests 7 to 9, the $k$ value of test 6 was used.

Table 14. Average conductance and holdup values for ring 01 (bottom active ring on conductivity probe) as function of gas rate.
Fig. 64. Normalized conductance profile in the froth as a function of bias rate.

1.5.2. Column Scale-up

At Falconbridge's Strathcona Mill two 2.12 m diameter columns, each about 13 m high, were commissioned in June 1989. The diameter and height of the columns were determined from the kinetic data arrived at in laboratory and pilot scale test work, while taking into account the projected forecast of copper recovery for the next ten years.

Experimental set up

The experimental set up of the laboratory scale column consisted of an 0.06 m diameter circular plexiglass column, 12 m in height. A sketch of the set up is given in Figure 65. It could be serviced from a platform near the top of the column, where also all feed pumps and controls were located. A peristaltic pump delivered feed to the column, which entered about 2 m from the top. The tailings flow was also controlled by means of a pump, whereas concentrate freely flowed across the lip into a separate container. Level was maintained at about 1 m from the top by varying the speed control on the tailings pump. In the bottom cone the column was provided with a porous stainless steel sparger for gas supply, which
was manually regulated and measured with a rotameter. A typical bubble diameter produced by the porous steel sparger was approximately 0.8 to 1.0 mm, about 0.1 to 0.2 mm smaller than is being produced by a cloth-type sparger at the same gas rate. Typical gas flow rates are between 1.5 and 2.0 cm/s. Wash water was added through a copper ring just below the surface in the froth. Its initial rate was set prior to the test by estimating the bias (normally between 0.1 and 0.5 cm/s) and adding the expected flow of water to the concentrate.

The laboratory scale column was operated in a semi batch mode. To generate the grade-recovery curve the residence time in the column was altered by adjusting the tailings flow rate. Feed rate and wash water were adapted accordingly to observe a positive bias at a fixed rate. The gas rate was set independently at predetermined levels. After at least two residence times steady state was reached and sampling started. Timed samples of feed, concentrate and tailings were taken for assay. Measurements were made of the volumetric flow rates of launder and wash water. The air flow was read from a calibrated rotameter.

Fig.65. Schematic of the experimental setup of the laboratory scale flotation column.
Fig. 66. Example of comparable metallurgy, nickel grade versus pentlandite recovery from recleaning Strathcona Mill secondary rougher concentrate.

Simultaneously, timed batch tests were conducted in a standard Denver mechanical laboratory cell. Grade recovery curves, obtained from testing in the mechanical cell and the small diameter column on the same feed material, are presented in Figure 66. The material recleaned was the secondary rougher concentrate, containing mostly pentlandite, pyrrhotite and some chalcopyrite.

Falconbridge's pilot column measured 0.91 m in diameter and was 13.7 m high. The column was fully equipped with instrumentation to allow process control. All input and output signals from the instruments were fed into a Analog Devices micromac 6000 computer. The micromac was connected with an IBM-AT, for data collection and trend display. Display of the process variables in the control room was attained through Taylor Mod 30 controllers. Feed and tailing mass flows were measured with Krohne magnetic flow meters, compensated for the magnetic contents in the streams (pyrrhotite); the spray water was measured using an orifice plate. Ohmart nuclear density gauges were installed on the feed and tailings lines. Feed was pumped from a plant tank into a separate tank ahead of the column. The level in this tank was maintained at setpoint in a control loop that involved a Milltronics Microranger II ultrasonic level detector and a Fisher control valve in the line between the plant tank and the column feed tank. It was possible to control the feed density through automatic water addition to the column feed pump. Concentrate flowed by gravitation into a concentrate tank at the base of the column. Centrifugal pumps were used to pump feed and tailings, but a Galligher sump pump suspended in a narrow concentrate tank transferred the concentrate to the copper thickener. The column was equipped with
three sight glasses that served as manometers, and three pressure transducers as described in the previous subchapter and in Figure 60.

Initially, air was supplied to the column through an array of steel pipes clad with two layers of fabric, as is described in detail in Chapter 2.9. This arrangement was replaced with an US Bureau of Mines external bubble contactor. Air supply was controlled in a separate loop, and was tied in with the calculated holdup. Water distribution was tested with both an overhead tank and a submerged spray header. Provision was made to facilitate changes in vertical position of the distributor. Wash water was tied in with level, level being maintained by changes in water addition. For bias control, the tailing flow equalled the setpoint of the feed flow to which a changeable constant was added through the computer.

Flotation rate constants

Scale-up from kinetic data can be approached two ways: using an overall rate constant $k_{te}$, or using a collection zone rate constant $k_c$ along with a value for froth recovery $R_f$. The correlation between the two rate constants can be described when considering two extreme mixing models for first order kinetics, perfect mixing and plug flow.

For a first order rate of reaction with a rate constant $k_{te}$, exhibiting perfect mixing with a mean residence time $\tau_p$, the overall recovery $R_{te}$ is given by

$$ R_{te} = 1 - (1 + k_{te} \tau_p)^{-1} \quad (76) $$

In considering a conceptual configuration of collection and froth zones for a column arrangement, the overall recovery can also be expressed by [145]

$$ R_{te} = R_c \cdot R_f / (R_c \cdot R_f + 1 - R_c) \quad (73) $$

where $R_c$ is the recovery in the collection zone, and $R_f$ refers to the recovery in the froth zone. The calculation of $R_c$, the recovery in the collection zone was given by Levenspiel [176]

$$ R_c = 1 - \left\{ \frac{[4a \cdot \exp(1/2N_d)]}{[(1+a)^2 \cdot \exp(a/2N_d)-(1-a)^2 \cdot \exp(-a/2N_d)]} \right\} \quad (77) $$

where $a = (1+4k_c \cdot \tau_p \cdot N_d)^{e}$

and $k_c$ is the collection zone rate constant, $N_d$ the vessel dispersion number (reciprocal of the Peclet number), and $\tau_p$ the mean residence time.

Replacing in (76) the $k_{te}$ by $k_c$ gives $R_c$, the recovery in the collection zone. In combination with (73) and after cancelling terms, equating these equations yields

$$ k_{te} = k_c \cdot R_f \quad (78) $$
Fig. 67. Rate constant determination for each mineral species in a copper nickel separation application (A gangue, B chalcopyrite, C pyrrhotite, D pentlandite)
For a plug flow case the overall recovery is given by

\[ R_{fc} = 1 - \exp(-k_{fc} \cdot \tau_p) \]  

(79)

After a similar manipulation as for perfect mixing, combining (79) with (73), and with equation (79) written for \( k_c \) to calculate \( R_c \), yields

\[ k_{fc} = \frac{1}{\tau_p} \cdot \ln[\exp(k_c \cdot \tau_p)R_f + 1 - R_f] \]  

(80)

\( k_c \) pertains to the collection zone and is a true rate constant [163]. \( k_{fc} \) on the contrary is a function of \( \tau_p \) as (80) indicates, and does not have, therefore, the properties of a rate constant under plug flow conditions. Only in the case of perfect mixing, \( k_{fc} \) may be classified a true rate constant, when \( R_f \) is constant.

It becomes evident that \( R_f \) must be known. Under plug flow conditions, \( k_{fc} \) becomes a reasonable approximation for \( k_c \) at high values of \( k_c \) regardless of \( R_f \). Then the \( k_{fc} \) measured in a laboratory column can be used to estimate \( k_c \), and can be expended in scale-up models. Plug flow models no longer apply in plant scale columns, and the froth recovery becomes important. At this place, it may be worthwhile to remind that froth drop back equals 1-\( R_f \) [145].

The experimental technique to measure \( k_{fc} \) in a laboratory scale column has been described by Dobby and Finch [177]. A number of column tests are conducted sequentially at different nominal liquid residence times, ensuring results cover the plant recovery range. Figure 67 depicts the results of a typical test series, in which the objective was to separate chalcopyrite from pentlandite, pyrrhotite, and gangue.

From this Figure one may observe that first order kinetics do not always fit the experimental data well. The flotation rate is recovery dependent and the two-slope straight line represents better the actual non-linear process. It may be interpreted as two species of the same mineral are floating at different rates, one slow and one fast. This phenomenon was studied elsewhere [109]. The usual approach is to use the rate constant that corresponds with the target recovery. When a single rate constant does not represent the entire process, simultaneous simulation of grade and recovery becomes difficult.

**Scale-up model**

The intrinsic of the simulator used to scale-up results from a small column to any diameter larger column, has been described by Yianatos et al. [178]. Figure 68 reproduces a flowchart of the sequential steps used in the simulator. Calculation procedures for each step are well documented elsewhere [177][159]. In the case of Falconbridge the model was employed to design a 0.91 m column from the results obtained in a 0.06 m column, while data from the latter column formed the basis to scale up to a 2.12 m column. In the example below, and pictured in Figure 69, simulation studies refer to scale-up from the laboratory size column directly to plant size. It eliminates the intermediate step to observe the consequences of this move.
Fig. 68. Flowchart of calculation steps employed in the simulation model.

The following example guides the reader through one iteration of the scale-up routine. Values referred to in this calculation are summarized in Table 15. To calculate the collection zone recovery for any mineral requires knowledge of the particle residence time $\tau_p$ and the vessel dispersion number $N_d$. These can be calculated from the particle slip velocity $u_{sp}$ and interstitial liquid velocity $u_i$ through established formulas [176,150]
<table>
<thead>
<tr>
<th>COLUMN SPECIFICS</th>
<th>PLANT SCALE</th>
<th>LAB SCALE</th>
<th>UNITS</th>
</tr>
</thead>
<tbody>
<tr>
<td>Base unit diameter, $d_c$</td>
<td>2.120</td>
<td>0.057</td>
<td>m</td>
</tr>
<tr>
<td>Column area, $A_c$</td>
<td>3.53</td>
<td>0.0026</td>
<td>m$^2$</td>
</tr>
<tr>
<td>Number of sections, $N$</td>
<td>4</td>
<td>1</td>
<td></td>
</tr>
<tr>
<td>Height collection zone, $H_c$</td>
<td>10</td>
<td>10</td>
<td>m</td>
</tr>
<tr>
<td>Bias rate, $J_g$</td>
<td>0.05</td>
<td>0.17</td>
<td>cm/s</td>
</tr>
<tr>
<td>Gas holdup, $\epsilon_g$</td>
<td>0.14</td>
<td>0.14</td>
<td></td>
</tr>
<tr>
<td>Gas rate, $J_g$</td>
<td>1.4</td>
<td>2.0</td>
<td>cm/s</td>
</tr>
</tbody>
</table>

**FEED CHARACTERISTICS**

<table>
<thead>
<tr>
<th></th>
<th>PLANT SCALE</th>
<th>LAB SCALE</th>
<th>UNITS</th>
</tr>
</thead>
<tbody>
<tr>
<td>Feed rate</td>
<td>34.1</td>
<td>0.02</td>
<td>t/h</td>
</tr>
<tr>
<td>Feed percent solids</td>
<td>35.0</td>
<td>28.7</td>
<td>%</td>
</tr>
<tr>
<td>Feed slurry density, $f_{sl}$</td>
<td>1.37</td>
<td>1.28</td>
<td>g/cm$^3$</td>
</tr>
<tr>
<td>Number of mineral species</td>
<td>4</td>
<td>4</td>
<td></td>
</tr>
<tr>
<td>Solids specific gravity, $f_s$</td>
<td>4.3</td>
<td>4.3</td>
<td></td>
</tr>
<tr>
<td>Specific gravity of cp</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>4.2</td>
<td>4.2</td>
<td></td>
</tr>
<tr>
<td></td>
<td>4.8</td>
<td>4.8</td>
<td></td>
</tr>
<tr>
<td></td>
<td>4.6</td>
<td>4.6</td>
<td></td>
</tr>
<tr>
<td></td>
<td>2.7</td>
<td>2.7</td>
<td></td>
</tr>
<tr>
<td>80 % passing size, $d_{80}$</td>
<td>50</td>
<td>50</td>
<td>μm</td>
</tr>
<tr>
<td>Mineral feed grade, cp</td>
<td>32.4</td>
<td>24.7</td>
<td>%</td>
</tr>
<tr>
<td></td>
<td>36.9</td>
<td>37.2</td>
<td>%</td>
</tr>
<tr>
<td></td>
<td>21.7</td>
<td>20.6</td>
<td>%</td>
</tr>
<tr>
<td></td>
<td>9.0</td>
<td>17.6</td>
<td>%</td>
</tr>
</tbody>
</table>

**B Rate constants**

To calculate point M in Fig.69:

\[
\begin{align*}
k_{fc} & \text{ for cp} & 0.2855 & \text{min}^{-1} \\
      & \text{pn}      & 0.0009 & \text{min}^{-1} \\
      & \text{po}      & 0.0039 & \text{min}^{-1} \\
      & \text{gangue}  & 0.0146 & \text{min}^{-1} \\
\end{align*}
\]

After calibrating simulator:

\[
\begin{align*}
k_e & \text{ for cp} & 0.2240 & \text{min}^{-1} \\
      & \text{pn}      & 0.0068 & \text{min}^{-1} \\
      & \text{po}      & 0.0096 & \text{min}^{-1} \\
      & \text{gangue}  & 0.0147 & \text{min}^{-1} \\
\end{align*}
\]

Table 15. Characteristics of column and feed to the column, used in the calculations of scale-up.
Step a) Calculation of slurry flow rate $J_{sl}$

$$J_{sl} \approx J_{r} = \left[ Q_{fw} + (1 - R_{tot}) \cdot Q_{rs} \right] / A + J_{g}$$

$$= [15981 + (1-0.30)(2001)] / (35299) + 0.05 = 0.54 \text{ cm/s}$$

where $Q$ denotes a volume flow rate in cm$^3$/s and $R_{tot}$ is an estimated value of the total mass recovery. On first iteration $R_{tot}$ would have been zero; for the purpose here the total mass recovery is estimated at 30%. At the end of this subroutine the calculated value will be checked against this value. In case of deviation, more iterations are required. Subscripts $f_{wp}, f_{ps}, p_{r}, r, t$, and $A_{st}$ are abbreviations for respectively feed water, feed solids, bias, tailings, and slurry; $J$ represents the superficial velocity, and $A$ is column area.

Step b) Calculation of liquid retention time $\tau_{l}$

$$\tau_{l} = H_{c} \cdot (1 - \epsilon_{g}) / J_{sl}$$

$$= [(10) \cdot (0.86) / (0.54)] \cdot (100/60) = 26.5 \text{ min}$$

where $H_{c}$ is the height of the collection zone, and $\epsilon_{g}$ is the gas holdup. The factor $(100/60)$ converts m to cm and s to min.

Step c) Calculation of particle slip velocity $u_{sp}$

This is an iteration procedure that calculates $u_{sp}$ from the general Masliyah equation [160] and the particle Reynolds number $Re_{p}$:

$$u_{sp} = \left\{ g d_{s}^{2} \cdot (f_{p} - f_{sl}) \cdot (1 - \phi_{s})^{2.7} \right\} / \left\{ 18 \mu_{sl} (1 + 0.15 \cdot Re_{p}^{0.687}) \right\}$$

and

$$Re_{p} = d_{p} u_{sp} \cdot f_{l} \cdot (1 - \phi_{s}) / \mu_{sl}$$

where $\phi_{s}$ represents the volume fraction of solids in the column, and $\mu_{sl}$ the slurry viscosity, $f_{p}$, $f_{l}$, and $f_{sl}$ refer to the densities of the particle, the liquid, and the slurry, and $d_{p}$ is the particle diameter. Using the values in Table 15 and for $\mu_{sl} = 0.01$ poise and $\phi_{s} = 0.05$, these formulas iterate to $u_{sp} = 0.36$ cm/s.

Step d) Calculation of particle retention time

$$\tau_{p} = \frac{\tau_{l} \cdot [J_{sl} / (1 - \epsilon_{g})] / [J_{sl} / (1 - \epsilon_{g}) + u_{sp}]}$$

$$= (26.5) \cdot [(0.54) / (0.86)] / [(0.54) / (0.86) + (0.36)]$$

$$= 16.8 \text{ min}$$

Step e) Calculation of particle vessel dispersion number $N_{d}$ from equation (63)

$$N_{d} = 100[0.063d_{c}(J_{g} / 1.6)^{0.3}] / \left\{ [J_{sl} / (1 - \epsilon_{g}) + u_{sp}] \cdot H_{c} \right\}$$
= 100[((0.063)(2.12)(1.4/1.6)^0.3])/[((0.54/0.86+0.36)· 10]
= 1.30

where \( J_g \) is the superficial gas velocity. Note that this number is calculated for a non-baffled column. Baffling a column requires adjustment of \( d_c \), the basic column diameter to a hydraulic diameter of the newly formed compartment. The compartment area must then be multiplied by the number of compartments to arrive at total area available. Also at this stage, a check is made for vertical alignment of column of baffles. If not, adjustment of \( N_c \) is required using equation (64).

Step f) Calculations of mineral recoveries in the collection zone, with aid of equation (77)

for chalcopyrite:

\[
R_c = 1 - \frac{[4(4.54)\exp(1/2.6)]}{[5.54]^2\exp(4.54/2.6) - (-3.54)^2\exp(-4.54/2.6)}
\]

\( R_c = 0.85 \)

and \( a = [1+(4)(0.224)(16.8)(1.30)] \)

\( a = 4.54 \)

for pentlandite the recovery calculates to 0.104

Step g) Calculation of the total recovery \( R_{tc} \).

For an assumed value for \( R_t \), the froth recovery, calculations with equation (73) yield:

<table>
<thead>
<tr>
<th>( R_t )</th>
<th>( R_{tc} ) (cp)</th>
<th>( R_{tc} ) (pn)</th>
<th>( R_{tot} )</th>
<th>cp grade</th>
</tr>
</thead>
<tbody>
<tr>
<td>50</td>
<td>73.9</td>
<td>5.5</td>
<td>26.0</td>
<td>92.1</td>
</tr>
<tr>
<td>30</td>
<td>55.6</td>
<td>3.4</td>
<td>19.3</td>
<td>93.3</td>
</tr>
</tbody>
</table>

Initially in step a) total mass recovery was assumed at 30 %. The Table above shows that the result from the first iteration yields lower values for \( R_{tot} \) than assumed initially in step a) (26.0 versus 30.0 %). This implies that additional iterations are required. Furthermore, it was noted that the calculations in this example pertain to a non-baffled column. Chemical engineering principles, supported by experience, imply that baffling a column of a diameter larger than 1 m makes sense as the system will then more closely approach a plug flow concept. In a subsequent step, the model also corrects for carrying capacity limitations, according to equation (72).

**Simulator application for Falconbridge**

Results of the two approaches to scale-up as alluded in the text are illustrated here. The data used in the model are summarized in Table 15. It should be noted that copper-nickel
separation in a column at Falconbridge is not carrying capacity limited, but rather is a function of retention time.

Approach a): Use of overall rate constants obtained from batch tests in a small diameter column.

Except for gangue, the rate constants inserted in the simulator were those obtained from Figure 67. For gangue a lower value was used than obtained from this Figure as such value was the result of various repeated tests in subsequent experiments. The rate constants from Figure 67 are denoted \( k_{rc} \). Ideally only for chalcopyrite the high value of \( k_{rc} \) could be used as an approximation of \( k_c \). However, rate constants thus obtained for all mineral species involved were inserted in the simulation model. As these constants referred to flotation overall, the froth recovery was set to 100 %, i.e. zero froth drop back. For a single column unit, point M in Figure 69 represents the outcome from the simulation. Coordinates are 30 % copper concentrate grade at a recovery of 93 %.

Fig. 69. Comparison of actual results attained in a 2.1 m diameter column at Falconbridge to those predicted by the simulator. Point M reflects the outcome of simulation prediction using directly the rate constants obtained from tests in a 0.06 m diameter column. Lines A and B are the predictions for a single column and a column circuit, after calibrating the simulator to adjust the rate constants.
Approach b): Use of overall laboratory column grade and recovery to scale up.

To incorporate froth drop back, the following steps were taken. Test results from the 0.06 m diameter column were inserted into the simulation program. Rate constants were adjusted to match the results predicted by simulation to actual results. In this calibration step the small column was assumed to have reached a froth recovery of 80%. From Table 15, part B, the degree of adjustment can be observed to complete calibration.

The next step involved the use of the simulator to predict results for a 2.12 m diameter column, using the new rate constants. Large diameter columns do not achieve high froth recoveries. $R_f$, therefore, was varied down to 40%. This way a grade-recovery curve can be obtained, which is included in Figure 69 as line A. The large column is baffled; this was taken into account by inserting a hydraulic diameter per compartment, as well as the number of compartments created (four).

Similarly, prediction of the performance for a two-column circuit was made with the simulator. The tailings of the first served as feed for the second column, while the concentrate of the second was combined with the feed for the first. To balance the circuit iteration is required, but usually is reached after three iterations. Curve B in Figure 69 presents these predictions obtained after the simulation procedure. The numbers on both curves A and B refer to the froth recovery assumed for that point.

<table>
<thead>
<tr>
<th>Rate Constants: Mineral</th>
<th>From Tests in Pilot Column</th>
<th>After Calibrating Simulator</th>
</tr>
</thead>
<tbody>
<tr>
<td>Chalcopyrite</td>
<td>0.1700</td>
<td>0.1870</td>
</tr>
<tr>
<td>Pentlandite</td>
<td>0.0007</td>
<td>0.0008</td>
</tr>
<tr>
<td>Pyrrhotite</td>
<td>0.0048</td>
<td>0.0053</td>
</tr>
<tr>
<td>Gangue</td>
<td>0.0130</td>
<td>0.0143</td>
</tr>
</tbody>
</table>

Table 16. Rate Constants obtained from tests in the 0.91 m diameter pilot column, and adjusted by calibrating the simulator.

Falconbridge's 2.12 m columns were scaled up from pilot results, not from laboratory column results. In reality, Falconbridge's two copper circuits each consist of a column, followed by a row of conventional cells, scavenging the column tailings. Their concentrate flows back to the feed tank of the column and is combined with the original feed. A survey around the column and of the entire circuit produced results, which have also been included in Figure 69. In fact the process behaves slightly better than expected. This fact is contributed to the large circulating load that exists in between the froth and collection zones. It has not been accounted for in the simulation model.
A second point of interest is how point M compares to line A. Obviously, froth drop back should be considered. If not, lower than expected recoveries may ensue. $k_f$ is a function of $r_p$ and cannot be extrapolated to a large diameter column. The approach to scale up from overall flotation rate constants may result in an incorrect representation of a final installation.

Fig. 70. Comparison of actual results attained in a single column and in a circuit at Falconbridge to those predicted by simulation from test results from both a laboratory scale column as well as a 0.91 m diameter pilot column.

It is seen in Figure 69 that the predicted results do not coincide entirely with actual measurements. A similar scale-up routine was followed, therefore, with results that were obtained from test work in a 0.91 m diameter column. Table 16 summarizes the rate constants that were calculated from the test work, as well as the adjusted values after calibration of the simulator program. For the conditions summarized in Table 15, and using approach b, Figure 70 could be composed analogous to Figure 69. It compares predicted scale-up results from test work with the 0.06 m and the 0.91 m column to actual plant results in a single column and in the entire column circuit. From these results it may be concluded that scaling-up from a small diameter, laboratory sized column might be done, but only with extreme caution, whereas scale-up from pilot columns should not pose problems. The reason is that the degree of froth drop back and froth recovery that occurs in large diameter columns, is difficult to reproduce under laboratory conditions.
1.6. **Concluding Remarks**

In this dissertation the interaction between grinding and flotation was discussed in detail. Neither of these two unit processes can be viewed independent from each other. The various important issues are highlighted below.

1. To meet product specifications two technical problems must be resolved:
   - concentration of minerals to a grade suitable for subsequent processing.
   - concentration of minerals to the desired particle size.

   Each problem is related to internal factors, signifying mineral occurrence in relation to each other, and external factors, which are inherent to the situation of the ore body. Especially the first dictate the interaction between grinding and flotation.

2. The type of mineral and host rock, the mineral occurrence in an ore body pertaining to size and intergrowth, dictate the selection of grinding equipment, the desired product size, and the number of flotation stages.

3. To obtain an adequate response in flotation the extent of mineral surface area exposed is important. Particles do not necessarily need to be ground to complete liberation; smaller sized grains would ensue that could not even be recovered by flotation.

4. Multiple stages of grinding and flotation are desired to separate minerals as coarse as possible to minimize costs and to prevent over grinding. Middlings streams can be prepared for subsequent re-treatment, while the overall selectivity is enhanced. In addition, the individual recovery of different valuable minerals is optimized, because each may respond best in a certain size range.

5. Liberation may be important but the mineral distribution-by-size should be combined with recovery-by-size curves. This would incorporate the phenomenon of preferential grinding.

6. It is not just liberation and recovery-by-size flotation response that results in optimum metallurgy. Stability in a grinding circuit to produce a product consistent in size and feed density, is a prerequisite. If this condition is not met, one cannot expect to prevent the fluctuations in grade and/or recovery in a flotation circuit, which lead to losses in metal.

7. The stability in a grinding circuit can be attained by implementation of process control. In some flotation circuits the correct particle size determines the recovery attainable. In other circuits the use of control at particle size simply could mean a high degree of stability as this is the most sensitive parameter. Generally, once fluctuations in size are controlled to a minimum, other parameters, such as pulp density, are also maintained constant. This
logic does not apply vice-versa. Limitations in a grinding circuit dictate the use of constraints. Variations in ore may cause the grinding process to exceed boundaries of stable operation; examples are mill power and cyclone underflow density. Grinding control has been applied successfully worldwide.

8. The conditions in a ball mill have a pronounced effect on the subsequent floatability of the material. Lack of air supply, but more importantly, the type of grinding media, may induce a reducing atmosphere when abraded steel from the grinding media oxidizes. As a result, the surface redox potential will drop to a low cathodic potential value. Mineral flotation is a function of the redox potential; controlling its increase after grinding due to oxygen uptake, would make selective flotation possible.

9. Another result of the grinding stage is the intimate contact between the minerals themselves, and between the minerals and the grinding steel. Galvanic interactions will promote the rate of sulphur formation on the surfaces of galena, sphalerite, and chalcopyrite. The flotation response of pyrrhotite in contact with steel balls drops dramatically, for example from 90 to 23 %, but only to 70 % in the presence of pyrite.

10. A third effect as a result of grinding is that abraded steel particles adhere to the minerals and oxidizes to ferric hydroxide. This competes with the formation of polysulphides or a metal-deficient sulphide layer on the surfaces of the minerals, and results in less available space for collectors to adsorb onto.

11. Among other factors, the choice of frother depends on particle size, which may give an opportunity for a slightly coarser grind.

12. If the particle size produced by the grinding circuit is relatively coarse, but suitable for flotation, then the pulp density in the flotation circuit should be adjusted accordingly. Bubble loading is restricted by buoyancy. Less coarser sized particles can adhere to a bubble than fine grains if the bubble-particle aggregate is to reach the interface. In a pulp of reduced density, competition for bubble space is less severe, thus enabling a higher recovery of the coarser size fractions.

12. Coarser particles require a longer contact time, but plateau where this time becomes independent of the particle size. By enlarging the bubble size, the contact time can be increased further. In a pulp of higher density, the contact time required will increase. Research data also suggest that the attachment time for a particle increases linearly with particle size on a log-log scale. Large particles show a poor flotation performance as their attachment time required generally exceeds the contact time. For more hydrophobic particles the mentioned straight line relationship shifts to lower values, inferring a requirement for shorter attachment time. This can be effected by the addition of reagents.
13. A model [74] that relates the collection efficiency to particle size and bubble diameter demonstrates a peak in efficiency. It corresponds well to the expected maximum in size-by-size recovery data.

14. A costly capital investment to improve or expand a current operation should be preceded by a managerial philosophy that spells out subsequent steps to eliminate bottlenecks. The procedure advocates the knowledge of process capacity and maximum utilization of the various unit processes. It must be rational, take an incremental approach, start at the beginning of the process, incorporate a cost-benefit analysis for each step, and must view the desired final plant lay-out. Generally, optimization in the flotation circuit follows stabilization, and maximization in the grinding circuit.

15. Redox potential control may well become one of the most important control variables within the flotation circuit, once adequate instrumentation to measure the potential has been developed.

16. The hydrophobic entity of sulphide minerals appears to be either elemental sulphur, a polysulphide or a metal-deficient sulphide layer, formed by oxidation of the mineral surface. This layer renders sulphide minerals amenable to collectorless flotation. For some sulphide minerals, however, the addition of a collector is required to attain the best recovery possible.

17. In column flotation the particle size of the concentrate to be produced dictates the carrying capacity. This bears relevance especially in cleaning applications, where the quantity of concentrate produced is high in relation to the column feed.

18. Columns are more conducive to complete process control than conventional cells, because of the relative compactness of the unit process.

19. The measurement of conductivity in a column to infer gas hold-up, bias and level, reduces the inaccuracy normally a result of conventional level measurement techniques. This is due to the relative insensitivity to changes in operating conditions, notably pulp density, bubble size, gas rates.


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PART TWO - PAPERS

2.1. AUTOMATIC GRIND CONTROL AT THE CUAJONE COPPER CONCENTRATOR

B.J. Huls and E. Socolich


For many years on-stream particle size measurement capability has been available to the base metals industry. Over 200 systems were installed worldwide by Armco-Autometrics, an American company, specialized in the development of electro-mechanical devices for industrial process control applications. Recently, their newest particle size monitor, the PSM-400 was installed in one of eight primary mill grind circuits at the Cuajone copper concentrator. In combination with a multiloop controller, the system is expected to generate annually a return on investment to more than offset its capital costs by operating the mill at maximum throughput, maintaining a desired particle size in the cyclone overflow. It would provide the capability of handling mill circuit upsets and ore changes without disturbing significantly the particle size. The fast responding particle size signal affords an excellent visibility into circuit operation. When an operator adjusts manually the feed rate or dilution water, he can view the results of his change immediately. Without particle size control the operator would adjust, for example, the feed rate belatedly according to size analysis from samples drawn from the cyclone overflow in the previous shift. The particle size monitor thus has the advantage of utilizing an actual measurement for control, rather than a difficult-to-maintain inferred value.

INCREASED THROUGHPUT DESIRED

The Cuajone copper concentrator treats about 52,000 tons of ore daily. It passes through a 60" x 89" Allis Chalmers primary crusher and, via two Rexnord Standard Cone crushers, continues to six Rexnord shorthead cone crushers to be collected at the fine ore stockpile (Figure 1). Twenty four apron feeders supply eight 16' x 20' Allis Chalmers ball mills; each mill is in close circuit with a cyclone cluster of four cyclones. The cyclone overflow is pumped to a second set of cyclones to separate the slimes from the sands for individual rougher flotation treatment. The rougher concentrates are combined, reground, and cleaned in two stages. The final concentrate is thickened for subsequent separation of molybdenite. Both the molybdenite and copper concentrates are thickened, filtered, and dried to be shipped to smelters elsewhere.

Originally, the plant was designed for 45,000 tpd. By taking advantage of many provisions and capacities originally installed for future expansion, we are now able to process 52,000 tons of ore daily. It is suspected, however, that, even at this increased mill throughput, the primary grinding mills do not always work at maximum load. A detailed flowsheet of the
grinding circuit, including the existing control and monitoring devices to aid operation is depicted in Figure 2. Apron feeders deposit the ore on an inclined belt equipped with a Ramsey belt scale. It enters the ball mill together with the cyclone underflow and process water. The mill discharge is collected in a sump provided with a level indicator and alarm for high and low level. Fresh water and reclaimed water is pumped into the sump at a rate to ensure a set cyclone overflow density. Mostly, however, the sump is running near empty. Mill control consists of hourly samples of cyclone overflow and mill discharge to check slurry densities. Also, a screen analysis is provided of a shift composite of the cyclone overflow to enable the mill foreman of the following shift to adjust the mill feed according to particle size.

For efficient operation of the flotation circuit a constant particle size distribution is desired. In addition, a reduction in variation of particle size may provide increased metal production through closer metallurgical control and maximum mill feed. It may also be proven that, at a coarser grind than usual, the resulting increased mill throughput would more than compensate for a lower, but acceptable, metal recovery.

Fig. 1. Schematic flowsheet Cuajone copper concentrator.
Fig. 2. Detailed flowsheet of primary grind circuit.

THE INSTALLATION OF THE PSM; ITS OBJECTIVES

In June this year the particle size monitor, the PSM-400, was installed in one mill circuit to meet the following objectives:
- a more stable grinding circuit performance
- increased tonnage
- improved control system reliability and availability
- increased copper production
- improved operational flexibility

The first objective entails that using size based control, the feed to the flotation circuit is more uniform, which would improve flotation operations. Because of the narrower particle size distribution, a slightly higher tonnage is expected at increased metal recovery. Improved flexibility is attained because the fast responding particle size signal affords mill operation at the selected grinds. The objectives of mill operation can now be based on economical and metallurgical considerations, rather than on control system limitations. Records of PSMs installed worldwide indicate low down-time and high operator acceptability, which contributes to a high control system reliability.
WHAT IS A PARTICLE SIZE MONITOR?

Basically, the PSM consists of an air eliminator and two pairs of sensors (Figure 3: the PSM). Each pair of sensors contains one transmitting and one receiving transducer, made of piezoelectric barium titanate crystals. At certain discrete frequencies an ultrasonic signal is attenuated (reduced) by the number of particles. This attenuation is not affected by size, but strictly by density, and the signal can be displayed immediately. At other frequencies a signal is attenuated by both the size and number of particles. From this signal the percent solids information had to be removed before it can be displayed as size. The sensors, however, can not tell the difference between air bubbles and particles; removal of air is necessary and is done in the Air Eliminator.

Fig. 3. The particle size monitor.
The Air Eliminator is a rubber encased impeller of elliptical cross section. It utilizes a combination of centrifugal force and reduced pressure to remove air that would affect the accuracy of measurements. The pressure inside the impeller is reduced to about 18 to 20 inch vacuum by a water powered aspirator mounted on top of the Air Eliminator. The sample flow stream (about 15 USGPM) enters through a stationary tube into the impeller, from which it is forced out through ports on its periphery. These are well located below the fluid level in the outer tank to avoid reentrainment of air. The flow then continues to the sensor unit.

Fig. 4. Sampling box in cyclone overflow for PSM sample.
Because the pressure inside the impeller is reduced, the sample may be lifted from a sample point several feet below the Air Eliminator. The distance between the sampling point and the Air Eliminator is kept to a minimum to reduce the pressure drop in the intake box, to minimize the time lag in measurement, and to avoid sanding in the lines.

Some removal of air already occurs in the sampling box (Figure 4), which is positioned in the overflow of the cyclone cluster. A plate in this box forces a diversion of flow, removing some air. The sample intake filter has been placed in the upward stream of the pulp. There, a strong turbulence warrants a sample that is a representative for the total flow.

THE PSM MEASUREMENT RANGE

The PSM measures the total size distribution. It assumes a similar size distribution for all ore types in one mine, which would merely shift with the hardness of the ore. The larger the shift, the greater the error from standard conditions. The accuracy of particle size measurements is directly affected by the choice of coarse endpoint of the measurement band. Maximum accuracy is attained when the coarse endpoint is less than 33 o/o (cumulative) retained or greater than 67 o/o (cumulative) passing. Greater accuracy is thus obtained when coarser reference screens are selected.

Earlier PSM models include the PSM 100, 200, and 300. For the 200 the measurement band was greatly enlarged over that for the 100 (see Figure 5), allowing finer grind to be measured. The band for the 300 and 400 is identical to that of the 200. For example, if a 150 mesh screen is chosen as reference, the measurement band is 95 o/o to 45 o/o (cumulative) passing for a PSM 100; for a PSM 200 the band measurement has been expanded to almost 100 o/o to 45 o/o (cumulative passing). For Cuajone, the cyclone overflow should operate between 8 to 10 o/o + 65 mesh. The measurement band for a 65 mesh screen is dictated by Figure 5 to be 0 o/o to 50 o/o (cumulative) retained; it may, therefore, be used as reference screen.

CALIBRATION

Before the PSM could be put on stream, a thorough and accurate calibration of the two pairs of sensors was pertinent. Samples were taken from the slurry as it passed the PSM sensor unit, while simultaneously Gamma and ASN values, for respectively the size and solids channels, were generated by the digital control unit. These values represent loss values (attenuation) in decibels (dB) of the ultrasonic signals, transmitted by one of the transducers in the sensor unit. The mathematical approximations relating to the attenuation of the ultrasound signal to the Percent Solids Material and Particle Size are respectively:

\[
\text{Percent Solids} = S_0 + S_1(\text{ASN}) + S_2(\text{ASN})^2
\]

\[
\text{Size} = P_0 + P_1(\text{GAMMA}) + P_2(\text{GAMMA})^2
\]
A least square fit is used by the PSM digital control to generate the pertaining S and P coefficients of the curve. The PSM system is calibrated by direct comparison of PSM

![Diagram](image)

**Fig. 5.** Particle size measurement range.

reading to screen analysis of the samples. Accurate calibration depends almost entirely upon accurate screen analysis. Each size fraction was, therefore, weighed on an analytical balance. For the PSM system accuracy is one percent or better; a goal of one-half percent error in screen analysis should be set. To achieve this a combined wet and dry screening method is used. Initially, the samples are wet screened at 200 mesh to remove the fine particles that could agglomerate, blind the screen opening, or adhere to coarser particles. Both the plus and minus fraction are dried, which, subsequently, is followed by dry screening of the coarse fraction.

**THE DEVELOPMENT OF THE CONTROL STRATEGY**

To develop a control strategy that would work for the Cuajone grinding circuit auxiliary equipment was installed, which might not be required for ultimate control configuration. Later, this equipment could be eliminated.
On top of the cyclone cluster a pressure transmitter was installed and wired to a digital MICON multiloop controller. Other transmitters included non-intrusive, ultrasonic doppler flowmeters to measure the head water flow, the sump water make-up flow and the cyclone feed flow. An Ohmart ultrasonic density gauge to measure the density of the cyclone feed flow already existed. Its signal was now fed to the multiloop controller. This signal in combination with that of the Ramsey belt weightometer, measuring the fresh feed to the mill, aided in continuous display of the mill circulating load. The mill power was also recorded continuously, a real benefit to early observe any sign of overloading.

![Graph showing corrected efficiency curves for cyclones of mills 2D and 2E at 280 t/h.]

Several bottlenecks were eliminated during strategy development. Firstly, the sample box positioned in the cyclone overflow proved too small and was replaced by an enlarged copy. Secondly, some modifications on piping were made to allow better flow measurements of water entering the cyclone underflow bin. Furthermore, fear of sump overflow creates the tendency to operate the cyclone feed pump near empty; sump level control without the aid of instruments is difficult. For good cyclone efficiency, however, a constant sump level is pertinent to provide a constant cyclone pressure and, consequently, a constant cyclone split. During strategy development, the sump level was raised and held constant; accordingly, the cyclone pressure increased somewhat, because more water was pumped through the cyclone. As a result, the backpressure from the plate in the sample box to the overflow bin mounted and caused spillage from the cyclone overflow bin. In an attempt to overcome this problem,
the lip of the overflow bin was raised three inches, but the remedy proved to be insufficient. Then the plate was removed from the sample box rendering the PSM intake sample non representative of the total flow. Finally, it was resorted to raise the cyclone cluster two feet to enlarge the overflow bin. Simultaneously, the pump speed was increased to process additional sump water make-up flow, which resulted from maintaining a higher and constant sump level. This water leaves the circuit at the cyclone overflow; the split at the identical mill throughput became finer. We proved this in a series of tests between two adjacent mills. For two consecutive days the feed to one mill (mill 2D) was maintained at 280 tph; the feed to the test mill (mill 2E) was raised 15 t/h every one hour and a half, starting from 280 t/h, and up to about 360 t/h. On the first day the ore was hard; on the second day it was soft. Samples were taken from the cyclone overflow, underflow, and mill discharge to calculate the circulating load, the cyclone efficiency and size fraction + 65 mesh in the cyclone overflow. Figures 6 and 7 illustrate the differences in cyclone efficiency. In the first Figure, the efficiency curves were drawn for a throughput of 280 t/h for each mill. It may be noteworthy that for each day the 50 o/o cyclone split was 30 micron finer for mill 2E. The second Figure indicates that for approximately similar efficiency curves the tonnage for mill 2E was 30 t/h higher for day one, and 60 t/h higher for day two.

Fig. 7. Corrected efficiency curves for cyclones of mills 2D and 2E.

THE CONTROL STRATEGY CHOSEN FOR CUAJONE

Without sump level or particle size control, large variations occur in cyclone pressure and cyclone feed flow. A near empty sump causes the pump to surge. Obviously, the particle
size cannot be constant. Actually, it varies significantly, although the percent solids in the overflow remained fairly constant. This is demonstrated in Figure 8. For reliable and constant cyclone operation a constant cyclone feed flow at constant cyclone pressure is important. To achieve this, one of the first steps in strategy development was to control the sump level. This can be done in several ways. A flow transmitter was fitted on the pipe for

Fig. 8. Normal Cuajone control, no size, nor level control.
reclaimed water that enters the pump sump. The fresh water flow to the sump was not monitored but kept constant. Level control was established by variations of the reclaim water flow, a signal being transmitted from the level indicator to the valve on this line via the multiloop controller. The desired sump level was manually set at the controller. In one attempt sump level control was linked with the measurements of the particle size in the cyclone overflow. If the size appeared too coarse the reclaim water flow was increased. This, in turn, resulted in a decrease in fresh feed to the mill, because of a rise in sump level, which would not be constant but vary continuously. The rate of response to a change in cyclone feed dilution proved to be very rapid, making this system too dynamic. The mill feed rate could not adjust quickly enough to the changes in sump level.

At present, the fresh mill feed is directly controlled by particle size in the cyclone overflow, as indicated in Figure 9. Less feed is called for when the size is coarse. The feed receives a remote signal from the process output of the particle size and changes accordingly. Particle size control is automatic; its setpoint is set manually, sump level control is now being established in a different loop. The valve on the reclaim water line is directly connected with the sump level transmitter, maintaining the level at a constant, manually preset, level. Figure 10 illustrates the influence of level control at a constant level on the overflow particle size. On this chart the feed was not controlled by size, but set manually at a level predetermined by the reading of the particle size monitor.

![Diagram](image)

Fig. 9. Detailed flow sheet of the mill circuit that illustrates the control strategy.
In cascade mode, the amount of mill head water that enters the cyclone underflow box is rationed by the amount of fresh feed entering the mill. A normal ratio of water to feed is about 0.28. A part of the recorder chart to show feed controlled by size is depicted in Figure 11.

Fig. 10. Level control only, no size control.
From Figures 10 and 11 it is evident that at least sump control is desired to maintain a fairly constant particle size in the cyclone overflow. With level control only the particle size is seen to drift a little over a larger time period. This may depend on the hardness of the ore, or on its coarseness. Control of particle size avoids this drifting altogether as seen in Figure 11. Fluctuations in particle size are $\pm 0.5 \, \text{o/o}$ absolute from the setpoint.

Fig. 11. Size and level control.
PROBLEMS

Some problems became apparent during the test period. One stems from the fact that the Armco PCDP (Process Control Development Package), or MICON multiloop controller, cannot be linked directly to the belt tachometers regulating the belt speed. The PCDP output signal is digital, rather than analog, which is needed for the tachometer relay switch. The output signal of the DCDP, therefore, has to be fed to the feeders via the existing PEIC (Periodic Electronic Integrating Control) unit. This unit was found to respond much slower than the PCDP and was also less accurate. For example it had not yet dispatched a speed signal to the feeders while the PCDP was already indicating a following change. Effective control at Cuajone at present, is only possible after incorporation of special control techniques to delay signals in combination with feed forward control.

![Graph showing performance metrics.](image)

Fig. 12. Economic performance.

Another, more serious problem, is scaling. Like in any high lime circuit, carbonate scaling occurs on all equipment in touch with the pulp flow. Within three to five days a scale has formed at the surfaces of the sensors that interferes with the attenuation of the signal. Scale formation reduces the intensity of the transmitted signals and hampers its reception. Consequently, the monitor will display a higher percent solids, and worse, a larger particle size than actual, incurring loss in metal production. A program will have to be initiated to remove the scale twice a week to ensure maximum production. Normally, carbonate scales are removed by concentrated hydrochloric acid. Unfortunately, the epoxy resin which is
encapsulating the sensor crystals, is affected by this acid. Another solution to the problem seems to be dilute phosphorous acid that softens the scale to be scraped off by e.g. a knife.

ECONOMIC JUSTIFICATION

An economic justification of the project might clearly be demonstrated by a higher mill throughput at similar particle size in the cyclone overflow. At this stage, most time was spent in the development of the control strategy and in fine tuning. Actual studies to prove a possible increase in mill throughput or Cu-recovery increase were not yet undertaken. One improvement though, manifested itself in the test period:

Cyclone feed sump level control and increased pump speed could result in an increased mill tonnage of more than 30 t/h (See Figures 6 and 7). To achieve this no particle size control would be required. For grinding at maximum mill load, while obtaining the desired particle size, control of the latter would be mandatory. Figures 12 and 13 illustrate potential losses in revenue because of overgrind or undergrind, if no size control were applied. Variation in particle size could be limited to a narrow band, thus allowing maximum recovery.

The control charts prove that, indeed, a more stable grinding circuit performance is achieved using the PSM. Improved operational flexibility was also apparent, especially because of the use of the multiloop MICON controller. Any change in setpoints was reflected immediately in power output, valve settings and operational levels.

Fig. 13. Revenue and particle size.
2.2. DISTRIBUTED DIGITAL CONTROL AT CUAJONE

B.J. Huls, P.L. Dunstan, S. Donges


ABSTRACT

The commissioning of a Distributed Digital Control System for the grinding circuit at Cuajone, was completed by mid August, 1985. Development of a successful strategy, excellent operator acceptability and the visual aspects offered by the display of the central console, improved mill throughput significantly. The inherent features of a digital system enhanced the flexibility in grinding control and allowed the mills and classifiers to be operated at maximum capacity. This paper discusses the strategy development and the production restrictions encountered, and evaluates the overall performance of the digital control system.

INTRODUCTION

In November 1976, The Cuajone Concentrator, located on the western slope of the Andes in Southern Peru, was commissioned to treat a nominal 41 kt/d of ore. A systematic elimination of production restrictions has resulted in a continual increase of concentrator throughput to 48.7 kt/d. This was before the grinding control was installed.

BACKGROUND

Ore reserve analysis indicates that the ore grade will drop from its current level of 0.9 % Cu to 0.85% Cu by 1988, where it will remain for the life of the mine. Several testwork programs were initiated to study new equipment designs and process strategy to offset the falling ore grade and to maintain the copper production. This would require a plant expansion to at least a nominal 59 kt/d of ore. One phase of the testing program studied the use of an Armco Autometrics Model 400 Particle Size Monitor (PSM) to control the product size of the primary mills. The following conclusions were made as a result of the testwork:
- A mill being controlled by a PSM will produce at least 5 percent more tonnage than a mill operating without on-stream size control.
- Without size control, the flotation feed size varies about ± 2% from its target; with size control, the deviation is ± 0.5%
- Significant milling rate increases are very sensitive to the physical condition of the cyclone feed pump, the cyclone classifiers, and the shell and head liners, plus variances in grinding ball load.

The use of digital control in the early stages of the development work showed the benefits of digital control over an analog system. Details of the development of the control strategy were described by D.M. Podobnik et al. (1984).

OBJECTIVES AND SPECIFICATIONS

The criteria that would be used in the selection of the distributed digital control (DDC) System for the Cuajone concentrator would include the following requirements:

- The system would measure field variables whenever possible to reduce the dependency on any model.
- Each grinding mill would be subject to dedicated control; slaving of other lines from one controlled line would be avoided.
- The system would be completely digital for quick response and to eliminate the hardware required for a similar analog system.
- The availability would be enhanced by functional distribution and redundancy.
- The dependency on a central console station would be eliminated through independent loop controls by various microprocessors.
- The individual operator stations would resemble the familiar analog lay-out as closely as possible.
- The individual control modules would make it feasible to adapt major strategy changes through configuration of control loops.
- The system would be backed up by another power supply and by a series of batteries in case the main power supply failed.
- The system would have a high degree of self-diagnostic capability.

After considering various systems, the Bailey Network 90 was selected.

SYSTEM DESCRIPTION

The Bailey Network 90 distributed digital control (DDC) system includes two central consoles (OIQ), a panel containing 90 digital control stations and 16 digital indicator stations grouped for individual mill and section control, a printer, data storage, historical trending and graphics display. Signals transmitted from the field arrive at termination units located in each of the four process control units (PCU) in the central control room. Within a PCU, several controllers containing microprocessors are dedicated to control the operation of an individual mill. Individual critical process loops can be spread over several
modules. In case one fails, only a small part of automatic control is lost. Since the operator stations on the panel are hardwired from the termination unit independently from the controller modules, manual control will always be possible. The operator stations form a fail-safe backup unit. Communication between the process control units and the central consoles is established through two independent plant loops. For actual control, the central consoles would not be absolutely necessary. They proved, however, their value in monitoring the grinding operation as they provide detailed data that enables the operator to understand the grinding system much better than before. Figure 1 illustrates a trend display for a combination of operating parameters.

Figure 2 displays a graphic in use for each mill. It updates the field values every minute. Actual changes in setpoint values may be done from this graphic. This form of OIU also has configuration capabilities that allow the entire system to be configured from one location. The OIU also allows the saving of module configurations on diskettes. These diskettes then provide a back-up to the modules own nonvolatile memory, in the event the modules need to be reconfigured.

![Image of a trend display]

Fig. 1. An example of a trend display

**Power Backup.** To protect the DDC system from catastrophic power failure it is supplied by one of three alternative power sources. The normal power supply is received from a motor control centre and converted to 110 V dc as it passes through the uninterruptable power supply (UPS). It is then reconverted to 110 V ac, 60 Hz before it enters the power
entry panels in the PCUs. A third source of power is available from another feeder that bypasses the UPS and is fed directly to the power entry panel after being transformed to 110 V ac.

Fig. 2. An example of a graphic display, here showing the circuit of one grinding line

Configuration. Actual control is distributed among the various microprocessors in the controller modules, although a computer may be hooked up to the system to make all the decisions in respect of strategy. It is not a computer program that regulates the control strategy, but a configuration of functional blocks, the function codes of which are related to those used in analog control. No additional hardwiring is required to compose different strategies by varying the sequence of the function code blocks. Flexibility in the control strategy compared to an analog control system is therefore enhanced significantly. Features like adaptive tuning, setpoint ramping, and algorithm capability also add to the flexibility.

OPERATING STRATEGY

Before the testing with the PSM began, it was thought that the grinding circuits were being operated at their capacity limits. The testwork results achieved by the PSM and the digital controller clearly demonstrated that this was not the case and that the development of a much more sophisticated control strategy was required.
Plant Water Ratio Control

The plant water system for the Cuajone Concentrator is different from most other plants because two independent water systems - one fresh water and one reclaimed process water - are used.

A simple water control strategy based on a constant consumption of fresh water was not possible because of the unusual requirements of the hydroelectric plants. This demand causes major fluctuations in the freshwater flow rate to the plant reservoir, and the limited reservoir capacity requires that the freshwater takeoff be varied as well. The total water demand in the plant is fairly constant, so the reclaimed process water consumption must be adjusted to compensate for the changes in freshwater consumption.

**Fig. 3. Dust slurry water control.**

The water control strategy calculates the total water that is required to be added to the mill feed from a steady-state mass balance model and divides it between dust slurry water, an amount of which has to be used, and a discretionary amount of fresh and process water. Depending on the situation in each reservoir, the discretionary water addition is varied between the fresh and the process water. Due to problems with the measurement of mass flow to the cyclones, we have been unable to use the steady-state mass balance model to
control mill solids. Consequently, the operator now enters a setpoint value on a ball mill head water station. Actual measured values of ball mill solids are sent to the control room, and this setpoint is altered accordingly.

**Dust Slurry Water Control**

Dust slurry water is added to each mill scoop. The strategy demands that all dust slurry water be consumed and that the addition be distributed equally over all eight ball mills (see Figure 3). If the ratio control described previously is designed to minimize freshwater consumption, any freshwater addition not required by the dust collectors must also be minimized. These requirements will be met by automatic control valves in the dust slurry water lines to each mill so that constant levels are maintained in two dust slurry water surge tanks installed in the grinding area. If the incoming slurry water flow from the dust collectors drops and a surge tank level begins to fall, the valves will slowly close to a preset minimum flow. If a level falls to a low-low setpoint, another valve opens and adds sufficient fresh water to the surge tank to ensure enough water to wash the coarse cyclone underflow material into the scoop box. When the incoming dust slurry flow normalizes and the level in the tank rises, the fresh water addition is stopped. Krohne magnetic flow meters monitor the dust slurry water flow to each mill.

Fig. 4. Typical ball mill circuit and steady-state mass balance model.
Ball Mill Water Control

The percentage of solids of the pulp in the ball mill has proven to be a critical variable in grinding of Cuaone ore and needs to be controlled over a narrow range. Control of this variable begins with a mass flow measurement of the cyclone feed. The tonnage of solids calculated from the mass flow is equal to the total mill discharge solids. If the measured flow of cyclone feed dilution water is subtracted from the total water in the circuit as determined by the mass flow measurement, it will give the tons of water in the mill discharge. The percent solids in the mill discharge can then be calculated and controlled by variation of the water added to the ball mill feed.

Figure 4 details the steady state mass balance model being used in the control system. There is some concern over the assumption of steady-state conditions. Cuaone ores exhibit wide variation in hardness. As the ore hardness changes, the operating characteristics of the grinding-classification circuit will also vary widely. The circulating load in the circuit will be affected. This situation, which does occur from time to time, invalidates the steady-state assumption and correct use of the model.

One large drawback is that, for the present, the steady-state mass balance model as previously described cannot be used. This model requires input from the percent solids and flow measurements on the cyclone feed line. The density measurement was delayed until early 1986 when the 0.46-m cyclone feed pipes were exchanged for 0.51-m pipes. For this pipe diameter, new Ohmart Nuclear density gauges were purchased.

Polysonics ultrasonic flowmeters were installed on the cyclone feed line to measure the cyclone feed flow. Unfortunately, the meters produce inconsistent data. As a result, the ball mill water demand control, based on the calculation of the mill discharge solids, cannot be implemented. Further, calculation of the cyclone underflow percent solids and the circulating load ratio cannot be made.

Ball Mill Product Control

Control of the ball mill product, or the size distribution of the cyclone overflow, is the heart of the control scheme (Figure 5). Process water is added to the cyclone feed pump sump, and the addition rate is controlled by a sump level setpoint. A sample of the cyclone overflow product is sent to the particle size monitor, which determines the percentage of a specified size and the pulp density. The actual size reading is compared to a size setpoint and the control station then regulates the speed of the belt feeders. These change the tonnage rate to the mill until the size setpoint and process variable have come together.

The output of the feed station is directly checked against the rpm signal (0.5 rpm) returning from the belt tachometer in the field. For a deviation exceeding 0.25 rpm, a pulse positioner will transmit pulses through a relay to the electric servo motor on the hydraulic motor control, which alters the belt speed.

The time between the size measurement by the PSM and the effect of a change in belt feeder speed is about six to eight minutes. It was necessary, therefore, to add a derivative
function to the controller of the size station. The better anticipation of diminishing error between the process variable and setpoint value allowed faster normalization after an upset in size measurement. Cyclone classifier overflow product size is controlled to within ± 0.5 % +210 μm from its setpoint value. Adaptive tuning, a feature provided by the system, ensures the size specification to be maintained within the range. On occasions, this deviation becomes larger, and despite lengthy efforts to finetune the size and feed controllers, size control could not be returned to specification. Closer inspection of the feeders revealed that the gear boxes of the electroservo motors could not stand up to the continuous changes in desired belt speed. The gears strip and have to be replaced by a new set. Electrical variable speed drives for the servo would eliminate the problem, and this solution is being considered. With the new control system, we have developed the flexibility to use several modes of operation. We can operate the feed station in the automatic mode where throughput is independent of size control, or in cascade where the feed rate setpoint is determined by the output of the mill size station.

![Diagram of Ball Mill Product Control](image)

Fig. 5. Ball mill product control.

Before this value enters the feed station as setpoint, the operation is checked for any problem situation such as a PSM not in operation, a ball mill or classifier overload situation, a mill start-up, or an actual mill power draw below the critical setpoint. The setpoint output of the feed station will be equal to its input when the PSM is out of operation. In a mill overload condition, the feed setpoint will drop slowly depending on the
seriousness of the problem. When starting up a mill, a configured start-up button may be actuated from the central console that would ask for 365 t/h feed until the actual size of the mill product comes within 4% absolute from its setpoint value. The feed setpoint value will be lowered immediately once the power draw arrives at its critical minimum point.

System Constraints

During the testwork with the PSM and the digital controller, it was found that under certain conditions the normal control strategy would produce undesirable situations. As a result, certain constraints were built into the operating strategy to prevent these from occurring.

Tonnage Constraint. The head grade of the concentrator feed is variable, and large variations can be experienced over short time periods. If the control strategy was simply allowed to maximize tonnage, flotation efficiency may be seriously affected or other operating sections downstream might become overloaded. To prevent such problems, a maximum tonnage constraint will maintain the total tonnage produced by four mills below a setpoint maximum. This constraint would originally be an operator input based on the operating conditions in other parts of the process. To avoid an override of actual size control, the variation in the size setpoint will be very slow so the mill can keep up with it. The size setpoint will not vary when the section throughput is within predetermined boundaries and when deviation in actual size between each of the four mills is within 1.5% +210 μm. It will also vary if one of the mills experiences a large upset, such as a sanded pump, feeder problems or an overload condition.

Cyclone Underflow Density Constraint. When the hardness of the mill feed increases, the circulating load also increases and can do so up to the point where the hydrocyclone classifiers become overloaded, causing the underflow to "rope". The approach of this condition can be anticipated and steps taken to correct it before an overload occurs. The calculated cyclone underflow percent solids would indicate the approach of such a condition. When a certain preset constraint limit is reached, the control system will react by either reducing the amount of water being added to the cyclone feed pump sump, which reduces the feed rate to the mill, or by an immediate cut of the feed rate at a predetermined rate.

Mill Power Constraint. An early warning of a ball mill tending to overload is the rate of change of power draw (dp/dt), expressed in kW/min. This rate is calculated in a controller module using the quarter second module scan. In normal operation, this value is higher than its setpoint. When they are equal, that grinding mill circuit enters a soft constraint. It freezes the fresh feed addition to the mill. If the actual rate drops to 2 kW/min below its setpoint value, the overload is critical. The feed will be cut at a slow pace, depending on the drop in power. Incidentally, the setpoint is not a constant value, but changes reciprocally with the power draw.

On recovery from a critical situation, additional feed will enter the mill when the power draw again is more than 2 MW and if dp/dt once again is positive. A mill upset is prevented because more feed will enter the mill at a slower than normal rate until actual size matches its setpoint value. When a mill enters a power constraint, the mill size station reverts to the tracking mode: it tracks its own output and setpoint values, which then do not
change. The size station returns to its original mode of operation after the overload condition has been cleared. For normal size control, this would be the automatic mode.

The mill size and feed control stations operate in the cascade mode for full constraint control. If the plant tonnage constraint is limiting the maximum mill throughput, the mill size control stations receive an external setpoint value from the section tonnage control station. On recovery from a mill overload condition, this value will be set equal to the actual process variable. Subsequently, the setpoint value will rise slowly to restore the desired section tonnage rate.

**Cyclone Feed Pump Current Constraint.** A fourth constraint implemented in the strategy is the cyclone feed pump current. It sometimes happens that, after a pump service, the spacing between impeller and casing is set too close, which causes a higher power draw. To protect the motor from overload, an upper current limit value has been configured. On reaching this value, the sump level is lowered to relieve the motor.

**OPERATING EXPERIENCE**

On arrival in early March, 1985, the Bailey engineer began immediately with on-line control to one grinding line.

After operating in the manual mode for sufficient time for the operators to become accustomed to the new digital stations, the water stations were put into the automatic mode. The operators showed excellent response to the new station layout. Following this, the feed signal from the Ramsey belt scale and the signal from the PSM were transferred from analog to digital display. The crossover from the analog system to the DDC was conducted with the minimum downtime of any operating equipment.

Before the feed station was operable, using the new control strategy, it was necessary to change the 1 to 1.9 V on the tachometer output to a signal of 4 to 20 mA. Relays located on the termination control unit in the PCU transmit the pulses to the electroservo motor on the hydraulic drive of the feeder belts to alter belt speed as desired.

Isolators were installed on the cyclone feed pump current signal and on the signal from the sump level probe to enable the Bailey system to accept these signals. The 2.1 m long Bindicator sump level probes were replaced by 3 m probes in order to reach the bottom of the sump.

In this order, the signals from the field were transferred to the new system. Toward the end of the commissioning period, the uninterrupted power supply (UPS) was brought on-line and has since worked satisfactorily.

Several failures were encountered in field instrumentation and in the DDC equipment. The majority of the failures in the DDC equipment occurred during the start-up period between April and May 1985. A failure in the power supply in PCU 1 damaged nearly all the
modules installed in that unit. The other major failure was that of one OIU. The Winchester drive and drive board needed to be replaced.

The controller modules developed many EEROM errors. Bailey Controls, in the meantime, frequently updated the software in these controllers, which appeared not to be compatible with the existing system. Replacement of chips in the controllers, therefore, would not always guarantee that the modules would work. Other problems with the Network 90 system might be traced back to what is now believed to be a poor environment in which the system operates. Steps have been undertaken to overcome the dust, the high temperature, and the low humidity problems. Since the temperature has been maintained to below 27 °C, many of the software problems have disappeared.

OPERATING RESULTS

Results are shown for the periods in which no size control strategy had been developed, size control strategy was in place without plant-wide control, and when plant-wide control had been adopted. The histogram in Figure 6 illustrates the percentages of change of various operating conditions, using the first period as base.

![Histogram showing changes in mill throughput, size distribution, sulphide copper recovery, power consumption, and ball consumption.]

**Fig. 6.** Effect of control on plant operation.
Mill Throughput

An analysis of the mill throughput data, prior to installation of the PSM, shows that 252 t/h per mill was achieved consistently between 1980 and July 1983. The introduction of a single PSM in July 1983 produced a significant increase in mill throughput to 260 t/h. During the period since the complete control system has been in service, the average mill throughput has been 271 t/h, which denotes an overall increase of 7.5%.

Mill Flotation Efficiency

The best parameter that senses the flotation efficiency is the indicated copper sulfide recovery as determined by the metallurgical balance around the copper flotation plant. It shows that there has been a reduction of about 1% since installation of the grinding control strategy. A factor to be considered, however, is that with the grinding circuits being pushed up to their apparent maximum tonnage, operation of the flotation plant has become much more difficult. In some cases, it has required different operating strategies. As adjustments are being made, the flotation performance since the changeover has been consistently improving. Copper sulfide recovery is expected to be maintained above 88.7% without further plant modification. Conclusively, the plant recovery will fall by 0.2% to 0.5% due to a reduction in flotation retention time and to the acceptance of coarser flotation feed size distributions.

Size Distribution-Flotation Feed

The reason for the increased coarseness of the product from the grinding circuit under control, from 10.1% to 11.8% +210 μ, is that several of the components within the circuit are overloaded and are imposing constraints on the control system. This prevents the production of a finer size distribution. The major factor is that the pipeline collecting the product from the cyclone overflow stream is full and thus reduces the amount of dilution water that can be added to the cyclone feed pump sump. Earlier testwork has proved the benefit of operation with full cyclone feed sumps. Once the constraints are eliminated, the size distribution of the flotation feed will be controlled at the same levels achieved before the control system was in use, but at a higher mill throughput.

Mill Power Consumption

The control strategy employed ensures a mill load up to the point where a rapid reduction in power draft occurs. As this point is being maintained more precisely, the energy efficiency has improved continually since installation of both the PSM and the control system. The reduction in unit power consumption is expected to be about 0.6 kWh/t of ore.

Grinding Ball Consumption

Although there was some expectation of reduced grinding ball consumption when grinding control was implemented, the data in Figure 6 indicates only a very slight trend toward lower steel consumption. To date, the character of the ore milled has had a much more
significant impact than any change in operating strategy. Further data is required before any
conclusion can be drawn about changes in ball consumption.

There was concern that one of the problems of introducing the new control strategy and
hardware would be a low acceptance by the hourly operating staff. Therefore, much effort
was put into exposing all personnel involved in using the system to as much formal training
as possible. Emphasis was placed on how the system would assist them in carrying out their
duties. This has been successful, with the result being a high level of acceptance by all
operations personnel. The real-time data provided by the system has not only allowed the
operator to know what the actual operating conditions are, but also has gradually given him
an understanding of the dynamic conditions that exist in the grinding-classification circuit.
We are convinced that the economic benefit of the development of control operators to the
point where they understand what a grinding-classification circuit is all about is very
significant.

REFERENCE

presented at the Fourth Arico Grinding Symposium, Chile.
2.3. MECHANICS OF COLUMN FLOTATION OPERATION

R.L. Amelunxen, R. Llerena, P. Dunstan, B.J. Huls


ABSTRACT

A single four feet diameter cell, treating a Cu/Mo rougher concentrate, has been fully instrumented and linked to a Distributed Control System. This has enabled the writers to further understand the mechanics of operation of the cell, particularly that of the "TRANSITION ZONE". This is the volume between the feed port and the pulp-froth interface. The results have shown a need for the re-designing of the control loops whereby the Transition Zone volume (or pulp level) becomes the single most important control parameter to achieve the desired recovery. Air consumption, previously thought to have played this role, is now kept at precoalescence levels for maximum efficiency.

INTRODUCTION

Since the advent of column flotation, strategies regarding the control of column cells have been few. The strategies used can be summed into two categories; a) one that aims to achieve an improved grade of concentrate and b) one driving towards maximizing the economic benefits of both recovery and grade. The former would be employed when the column is used as a final cleaner stage, where shortcomings in the recovery could be absorbed into a circulating load. The latter, would be utilized when a column is in a cleaner circuit or a roughing stage of a separation circuit where the recovery plays just as an important role. This work aims to shed more light into the mechanics that govern the operation of these units, using strategy b).

BACKGROUND

Previous work by Amelunxen (1) suggests a Washair Interactive System, whereby the spraywater requirement is precalculated and adopted to yield the set point of the air holdup. Regulation of the air holdup would then cause the degree of "pulling" of the concentrate to change. The principle generally works well in that increasing the air causes an increase in the concentrate production, hence an increase in the washwater consumption. The instrumentation diagram for this strategy, is shown in Figure 1.
With time, flaws were found in this strategy. For instance, the column could not recognize the occurrence of "impending" holdup. In these situations, rapid increases of holdup occur, without changes in the air addition rate. This condition is characterised by low pulp levels (i.e. 4 to 6 feet below the overflow lip) resulting in a lack of froth buoyancy to carry the concentrate over the lip of the cell. An accumulation of the floatable species forms below the pulp-froth interface as a mushy conglomerate of air and pulp. The effect of this accumulation at the interface is to increase the air holdup resulting in the control strategy asking for less air. This change causes the cell performance to diminish even further.

It is often found that the air-interactive control loop cannot comply with the degree of "pulling" required and is therefore not capable of meeting the washwater demand. This is usually the case when air coalescence occurs while the control strategy is demanding the holdup to increase.
Two new points in the control strategy became rapidly apparent from the way in which the operator responded to these symptoms. In the case of approaching “impending” holdup, the level of the pulp is raised to increase the rate of concentrate removal. In the case of coalescence, the column air is allowed to reach pre-coalescence levels and not surpass it by using an algorithm on the holdup/air flow rate loop.

The need to further understand the column operation prompted a series of tests with the aim of studying the role of the Transition zone in the column operation. Considering that the strategy used to control "impending" holdup was to raise the level of the pulp, testwork was geared towards the study of pulp-froth interface position versus column performance.

**TEST PROCEDURE**

**Process description:** Cuajone's concentrator has a capacity of 57,000 tpd with two parallel circuits treating equal tonnages. The production of each cleaning circuit is 800 tpd copper-moly concentrate assaying 28% Cu, 9.5% Insolubles and 1% MoS2. The one-column test circuit is fed by about 10% of the primary rougher concentrate production (approximately 7 to 12% Cu). The remaining 90% feeds the two conventional cleaning circuits. Two stages of cleaning are followed by a cleaner-scavenger which produces a final tail. Cuajone’s interest in column technology was prompted by the need to lower the insoluble content of the final concentrate. This translates to immense savings in the smelting costs.

**TEST A)** Level versus performance: In this series the column was operated using the Wheeler (3) strategy, i.e. with a BIAS loop and a washwater-level control loop. Samples were taken of feed, concentrate and tails at varying levels of the pulp-froth column interface. Pulp flows, air, holdup and % solids were measured to calculate a mass balance and a general assessment of the cell performance to be made. Copper and insolubles assays were made on the feed concentrate and tailings streams. The results appear in Table I.

<table>
<thead>
<tr>
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<tr>
<td>TEST            : TOP LEVEL VS RECOVERY</td>
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<td>FEED FLOW       : 176 GPM</td>
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<tr>
<td>TAIL FLOW       : 195 GPM</td>
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<td>AIR FLOW        : 67 SCFM</td>
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| FT    | % HOLD-UP | GPM | % WATER | FEED ASSAYS | CONC ASSAYS | TAIL ASSAYS | LBS CONC/ |
|-------|------------|-----|---------|-------------|-------------|-------------|FIT AIR    |
| 4.00  | 13.33      | 58  | 13.31   | 16.71       | 33.96       | 24.21       | 5.36      | 9.85      | 1.14      | 70.90     |
| 3.45  | 13.73      | 48  | 13.31   | 15.10       | 32.97       | 24.56       | 4.96      | 9.54      | 1.09      | 73.80     |
| 2.56  | 9.14       | 67  | 13.41   | 15.33       | 33.56       | 20.90       | 4.67      | 9.60      | 1.13      | 75.71     |
| 1.50  | 14.36      | 121 | 13.50   | 14.78       | 31.38       | 13.70       | 3.97      | 8.12      | 1.25      | 80.82     |
| 0.97  | 15.20      | 103 | 13.70   | 17.56       | 29.99       | 13.35       | 4.02      | 9.34      | 1.62      | 81.60     |
| 1.13  | 14.19      | 42  | 13.46   | 17.78       | 31.28       | 23.66       | 3.97      | 10.03     | 1.53      | 80.73     |

Table 1 Summary of results of test A.
TEST B) The results of TEST A prompted a further investigation of the column behaviour in the Transition zone (the volume between the feed inlet and the pulp-froth interface). This test was done twice to verify the results.

Fig. 2. Dimension of the test column.

A profile of the column’s interior was taken by sampling the column at various depths as shown in Figure 2.

These samples were assayed for copper, insolubles and % solids. The changes in the copper recovery in the Recovery Zone were calculated by analysing a mass balance of a constant
volume at each sample station. An attempt was made to use the same technique in the Transition Zone, but interference by the incoming feed distorted the results. The recovery changes in the Froth Zone were calculated by knowing the product concentrate weight, its copper content, and working backwards towards the Transition Zone. Using the steady state equation for mass, assuming no rate of accumulation in the froth zone, recoveries in each of the sections can be calculated. These results are shown in Figure 3. The recovery profile curve, as can be seen, fits well except for the transition zone. The erratic results in the transition zone prompted one further test.

![Graph](image)

Fig. 3. Results of the profile test.

**TEST C** This test aimed to resolve the question of whether the flotation rate of mineral particles in the transition zone is as high as in the recovery zone as indicated by TEST B. This test was performed immediately following TEST B. It consisted of running the column with no froth bed. That is, the recovery of copper would be a reflection of the performance in the Recovery plus Transition Zones. The results of TEST B and C are shown in Table 2.
DISCUSSION OF RESULTS

TEST A)  Figure 4 indicates a strong \((r = 0.994)\) relationship between the position of the pulp-froth interface and the column performance. The higher level (i.e. the smaller the froth bed) the better the recovery. The slope indicates an improvement of 3.4 recovery units per foot of level. Likewise, the grade of the concentrate obeys the grade-recovery curve and is therefore decreasing with the increases in the level. The question here is whether the improved performance is due to a reduced froth bed or to the increase in size of the Transition Zone. The belief among operators is that a deep froth bed would result in a higher degree of cleaning as the water spray would have more of a froth volume in which to clean the concentrate. These tests have indicated that the cleaning is more as a result of the grade-recovery relationship. Undeniably, liberated gangue would have a difficult

COLUMN CELL 48"

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<th>TOP LVL</th>
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Table 2  Summary of results of tests B and C.

time reporting with the concentrate by the principle under which BIAS controlled columns operate: the gangue free washwater is the only source of water reporting with the concentrate.

TEST B)  This test was primarily designed to determine the flotation characteristics within the column. The Recovery Zone is relatively easy to expose in that there is an
unidirectional flow of liquid with relatively low mixing characteristics of the solid species. A distinct near-linear recovery pattern emerges with column depth.

This gives a net downwards liquid flow with a rate that is generally quite small (i.e. 1 to 5% of the feed flow). In this zone we also find an upward flow of solids already experiencing the fluid drag of the BIAS portion of the washwater. Therefore the Transition Zone is one of relatively little activity except for it being the recipient of the column feed. The large degree of mixing that occurs in this zone due to the incoming feed, net downward flow of water and upward movement of the solids makes it virtually impossible to calculate a recovery characteristic.

The Froth Zone is one where the flow patterns are not well understood. Above the froth column the washwater is added, then it divides into two streams, the downwards flowing BIAS water and the water leaving with the concentrate.

![Graph](image)

Fig. 4. Results of Test A. Column performance with variation of level

As in conventional flotation, there is a loss of recovery in the froth zone. This loss has been determined as approximately 0.9 recovery units per foot of column depth. This partially explains why the column flotation recovery improves with increasing pulp level. The results
of this test, as shown in Figure 3, indicate that the recovery in the Transition Zone is characteristic of that of the Recovery Zone. However, the inconclusive data prompted the running of TEST C.

**TEST C** The results of this test indicated that the Transition Zone is working as a mere extension of the Recovery Zone. Figure 5 shows the recovery profile from TEST B, and in addition it depicts the copper recovery in a column with no froth bed (i.e. 78.7%), as obtained in TEST C. If an asymptote is drawn in the copper recovery profile curve and extended to intersect the zero depth coordinate, it yields a recovery of 77%. This confirms that not only is the Transition Zone an extension of the Recovery Zone, but the kinetics are also similar. This only applies to the column dimensions as tested. No data is available, at this time, which would indicate the maximum depth of transition zone that would function like a recovery zone.

**STRATEGY DEVELOPMENT:** Initially, the air-interactive system was used to control the column performance with regard to the grade-recovery curve. Deficiencies in that strategy prompted the operators to overcome the difficulties by adopting measures or programming algorithms that warned of undesirable situations. The level-interactive control strategy has emerged as a powerful tool to overcome all the deficiencies found with the other strategy. The options for its implementation are two-fold. It can be used as a stand alone control circuit, with the air at its maximum before coalescence, or it can be implemented as an air-level-interactive loop. This aims to keep pulp levels lower (if the air rate is well below coalescence levels) and works as air-interactive, but as changes in the feed prompts large increases in the air addition rate, nearing coalescence, the level-interactive loop comes into play. Cuajone ’s column presently operates the stand alone, level-interactive loop as it lacks a proper control valve for the air flow.

![Fig.5. Recovery vs column depth curve. Asymptote projects the recovery in the absence of a froth zone](image)
COLUMN PERFORMANCE WITH LEVEL CONTROL

The instrumentation for this control strategy is illustrated in Figure 6. It shows the Wheeler BIAS loop. It also contains the level-washwater control loop where the level is maintained with washwater additions. The third loop is the level-interactive loop. In this loop, information regarding the cell concentrate characteristics is essential, as it was for the air-interactive loop. If the relationship between the specific gravity of the concentrate pulp and the concentrate grade is known, it is possible to calculate the water content of concentrate overflowing a cell lip at the desired grade.

Fig. 6. Instrumentation diagram for level-interactive system
The principal conditions that are required for column flotation are that the tail flow rate must be larger than the feed flow rate:

\[ Q_t > Q_f \]

and,

\[ Q_{sw} = Q_c + \text{BIAS} \]

where \( Q_{sw} \) and \( Q_c \) are the flow rates for washwater and concentrate respectively, and,

\[ \text{BIAS} = Q_t - Q_f \]

In the level-interactive loop, air is still the driving force of flotation but, as in the Cuajone case, maintained at its maximum rate just below coalescence. The digital distributed control system initially calculates the washwater requirement, \( Q_{sw} \). The BIAS and \( Q_c \) are operator inputs and are calculated from various other measurements and inputs: a) from the operator, (in the future from an X-RAY analyser) the grade of copper in the feed. b) The recovery and grade desired. These conform to the established grade-recovery curve for that system and are pre-set within the computer. c) From field data, % solids in the feed.

The calculated \( Q_{sw} \) becomes the set point of washwater volume the column is required to maintain. This it does by comparing the actual washwater flow to the set point. If this flow is lower than the set point, it automatically raises the set point of the level. Since the recovery will increase with increased level of the pulp, this translates to a higher \( Q_{sw} \) consumption. The level is increased until the pre-set washwater demand is met. Figure 7 is an actual output of Cuajone's DDC Bailey 90 Network.

CONCLUSIONS

A strong relationship has been found between the pulp level and the performance of a flotation column. Decreases in the froth column depth will result in a significant increase in recovery. The investigations indicate the reasons for this are that the Transition Zone acts as an extension of the Recovery Zone and the reduction in the depth of the Froth Zone where the recovery decreases. The new control strategy that emerges from this investigation is a level-air-interactive loop as opposed to an air-interactive system. The benefits to be derived are a much more stable operation, which will result in higher concentrate grades and higher metal recoveries.
Fig. 7. Actual output of Cuaione 's DDC

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2.4. LEVEL DETECTION IN FLOTATION COLUMNS USING CONDUCTIVITY

C.O. Gomez, A. Uribe-Salas, J.A. Finch, B.J. Huls


ABSTRACT

A deep froth stabilized by a downward flow of washwater is used in column flotation to prevent the recovery of gangue particles by water entrainment. Froth depth or pulp level is a control variable influencing both grade and recovery. A common method of measuring level involves sensing the static pressure at some location below the level. The calculation of the froth depth from a static pressure reading involves the use of an equation whose parameters must be determined at the same operating conditions being used, and therefore, substantial errors result as a consequence of changes in slurry density, bubble loading and gas holdup. Electrical conductivity can be used to develop a method free of these sources of error. Large differences in conductivity have been measured for the froth and bubbling zones in a water-air system, as a result of their different water contents. Experiments completed in a laboratory column installed at Strathcona mill (Falconbridge Ltd), show that large differences in slurry and froth conductivity also exist when a third phase in the form of solid particles is present in the system. Conductivity profiles collected around the interface show that a gradient exist within each phase, a result with important implications in the design of conductivity probes to detect the interface level.

FLOTATION COLUMNS

Flotation is one of the most used methods to separate particles containing a valuable mineral species. The separation is based on the creation of an appropriate environment to promote the preferential attachment of particles containing the valuable component to small air bubbles which rise leaving the gangue mineral particles behind. Column flotation is a recent development in this area which offers some important advantages over conventional flotation.

As schematized in Figure 1, a flotation column is a device where a slurry and air bubbles flow countercurrently. A high length to diameter ratio increases the probability of collision between the mineral particles contained in the slurry and the rising air bubbles generated at the sparger located near the bottom of the column. The volume of the column occupied by the mixture slurry-air bubbles is called the collection zone because it is here where the particles containing the valuable components are collected by attachment to the air bubbles.
The air bubbles carrying mineral particles rise and form a layer of froth as they leave the collection zone. Water is introduced from the top of the column to stabilize this froth and to build a froth bed of packed bubbles about a meter deep, which is called the cleaning zone. In addition to stabilizing the froth, this downward flow of water washes from the froth particles of gangue material entrained in the water layer surrounding the air bubbles; this results in an increase of concentrate grade. The tailings flowrate defines the residence time within the column.

![Diagram of flotation column](image)

**Fig. 1.** Schematication of a flotation column

The froth depth (slurry-froth interface level) needs to be maintained within a certain range to have efficient operation of the column. Small froth depths result in a reduction of concentrate grades as particle entrainment is only partially eliminated, while excessive froth depths reduce recoveries as a consequence of the small volume of the collection zone, which means shorter residence times.

**TECHNICAL BACKGROUND**

The froth depth or interface level is a variable that currently is inadequately sensed. The common method of detecting this variable is based on the measurement of the static pressure below the interface, a method that reports the level with errors sometimes in the
range of meters. Electrical conductivity has been used to detect the pulp level on conventional flotation cells. Previous work in a laboratory column has shown that a major difference in conductivity exists between the froth and the slurry as a consequence of their different water contents. Yianatos et al. (1985) developed equations that allow the determination of gas holdups in air-water systems from conductivity measurements in each phase. Based on these equations, the following relationship between the ratio slurry to froth conductivity \( R \) and gas holdups can be obtained:

\[
R = \frac{2.315 \epsilon_f(1-\epsilon_S)}{[(1+0.55\epsilon_S)(1-\epsilon_f)]}
\]

where \( \epsilon \) is the gas holdup and the subscripts S and F refer to the slurry and froth, respectively.

The variation of \( R \) for different holdups in both phases is illustrated in Figure 2. Typical ranges for these variables during normal operation of flotation columns are also indicated in the figure, which shows that a conductivities ratio sufficiently large for level detection purposes can always be expected. However, the presence of a third phase in the form of solid particles may modify this picture. The effect of mineral particles in the conductivities of slurry and froth has not been described and the information available is so scarce that its prediction is hazardous.

![Figure 2. Slurry to froth conductivity ratios for different gas holdups in an air-water system](image-url)
EXPERIMENTAL SET-UP

The laboratory column used in this work is made of transparent Plexiglas tubing (5.04 cm internal diameter). Five sections of 2.1 m each give a total height of about 10.5 m. A porous stainless steel sparger is installed at the bottom of the column in a removable section that also allows for tailings exit. The pressure of the air supplied to the sparger is controlled using a gas regulator and a suitable flowmeter. Feed is introduced at about 2 m from the top. The top end of the column is surrounded by a square box that acts as an overflow reservoir to receive and guide the concentrate stream. Washwater is supplied using a perforated 1/4" copper tube ring installed at about 1 cm below the froth lip. Peristaltic pumps (Masterflex Cole Parmer, model 7520), are used to supply and control the flows of feed, tailings, and wash water stored in plastic containers.

Fig. 3. Arrangement of electrode pairs in the top section of the flotation column
Fourteen electrode pairs are installed in the top portion of the column to obtain conductance profiles and conductance readings that allow the calculation of the interface position within the column. In the case of conductance profiles, 11 electrode pairs (stainless steel plate, 1x2 cm), were glued to the column walls following vertical lines 180 degrees apart, and with a separation of 10 cm starting at 50 cm from the top of the column (see Figure 3). For level estimation, 3 electrode pairs are installed in the same way but on vertical lines displaced 90 degrees from those of the profile electrodes. A pair of long parallel electrodes (stainless steel plate, 1x100 cm), whose response is proportional to the position of the level, is installed at 50 cm from the top of the column. Two pairs of small electrodes for measuring froth and slurry conductances are installed at 40 and 110 cm from the top of the column, respectively.

The data are collected using a computerized data acquisition system formed by a microcomputer (Mackintosh, Apple compatible, 64 K of memory), a conductivity meter (Tacussel, model CD 810), an interface board which includes a clock and an 8-bit A/D converter, and a manual 10-channel selector. Figure 4 shows the arrangement used in this application.

Fig. 4. Schematization of the data acquisition system
EXPERIMENTAL RESULTS

The column was installed at the Strathcona mill, Falconbridge Ltd., in Sudbury. Several runs were completed using different solids contents and residence times. The interface level was controlled varying only one of two variables: feed or washwater flowrate. Air flowrate was maintained constant in all these experiments at 3220 cc/min (2 cm/s). After the column was operated at the selected conditions for 2 residence times, the data collection

![Graph showing conductance profile around the interface](image)

**Fig. 5.** Conductance profile around the interface (low level).

was initiated. For 15 minutes, each one of 10 electrode pairs linked to the channel selector were connected to the conductivity meter for 6 seconds (time between readings). The real position of the level, determined by visual inspection, was also recorded every minute.

The positions of the electrode pairs selected to collect a conductance profile were those at 40, 60, 80, 90, 100, 110, 120, 140, and 160 cm from the top. Typical results are presented in Figures 5 and 6, which were selected as to show profiles for low and high interface levels, respectively. The results show a large difference between the conductances of froth
and slurry, which is a necessary condition for the successful use of conductivity in the detection of interface level. Another important result is that the conductance varies within one phase as the position of the measuring point moves away from the interface (i.e., there is a profile). As will be discussed, this observation has an impact on the detection of the interface level.

![Graph showing conductance profile](image)

Fig. 6. Conductance profile around the interface (high level).

The detection of the interface level within the column is based on the conductance through long parallel electrodes; the conductance is proportional to the fraction of their length immersed in froth and slurry. The position of the level can be estimated from this conductance if the conductivities of froth and slurry between the level electrodes are known. As the conductivities of froth and slurry depend on the properties of the materials
being processed and on the values of operational variables such as air and washwater flowrates or reagent dosage, an accurate estimation of the interface level requires the measurement of froth and slurry conductivities at about the same time that the conductance through the level electrodes is determined. Initially unaware of the extent of variation within a phase, electrodes were placed 10 cm above and below the level electrodes to obtain a value representative of the froth and slurry conductivities, respectively. As differences in cell constant are smaller than 2% for all electrode pairs, the conductance profiles show that the selected locations away from the level electrodes were not the most appropriate. The conductance variation as the measuring point moves away from the interface implies that the values determined do not represent appropriate average values for the froth and slurry within the level electrodes. This is illustrated in Figure 7, which

![Graph showing calculated froth depth vs. froth depth](image)

**Fig. 7.** Level prediction using average values of slurry and froth conductivities around the interface and values determined away from the interface.

compares the position of the interface calculated from conductivity measurements with the value obtained from visual inspection. However, when an average value calculated from the profiles is used, the correlation between two values improves considerably, as also shown in Figure 7.
CONCLUSIONS

The results obtained in this work clearly establish that electrical conductivity can be effectively used to detect the position of the slurry-froth interface in flotation columns. A large difference in the conductivities of froth and slurry, which is essential for a successful estimation of interface level using conductivity, was measured at various conditions in a laboratory flotation column processing several streams. The difference was similar for a wide range of bubble loadings and particle compositions, which suggests that the most important parameter defining the conductivities of froth and slurry within a flotation column is the gas holdup and not the characteristics of the stream processed. Another crucial result is the detection of conductivity profiles within each phase, which must be considered for the accurate estimation of the froth depth.

REFERENCE

2.5. A CALIBRATION PROBE FOR LEVEL DETECTION SYSTEMS IN FLOTATION COLUMNS

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ABSTRACT

Froth depth or interface level is an important control variable in column flotation which can influence both grade and recovery. A common method of locating the level is based on sensing the hydrostatic pressure at one or more points along the column. The level indicated can be in considerable error, largely because of variations in froth and collection zone bulk densities. A portable conductivity probe has been developed to detect the position of the interface from conductance measurements across the interface. The location of the interface corresponds to the position where a large change in conductance is observed, which is primarily a consequence of the difference in gas holdup between the collection and froth zones. This probe detects the position of the interface to an accuracy of ± 5 cm. The probe can be used while the column is in operation, which allows for online re-calibration of the level detection system. Results obtained in its application in a pilot-scale column (0.91 min diameter) are described.

INTRODUCTION

A flotation column is a device where a slurry and air bubbles flow countercurrently (Figure 1). The region below the interface is called the collection zone because it defines the region where the particles containing the target components are collected by attachment to the air bubbles. The air bubbles carrying mineral particles rise and form a layer of froth as they leave the collection zone. A deep bed of packed bubbles builds above the interface which is called the cleaning or froth zone. The interface between the collection and froth zones is usually quite distinct. The froth zone typically has wash water added near the top (Figure 1) which helps stabilize the deep froth and wash back to the collection zone particles of gangue material entrained in the water layer surrounding the air bubbles in the froth, which results in an increase of concentrate grade.

One of the operational variables that needs to be controlled for stable operation of the column is the froth depth, or 'level' as it is commonly called. Small froth depths may result in a reduction of concentrate grade if particle entrainment is only partially eliminated, while excessive froth depths reduce recoveries as a consequence of the smaller volume of the collection zone. The froth depth is a variable that currently is inadequately sensed. The common method of detecting this variable is based on the measurement of the hydrostatic pressure below the interface, a method which requires knowledge of the bubbling and froth
zone bulk densities. These densities cannot be obtained from theoretical considerations and must be obtained by calibrations when one or two pressure transducers are used, or continuously determined using three pressure transducers conveniently located along the column (Dirus, 1988). Since these densities change in an unpredictable way upon variation in the operating conditions or stream characteristics, the estimated level experiences significant error (Moys and Finch, 1988). On the other hand, electrical conductivity offers a means of level detection whose accuracy is virtually unaffected by variation in operating conditions or stream characteristics. This paper describes the use of a conductivity probe designed to calibrate level detection systems in flotation columns.

![Schematic diagram of a flotation column.](image_url)

**Fig.1.** Schematic diagram of a flotation column.

**LEVEL DETECTION USING CONDUCTIVITY**

The conductance between two electrodes immersed in a liquid is proportional to the cross-sectional area of the electrodes and inversely proportional to the distance between them. When air bubbles are present, the conductance is reduced for two reasons: i) a smaller cross-sectional area occupied by the liquid, which is the conducting medium; and ii), a larger effective distance between the electrodes because the ions must follow a tortuous path. Therefore, the higher the gas holdup (volume occupied by gas), the smaller the conductance. Since, in flotation columns the collection and froth zones present different gas holdups, the level position should be detectable from the change in conductance. This assumes that the presence of solid particles does not affect adversely the conductance in each zone.
Figure 2, based on the air-water experiments of Yianatos et al. (1985), illustrates the range in the values that can be expected for the ratio of collection to froth zone conductance. This ratio is larger than one because the conductance in the collection zone is always larger than that in the froth zone, as a consequence of its smaller gas holdup. For level detection purposes, given that the noise in the signal is about 10%, a conductance ratio over 2 is considered sufficiently large for an accurate estimation of the transition between the collection and froth zones; as shown in Figure 2, this value is expected to be reached or exceeded in flotation columns.

![Froth Gas Holdup Diagram](image)

Fig.2. Collection to froth zone conductance ratio for different gas holdups in an air-water system (based on Yianatos et al., 1985).

Gomez et al. (1988) found that the presence of solids does not significantly change the value of the conductance ratio. Similar values of the conductance ratio obtained for a wide range of bubble loadings and particle compositions suggest that the most important parameter defining the conductance ratio remains the relative gas holdup and not the characteristics of the solids being processed. Small changes in solution chemistry did alter the value of the conductance ratio but the transition at the interface always remained clear. (It is possible, if the conductivity of the feed slurry water decreased substantially while that of the wash water increased, that the conductance ratio could approach one; this is considered an unlikely event and certainly not one that has been encountered). A recent report on the use of electrical conductivity to discriminate the position of the interface in flotation columns processing coal is consistent with these observations (Nicol et al., 1988).
DESCRIPTION AND OPERATION OF THE CONDUCTIVITY PROBE

A portable probe was designed to measure the conductance at different heights in a flotation column. The probe is made of aluminum and PVC tubes which can be assembled in a few minutes (Figure 3a). The central component is a supporting aluminum cube with threaded holes in which are inserted a vertical aluminum tube, two lateral PVC guiding tubes, and a PVC tube used to support the electrode. The electrode (Hanna Instruments HI 76301) has a PVC body and four stainless steel sensor rings. Wiring between the electrodes and a portable conductivity meter (Hanna Instruments HI 8633) is completed through the interior of the tubes. The vertical aluminum tube is graduated to indicate the depth of the electrode relative to the top of the column. Figure 3b is a top view illustrating how the arrangement of PVC tubes keeps the electrode close to the centre of the column, which minimizes signal distortions produced by the column wall that is electrically grounded.

Fig. 3a. Schematic diagram of the portable conductivity probe.
The probe is operated from the top of the column. To locate the interface, the operator slides the probe up or down looking for the position where a large change in conductance occurs (experience indicates that moving up from the collection to the froth zone is preferable because the change is detected more accurately). This change reflects the different gas holdups in the two zones. The accuracy with which the level is detected is better than ± 5 cm, as judged from laboratory columns where the interface can be seen (a typical conductance profile collected in a laboratory column is shown in Figure 4). The information collected can be used, for example, in two ways: i) to construct a calibration curve for selecting nominal levels used later as set-points in the control system; or ii), in cases where level is calculated from pressure, to recalculate froth and collection zone bulk densities which better represent those existing in the column during operation.

![Figure 3b](image)

**Fig.3b. Schematic view from the top of the column with the conductivity probe in its working position.**

**EXAMPLES OF APPLICATION**

The level determined using the conductance probe was compared with that indicated by the system installed on the pilot flotation column (0.91 m dia.) at the Strathcona concentrator (Falconbridge Ltd.). The system uses a single pressure sensor and corrects for slurry density using feed and tailings density readings. The comparison (Figure 5) reveals that the indicated level is underestimated by up to 1 metre. The system was re-calibrated using the
conductance level data and level estimation was improved. Figure 6 shows, however, that later processing of a different stream once more encountered errors meaning the system needed recalibration.

Fig.4. Typical conductance profile collected in a transparent laboratory column. (Dotted line indicates visually located interface).

Experiments were also conducted to assess the effect of different operating conditions using the level control strategy in use on the column. Part of the strategy is to maintain the level set-point when the air flowrate is changed by manipulating the wash water flowrate. The experiment asked the control system to keep a constant froth depth of 105 cm as the flowrate was increased from 10 up to 26 scfn (0.73 to 1.89 cm/s). The level indicated in the control room was constant at 105 cm during the whole sequence. The results obtained with the conductivity probe, however, reveal a different story (Figure 7). The level control strategy is effective only at low air flowrates; as the air flowrate increased, above about 1.1 cm/s the level actually moved up and eventually, at the largest air flowrate used, it reached
the top of the column. The strategy is basically correct—as air flowrate increases more water is transferred from the collection zone to the froth and level decreases calling for an increase in wash water—but the strategy demands excessive wash water at the higher gas rates.

It is beyond the scope of this communication to consider in detail the origin of the errors evident in Figures 5-7. It lies in part in the zone bulk density values and their unknown change upon altering stream characteristics and operating conditions. The conductance probe was instrumental in revealing the limitations in the installed level detection and control systems.

Fig. 5. Comparison of level estimated using conductivity probe and that reported by the installed system.
CONCLUSIONS

A portable conductance probe has been developed which determines level to within ± 5 cm. The probe is robust, simple to use and can be assembled and disassembled in a few minutes. It was used to test the level detection and control system installed on the pilot flotation column in the Strathcona concentrator (Falconbridge Ltd.); substantial errors were revealed. Using the conductance probe the system was recalibrated on-line.

Fig.6. Errors detected in the level reported by the installed system when the column is operated with a different stream to that used during its calibration.
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Fig. 7. Errors detected in the reported level as the installed level control system reacts to increasing air flowrates to maintain a constant level (105 cm).

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2.6. A LEVEL DETECTION PROBE FOR INDUSTRIAL FLOTATION COLUMNS

C.O. Gomez, A. Uribe-Salas, J.A. Finch, B.J. Huls


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ABSTRACT

The most common method for detecting froth depth (level of the interface between the bubbling and froth zones) in a flotation column is by measuring the hydrostatic pressure below the interface. The accuracy of this method is affected by variations in the operational variables and in the characteristics of the stream being processed. The error can be substantial. A new approach has been developed based on the different conductances in the bubbling and froth zones which are a consequence of their different gas holdups. The technique has been successfully used in laboratory and pilot-scale (0.91 m diameter) columns giving accuracies better than \( \pm 5 \) cm. Based on this experience, an industrial probe has been designed and built for use in production columns. This paper includes a description of the probe and results obtained during plant testing.

KEYWORDS

Mineral Processing; flotation; column flotation; flotation column control; level detection; level sensor; electrical conductivity.

LEVEL DETECTION IN FLOTATION COLUMNS: IMPORTANCE AND DIFFICULTIES

An interface, usually quite sharply defined, separates a flotation column into two zones (Fig. 1). The volume of the column below the interface is called the bubbling or collection zone. Here rising air bubbles collect and carry mineral particles across the interface to form a froth zone. Water introduced from the top of the column stabilizes the froth and at the same time washes back to the collection zone particles of gangue material entrained in the water layer surrounding the air bubbles. The froth layer is also called the cleaning zone. Froth depth or interface level is an important control variable in column flotation which can influence both grade and recovery. Small froth depths may result in a reduction of grade if particle entrainment is not completely removed. Excessive froth depths reduce
recoveries as a consequence of the smaller volume of the collection zone which means a shorter particle residence time.

Interface level is a variable that currently is inadequately sensed. The most common method for measuring level is from measurements of the hydrostatic pressure below the interface. As discussed elsewhere (Moys and Finch, 1988; Gomez and co-workers, 1989), this approach is subject to substantial error because it requires knowledge of the bulk densities in the bubbling and froth zones. These densities are influenced in an unpredictable way by variations in the operating conditions and the stream characteristics. A new approach is being developed based on the different gas holdup existing in the bubbling and froth zones (Yianatos et al., 1985). As electrical conductivity within a column is a variable primarily depending on the gas holdup, given that a sharp gas holdup transition exists around the interface, it is expected that this transition will be readily detected by conductance measurements. This approach has been successfully used in laboratory flotation columns (Gomez and co-workers, 1988). Interface level has been accurately measured in pilot-scale columns from conductance profiles manually collected using a portable probe, and the results used to calibrate the level detection system installed in the column (Gomez and co-workers, 1989). The extension of this approach to industrial flotation columns requires the development of a probe suitable for collecting conductance profiles along the column.

Fig. 1. Schematic representation of a flotation column.
MEASUREMENTS OF LOCAL CONDUCTANCE WITHIN A COLUMN

The major difficulty in the development of a conductivity probe for level detection is the selection of appropriate electrodes to measure local, representative values of the conductance. In the case of the laboratory column, the conductance profiles were collected using electrodes permanently attached on opposite sides of a 5 cm laboratory column (Gomez and co-workers, 1988). In the case of the pilot-scale column (0.91 m), the conductance profiles were collected using a commercial electrode installed in a portable structure that was slid up and down the column (Gomez and co-workers, 1989). Neither alternative is suitable for on-line use in production columns.

A literature review gave no indication of possible electrode configurations or geometries appropriate for this task. Work in progress, that will be reported elsewhere, has shown that the cell constant for electrodes where the current flow is not restricted by walls does not correspond to the ratio between the area of the electrodes and the distance between them. This means that local conductances can be measured with electrodes that need not necessarily be facing each other. This point can be illustrated using the results obtained during the testing of the electrode configurations and geometries described in Fig. 2.

![Diagram of electrode configurations](image)

Fig. 2. Different electrode configurations tested.

These different electrodes were installed one at a time in a portable structure that allows the collection of conductance profiles as it is moved up or down the column. Conductance profiles for the electrode configurations included in Fig. 2 were collected in a pilot-scale column installed at the Strathcona mill (Falconbridge Ltd.) which was processing a stream from the copper circuit. As expected, the results showed marked differences in the conductances measured, but when the profiles were displayed using a normalized
conductance, all the electrodes showed similar abilities to detect the conductance difference across the interface (Fig. 3). The conductance was normalized using the value obtained at 3 m from the top. The geometry offering easiest construction and maintenance is the ring (A in Fig. 3); the ring gave the second best sensitivity after the facing rectangular plates (D in Fig. 3).

Fig.3. Level detection using different electrode configurations. a) raw data; b) normalized, conductance relative to that 3 m from top of column.

LEVEL DETECTION PROBE AND ASSOCIATED DATA COLLECTION SYSTEM

The design of the stationary probe was based on a basic module formed by a non-conducting PVC cylinder shaped to support a stainless steel ring that acts as an electrode. This module was designed so that a probe can be assembled by mounting a series of them around a 2 inch stainless steel pipe; the number of modules used defines the active length of the probe. Every stainless steel ring is grounded to the internal stainless steel pipe, while those in between the grounded rings are connected to a relay board driven by a computer which activates them sequentially. The conductance is measured between the activated ring and those grounded on either side. Figure 4 shows details of the probe which includes 25 rings (12 active and 13 grounded); this allows detection of local conductances along 1.1 m
of the column at 12 positions at 10 cm intervals (separation of the active rings). Figure 4 also gives specifications of the two components of the basic module and illustrates the way they are assembled. The whole structure is installed at the top close to the centre of the column and parallel to its wall. The range of froth depths to be monitored is defined by the location of the probe relative to the top of the column. The probe was installed in this case to measure conductances at distances between 43 and 153 cm from the top.

Fig. 4. Schematic representation of level detection probe and specifications of module used in its construction.

Software was developed to collect and display conductance profiles automatically as well as to store data for further analysis. As indicated in Fig. 5, the data acquisition system is formed by a microcomputer (Comterm PC, IBM compatible, 256 K memory), a conductivity meter (Tacussel, model CD 810), a 24-channel relay board (Metrabyte, model ERB-24), an I/O interface to drive the relay board (Metrabyte, model PIO 12), and the A/D converter
of a Data Translation interface (model DT2801). User-defined variables are the time between electrode readings, normally selected as the response time of the conductivity meter (3 seconds), and the time between profile collection (normally 1 minute). To collect a profile, the computer sequentially connects active electrodes to the conductivity meter starting from the top of the column. The signal emitted by the conductivity meter, which is proportional to the conductance, is continuously fed to the A/D converter; its output after the time between readings has elapsed is stored in the computer memory and plotted.

![Diagram of data acquisition system]

Fig. 5. Schematic representation of data acquisition system.

RESULTS OF PROBE TESTING EXPERIMENTS

An experimental program was designed to test the performance of the probe in the pilot column. Three types of experiments were conducted: i) collection of conductance values for all rings in still water to define cell constants of the individual rings; ii) collection of
conductance values in single rings to define signal noise; and iii) collection of profiles for extended periods of time to test detection reliability and level calculation algorithms.

i) The collection of conductance values to define cell constants for each of the 12 active rings was completed using a high data collection frequency (1 reading every second) with the column filled with water. The results showed that the conductance measurements in still water were remarkably stable and very similar for the different rings, as illustrated in Fig. 6 by rings 1 and 4; the response of the other rings were all in between these two. Average conductance values were calculated for the different rings to normalize their responses; this is important because the readings of these different electrodes will be used to form a conductance profile along the column.

![Conductance Profile](image)

**Fig. 6.** Conductance variations in still water.

ii) The collection of conductance values in individual rings was also completed using a high data collection frequency (1 reading every second). Figure 7 includes typical results obtained for rings 2 and 11 with the column processing a stream from the copper circuit at normal operating conditions. This situation corresponds to ring 2 in the froth zone and ring 11 in the bubbling zone. As expected, the results showed that the noise of the signal is larger than that obtained in still water, which obviously is a consequence of the presence of rising air bubbles entering and leaving the confines of the volume acting as the cell for each electrode.
Fig. 7. Conductance variations at normal operating conditions.

Fig. 8. Typical conductance profile around the interface.

iii) Conductance profiles were collected once every minute for several hours, with five seconds between electrode readings; Fig. 8 shows a typical profile; conductance for each ring was normalized using the respective average conductance determined in still water. The performance of the probe was very satisfactory, the collected profiles showed a sharp conductance change across the interface.
A simple algorithm to estimate the froth depth from a conductance profile was used. By calculating the difference between the conductances measured with two consecutive active electrodes, the level was clearly detected by those consecutive electrodes giving the largest difference. The level was estimated to be at the half-distance between them. Since the separation of the active electrodes is 10 cm, the algorithm, therefore, reports the level with a maximum error of ± 5 cm (work is in progress to improve the accuracy of the detection using more sophisticated algorithms). Figure 9 illustrates the application of the conductance difference algorithm to the profile in Fig. 8. Figure 10 shows the position of the level calculated using this algorithm over several hours of operation.

Fig. 9. Illustration of conductance difference algorithm.

Fig. 10. Level variations detected by conductivity probe.
OBSERVATIONS AND CONCLUSIONS

A stationary conductivity probe for detecting interface level in industrial flotation columns has been developed. The design of the probe is based on a series of ring electrodes which allow the collection of a conductance profile along the column. The detection of the interface level is made possible by the difference in conductance between the bubbling and froth zones because of their different gas holdups. The selection of rings gave advantages for construction and maintenance purposes. The results obtained in a series of experiments completed to test the probe demonstrated that its performance was very satisfactory. The collected conductance profiles revealed a sharp transition between the bubbling and froth zones. The low noise of the conductance measurements in individual rings guarantees that the sharp conductance variation existing across the interface can be accurately detected. The probe can be installed or removed in a few minutes, which makes maintenance easy. The materials selected to build the different components showed no signs of corrosion after three months of use.

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2.7. LEVEL DETECTION IN FLotation COLUMNS AT FALCONBRIDGE

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ABSTRACT

Accurate level detection is required to make full use of level in optimizing control strategies for flotation columns. Techniques based on pressure and electrical conductivity are compared. Pressure and conductivity profiles give good agreement on level. Level calculated from single and multiple pressure values became more accurate using, in order, one, two and three sensor locations. The level estimate remained in error, however, because, unlike the collection zone, froth zone bulk density varies with height. A conductivity probe continually updating the conductance profile is a practical method of obtaining an accurate level.

INTRODUCTION

Flotation columns cannot be operated manually. The minimum requirement for stable operation is control of the pulp/froth interface position (i.e. froth depth, or "level").

Measurement of level in flotation columns poses some novel problems compared with conventional cells. Froth depths are much greater, commonly 1 m or more and, because of the column geometry, level is more sensitive to feed flow fluctuations.

The most frequent method of level detection is to infer from pressure measurements. Pressure has been determined in a variety of ways, including manometers, Metritape and electronic transducers. The simplest arrangement is a single sensor below the interface. While susceptible to error, if the requirement is simply stable operation this technique is often adequate. However, if level control is part of an optimizing strategy, accurate measurement is required, and this has spurred the development of improved level detection systems. These improvements have included multiple pressure sensors, and alternative techniques, such as the use of electrical conductivity. In this paper, the use of pressure (single and multiple sensors) and conductivity is compared for level detection.
LEVEL DETECTION SYSTEMS

The experimental work was conducted on a 0.91 m diameter pilot column and on one of two 2.13 m diameter plant columns. Each column measures approximately 13.5 m in height and is located at the Strathcona Mill in Falconbridge Limited near Sudbury, Ontario.

In the current process control strategy, level is maintained by manipulating wash water addition. Other controlled parameters are the feed flow rate which varies depending on the level in the head tank, the bias i.e. the tailings slurry flow setpoint in cascade from the feed flow, and the air flow rate and pressure on the plant columns.

The pilot column is equipped with two Taylor (model 532TB04112A) pressure transmitters. Both are located on the column wall below the froth-pulp interface at a distance of 3.50 m and 11.96 m from the lip. Calibration ranges 0 to 6 kPa and 0 to 207 kPa respectively. For the plant columns Rosemount pressure transmitters were selected (model number 1151 LT4EA6A23DC6). Three of these were mounted in each column and protrude about 6 mm from the wall to prevent solids build up on its surface. The respective distances from the lip of the column are 0.53 m, 2.21 m, and 9.30 m, and are calibrated from 0 to 10, 37 and 125 kPa respectively. The top pressure transmitter is generally located in the froth zone.

**Pilot column**

Level measurements were made using the hydrostatic pressure from a single pressure transmitter below the pulp interface (Figure 1a). The froth depth can be calculated from the pressure if appropriate values for the bulk density in the collection and froth zones are known. If \( h_b \) is the distance of the pressure transmitter to the lip and \( h_f \) is the froth depth, the pressure measured can be expressed as

\[
P_b = h_f + h_c g(h_b-h_f)
\]

where \( h_f \) and \( h_c \) are the densities of the froth and collection zones respectively. Equation (1) can be rearranged to determine \( h_f \):

\[
h_f = \frac{h_b g - P_b}{g(h_c - h_f)}
\]

A first estimate of the density in the collection zone can be obtained by measuring the column tailings density. The bulk density, however, is a function of several poorly controlled variables including gas holdup, the density of the slurry phase, and the density of the bubble-particle aggregate (Gomez et al., 1989 a).

When placing a second pressure transmitter in the collection zone (Figure 1b) the bulk density of the collection zone, in effect, is measured by
\[
\tau_c = \frac{P_c - P_B}{g(h_c - h_B)}
\]  \hspace{1cm} (3)

Inserted into equation (2), the level calculation becomes

\[
h_f = \frac{(P_c - P_B)h_c - P_c(h_c - h_B)}{(P_c - P_B) - f_f g(h_c - h_B)}
\]  \hspace{1cm} (4)

The results of the two level calculations were compared with measurements obtained by conductivity. Two conductivity probes were used. One measures conductivity at a single point. To find the pulp interface this probe needs to be adjusted up and down manually to obtain a conductivity profile in the froth and pulp. The probe was designed for on-line calibration of the pressure-based level detection techniques (Gomez et al., 1989a).

The second probe is stationary and measures conductivity at 25 points, at 10 cm intervals, automatically and sequentially. The probe is interfaced with a computer and updates the profile each minute. The level is inferred from the largest relative difference in conductivity between two consecutive points (Gomez et al., 1989b).

![Fig.1. Techniques for measuring position of interface (i.e. froth depth or "level") using a single or multiple pressure sensors.](image)

Production column

In the pilot column, no pressure reading is made in the froth zone, but here also the bulk density is affected by factors such as gas holdup, slurry phase density and bubble-particle aggregate density. For this reason the plant columns were fitted with a third pressure transmitter located in the froth zone (Figure 1c) to monitor the bulk density of the froth zone. Froth depth is now given by

\[
h_f = \frac{h_c(P_c - P_b) - P_c(h_c - h_b)}{h_c(P_c - P_b) - P_c(h_c - h_b)}
\]

(5)

This equation provides a more accurate level measurement because, in effect, the froth zone and collection zone bulk densities are being measured. The top pressure transmitter should be located as near the interface as possible to provide the best information on average froth bulk density. If for any reason the level would rise above this transmitter, equation (5) is no longer valid.

The estimate of level using equation (5) was compared with the measurement by conductivity, using the single point probe.

RESULTS AND DISCUSSION

Experiments were completed in the pilot-scale and in one of the production columns to test the performance of the different techniques available to measure the interface level. To compare the performance of the various level detection systems, experiments were designed to collect simultaneously: a pressure profile along the column, manual and automated conductance profiles around the interface, and the pressure transducer readings. The installation of the stationary probe in the production column was being completed at the time of this work and, therefore, automated conductance profiles for the production column could not be collected.

The accuracy with which the interface level is detected from static pressure readings depends on the magnitude of bulk density variations in the froth and collection zones. To define these variations, a device was developed to collect a complete pressure profile along the column by installing a pressure transducer (Druck, model PDCR86X, range 0-10 psig) in a submersible stainless steel cylinder. This cylinder can be reeled down from the top of the column to any depth using a 6-wire shielded cable through which the transducer is powered and its resulting signal is transmitted to a microcomputer. At any position, the pressure is considered as the average of 20 values collected one every second. The collection of the data requires about 12 minutes of stable operation (particularly a stable level).
Pilot column

A pressure profile was collected by measuring the static pressure at 30 positions along 12 m of the column. For the first 2 m the pressure was measured every 0.2 m, while for the remaining 10 m the separation was 0.5 m. The results are presented in Figure 2, which also includes the readings of the two pressure transducers permanently installed in the column as part of the level detection system. There is excellent agreement between the response of these transducers and the one used to collect the profile. These results indicate a pronounced bulk density variation in the froth zone and, rather surprisingly, a constant bulk density in the collection zone.

Fig. 2. Pressure profile collected along pilot column.

Figure 3 shows the data collected in the first 2 m from the top of the column in an expanded scale. The point where the line breaks indicates where the interface is located - 1.25 m from the top of the column in this case. The calculation of the level from the transducer readings, however, results in appreciable error because, as no pressure is measured in the froth zone, its bulk density cannot be estimated and is assumed as zero. The calculation under these circumstances can be graphically represented in Figure 2 by extending the slope of the profile in the collection zone until the y-axis is crossed (0.9).
Fig. 3. Pressure profile at top of the pilot column.

Fig. 4. Conductance profile collected with single-point probe in the pilot column.
The results obtained using conductivity probes are presented in Figures 4 and 5. Figure 4 includes the conductance profile collected using the single-point probe. These results indicated that the level is about 1.2 m from the top of the column. The profile presented in Figure 5 is from the stationary probe. This profile also showed that the level is located at about 1.2 m from the top of the column.

Production column

The same distance intervals used in the pilot column to collect the pressure profile were used in the production column. The results (Figure 6) showed similar trends as those obtained in the pilot column: a fairly constant bulk density in the collection zone and a substantial variation in the froth zone. As before, the readings obtained with the transducers permanently installed in the column are also included, but small differences were detected this time in the response of the top two transducers. Figure 7 illustrates this point more clearly as the pressures measured in the first 2 m from the top of the column are presented in an expanded scale. The position of the interface, defined by the breaking point, is about 2.1 m. The conductance profile collected with the single-point probe also indicates that the level is at 2.1 m (Figure 8).

![Conductance profile](image)

Fig. 5. Conductance profile collected with stationary probe in the pilot column.

The calculation of the level from the output of the three pressure transducers reports the level with error, which is now primarily a consequence of the inadequate value for the bulk density in the froth zone calculated from the reading of the top pressure transducer. The reason is that this transducer is too far away from the interface. The calculation can be
Fig. 6. Pressure profile collected along production column.

Fig. 7. Pressure profile collected at top of the production column.
Fig. 8. Conductance profile collected with single-point probe in the production column.

graphically illustrated in Figure 7 by drawing two lines, one going through the origin and the point representing the two transducers located in the collection zone. These lines represent the contribution of the froth and collection zones, respectively, to the pressure measured by the middle transducer B. The calculated level is defined by the point where both lines intersect (1.84 m in this case). Table 1 summarizes the levels reported by the different techniques used in this work.

<table>
<thead>
<tr>
<th>Level detection technique</th>
<th>Pilot column</th>
<th>Production column</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pressure profile</td>
<td>1.25</td>
<td>2.12</td>
</tr>
<tr>
<td>Conductance profile (manual)</td>
<td>1.20</td>
<td>2.10</td>
</tr>
<tr>
<td>Conductance profile (autom.)</td>
<td>1.20</td>
<td>--</td>
</tr>
<tr>
<td>Two pressure sensors</td>
<td>0.90</td>
<td>1.70</td>
</tr>
<tr>
<td>Three pressure sensors</td>
<td>--</td>
<td>1.84</td>
</tr>
</tbody>
</table>

Table 1. Levels reported by the different techniques used in this work.
CONCLUSIONS

The results indicate that pressure and conductance profiles report the interface level in close agreement. Pressure profiles collected along the pilot and the production columns showed that the bulk density in the collection zone is fairly constant and, therefore, its value can be accurately measured with two pressure sensors located within the collection zone. On the other hand, the bulk density in the froth zone showed substantial variations which make it difficult to obtain a representative average value using a single pressure sensor. The consequence of this is that the calculation of the interface level from the reading of two or three pressure sensors is subject to error. As expected, the use of three pressure sensors gives more accurate results than when only two (or one) are used where the bulk density of the froth is considered zero. The magnitude of the error in the case of three pressure sensors increases as the distance of the top ('froth') sensor from the interface increases.

Detection of level from the profile of conductivity is shown to be a practical and accurate method. It is less sensitive to uncontrolled variables in comparison with the pressure method based on 1, 2 or 3 sensors.

REFERENCES


2.8. GAS RATE AND FROTH DEPTH EFFECTS ON PERFORMANCE OF A CU-NI SEPARATION FLOTATION COLUMN

B.J. Huls, C.D. Lachance, G.S. Dobby


ABSTRACT

A 0.9 m diameter pilot flotation column has been in operation at Falconbridge's Strathcona concentrator for the purpose of Cu-Ni separation. The results from this column have been used to design a full scale column circuit to treat all of the Cu-Ni separation feed. To design a control strategy for the full scale column extensive tests have been run with the pilot column, which is instrumented to measure all process flows, froth depth, slurry density and gas holdup. This paper examines the effect of gas rate, froth depth, feed slurry density and water spray position upon metallurgical performance and discusses implications relating to control strategy.

KEYWORDS

Column flotation, Cu-Ni separation, froth depth, gas rate, slurry density, pilot column.

BACKGROUND

Falconbridge Ltd. owns and operates the Strathcona Concentrator about 45 km northwest of Sudbury, Ontario. Five mines currently feed the mill at a rate of about 9000 tonnes per day of ore containing about 1.5 % Ni, and 1.4 % Cu. The flowsheet described in an earlier paper (Scales et al. (1986)) has been largely unaltered since start up. The primary rougher concentrate is pumped to a copper-nickel separation circuit, from which the nickel concentrate is combined with the secondary rougher concentrate. The scavenger concentrate is subjected to magnetic separation to allow individual flotation of the non-magnetic and magnetic fractions for optimum rejection of pyrrhotite, a low nickel bearing iron sulphide.

In 1986, column flotation was proposed as an alternative to the conventional method of separating copper from nickel minerals, employing 1.4 m³ Denver flotation cells in a rougher-cleaner application. Copper-nickel separation is effected at high pH and with the addition of NaCN. Historically, it proved difficult to control the nickel grade in the copper concentrate, which typically assayed 0.8 to 0.9 % nickel and 27 to 29 % copper. It is desired to maintain the nickel content in the copper concentrate below 0.5 %. Chemical screen analyses of concentrates obtained from the conventional circuit indicated that in the fine fraction the nickel assay was excessive compared to the coarser fractions. In comparison,
a concentrate obtained from the column appeared uniform in nickel assay throughout the size fractions as illustrated in Table 1.

<table>
<thead>
<tr>
<th>SIZE FRACTION (μm)</th>
<th>% Ni COLUMN</th>
<th>% Ni CONVENTIONAL</th>
</tr>
</thead>
<tbody>
<tr>
<td>+ 106</td>
<td>0.28</td>
<td>0.23</td>
</tr>
<tr>
<td>74 - 106</td>
<td>0.32</td>
<td>0.19</td>
</tr>
<tr>
<td>54 - 74</td>
<td>0.41</td>
<td>0.24</td>
</tr>
<tr>
<td>37 - 54</td>
<td>0.35</td>
<td>0.70</td>
</tr>
<tr>
<td>OVERALL</td>
<td>0.35</td>
<td>0.59</td>
</tr>
</tbody>
</table>

Table 1. Typical chemical screen analyses of concentrates obtained from the column and from the conventional Cu/Ni circuit.

Through the froth washing in the column the positive drainage flux enhanced copper-nickel separation, especially the fine fractions. Column amenability tests were done by the University of Toronto (3.8 cm column) and were confirmed in studies by McGill University (5.7 cm column). Figure 1 compares the amenability test results with plant data. The results were encouraging enough to design a pilot column, which measured 0.91 m in diameter and 13.7 m in height. Data from this column is also shown in Figure 1. The pilot column performed marginally better than expected from the data attained in the laboratory scale columns.

Fig.1. Performance comparison between pilot column (0.9 m diameter), 3.8 cm column and typical plant (mechanical cell) performance.
The pilot column data was used in a simulation program (Castillo et al., 1988) to help design plant scale columns. Two 2.13 m diameter columns were constructed in the concentrator early 1989, one to handle the primary rougher concentrate and one to handle the rougher concentrate from a separate copper circuit. Each column will produce a final grade copper concentrate; both column tailings streams will be scavenged in conventional flotation cells (already existing), circulating the scavenger cleaner concentrate back to the feed tank of each column.

**Pilot Column Design and Operation**

The pilot column can treat up to 30% of the total primary rougher concentrate. As it operates in open circuit, potential losses in the tailings are minimized by retreatment in the conventional copper circuit. The operation of the column is almost entirely controlled by Taylor Mod 30 controllers. A few parameters, such as level and holdup, require calculations which are too complex for the Mod 30 to handle. These calculations are done by an Analog Devices Micromac 6000 computer before the calculated signals are output to the individual controllers.

The control strategy for the full sized columns will be similar to the strategy used in the pilot column. Bias and level are the two principal control loops. To ensure a positive bias, the tailings flowrate setpoint is cascaded from the setpoints value to the feed flow controller, to which a constant is added. In a separate control loop the feed flow is regulated with a flow control valve. Level is maintained by controlling water addition to the spray header. It is measured by a pressure transmitter located below, but near, the pulp interface. This signal is being calibrated frequently using a portable conductivity probe (Gomez et al., 1989). The same signal, together with the signal from an additional pressure transmitter located about 2 m above the spargers, is processed in the computer to calculate collection zone pressure drop. Air holdup is calculated using this data and the slurry density within the column; tailings slurry density is used to estimate slurry density within the collection zone. In the pilot column set up, control loops to regulate holdup and feed density can be activated when desired, but this has been done for test purposes only.

A cloth type sparger was employed in the testwork reported here. Test results for different sparger arrangements will be published elsewhere. For the pilot column the sparger system consists of an air manifold that feeds six pipes on either side of the column. The two sets of pipes are stacked above each other at a distance of 0.25 m. Each pipe is clad with two layers of cloth, a tightly woven T-100 Beane Bag as the inner cloth, and a Neotex 3570 on the outside to provide strength. The Neotex 3570 was chosen for its resistance to scaling under high pH conditions. Sparger surface area to column area ratio is at about 2.3.

Some operators prefer a shower-head type spray, located about 0.15 m above the froth. Most use a submerged spray header, its distance from the column lip seems arbitrary but, generally, varies up to a maximum distance of 0.25 m. Both types were tested in the pilot column.
Fig. 2. Copper circuit performance with and without the pilot column in operation.

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Range</th>
<th>Unit</th>
</tr>
</thead>
<tbody>
<tr>
<td>Feed Tonnage</td>
<td>5.5 - 6.0</td>
<td>tonnes/h</td>
</tr>
<tr>
<td>Feed Flow Rate</td>
<td>0.7</td>
<td>cm/s (superficial velocity)</td>
</tr>
<tr>
<td>Feed Density</td>
<td>25 - 35</td>
<td>%</td>
</tr>
<tr>
<td>Particle Size ($d_{50}$)</td>
<td>50</td>
<td>microns</td>
</tr>
<tr>
<td>Feed Grade</td>
<td>7 - 11.9 - 13</td>
<td>% Cu, %Ni</td>
</tr>
<tr>
<td>Gas Velocity</td>
<td>0.9 - 1.2</td>
<td>cm/s</td>
</tr>
<tr>
<td>Bubble Diameter</td>
<td>0.9 - 1.2</td>
<td>mm</td>
</tr>
<tr>
<td>Froth Depth</td>
<td>1.0</td>
<td>m</td>
</tr>
<tr>
<td>Bias Velocity</td>
<td>0.07 - 0.1</td>
<td>cm/s</td>
</tr>
<tr>
<td>Washwater Velocity</td>
<td>0.2 - 0.3</td>
<td>cm/s</td>
</tr>
</tbody>
</table>

Table 2. Typical operating parameters in pilot column operation
Figure 2 illustrates the performance of the column on a continuous base, and compares the result with the plant. The column appeared able to reproduce consistently a copper concentrate containing about 0.4 % Ni.
Typical conditions for the pilot column operation are summarized in Table 2.

EVALUATION OF SELECTED OPERATING AND DESIGN PARAMETERS

The pilot column has been run under carefully controlled conditions in order to assess the effect of critical operating and design parameters. For each test the column was brought to steady state, waiting at least three retention times from the time of parameter change to the start of sampling. Composite samples over a 30 minute period were taken, while feed flowrate and slurry density were controlled at the desired levels. The results that follow summarize observations on the effects of gas rate, froth depth, feed slurry density and water spray position. Unless otherwise noted, the washwater spray was located 10 cm below the lip. (A separate paper will describe sparger performance obtained with both filter cloth and the U.S.B.M. sparger systems.)

Gas Rate

Gas velocity was varied from 0.65 to 1.45 cm/s (STP) for feed flow at a typical rate (6.2 tph solids, 260 L/min slurry, 8.0 % Cu in the feed) as well as at a low rate (3.3 tph, 170 L/min, 11.0 % Cu in the feed). Figs. 3a and 3b summarize the metallurgical performance.

Gas flow above about 1.3 cm/s has a significant negative effect on concentrate quality while providing little, if any, additional Cu recovery. Results from chemical screening analyses indicated that the inferior concentrate quality was due to entrainment of -37 micron gangue and nickel mineral particles. As well, higher gas rates drive up the water consumption and yield a more dilute concentrate (Fig. 4). This effect has been observed by others, e.g. Kosick et al. (1988), in the flotation of galena.

Bubble diameter \( d_b \) was estimated from measurements of gas holdup, gas velocity and slurry velocity using a method described elsewhere (Dobby et al., 1988). For the tests at a feed rate of 6.2 tph, \( d_b \) varied from 0.9 mm at 0.9 cm/s gas velocity to 1.1 mm at 1.4 cm/s; for the tests at a feed rate of 3.3 tph \( d_b \) varied from 0.7 mm at 0.7 cm/s gas velocity to 0.8 mm at 1.3 cm/s. Thus, in the column test conducted at the lower feed rate, bubble size was smaller and retention time was longer, yet Cu recovery was lower than in the test conducted at the higher feed rate (at \( J_g \), less than 1.2 cm/s). This implies that the lower Cu recovery obtained at the low solids rate is not due to poor collection kinetics but rather to higher froth zone dropback, i.e. less froth stability due to low solids loading. Evidence to support this is seen in Fig. 3b where, at \( J_g = 1.3 \text{ cm/s} \), Cu recovery is similar, but Ni grade is considerably lower in the test conducted at low flowrate; this is an outcome expected from a high circulating load (Dobby, 1984).
Froth Depth

Froth depth was varied from 0.75 m to 1.5 m. As expected, copper recovery decreased and concentrate grade increased (% Ni in concentrate decreased) with increasing froth depth (Figs. 5a and b). At this gas velocity (0.93 cm/s) a froth depth of 1.0 to 1.2 m appears to be preferred; froths shallower than this result in considerable deterioration in concentrate quality, while deeper froths provide relatively little benefit to concentrate grade. Similar observations were made from tests conducted at low solids feed rate.

Fig.3a. The effect of gas rate on copper recovery to pilot column concentrate. $H_f = 1.0$ m.

Fig.3b. Effect of gas rate on Ni grade of pilot column concentrate.
Fig.4. Effect of gas velocity on washwater consumption.

Fig.5a. Effect of froth depth on copper recovery. Conditions: $J_g = 0.93 \text{ cm/s}$; feed rate = 6.2 tph.

Washwater consumption as affected by froth depth is shown in Fig. 6. The decreasing washwater rate with increasing froth depth is an indirect result of the control scheme employed. When froth depth is increased a greater proportion of the added washwater reports to the bias flow; therefore, to attain a certain bias (which remains constant in this
control approach) will require less total washwater. The increase in the proportion of water reporting to the bias flow for the deeper froth may be attributed to either increased water retention time in the froth or higher packing density of the bubbles which in turn causes the froth to be less fast flowing. However, at this point these are speculations, and improved understanding of the froth is needed.

Feed Slurry Density

The effect of feed slurry density upon copper recovery is shown in Fig. 7 for two sets of tests. In these tests slurry density was varied by diluting the feed slurry while still maintaining a constant solids feed rate, i.e. feed slurry rate increased and retention time decreased with a decrease in slurry density. In test set A copper recovery is high and only

![Graph showing the effect of froth depth on concentrate Ni grade.](image)

Fig.5b. Effect of froth depth on concentrate Ni grade.

slightly affected by slurry density, while in test set B copper recovery is low and increases with increasing slurry density.

It is felt that the difference in performance between the two sets of tests is a result of reagent conditions, as reflected by bubble size; the average bubble size was 1.0 mm for test set A and 1.7 mm for test set B. Plant conditions in test set B suffered from abnormally high dosages of reagents.
Fig. 6. Effect of froth depth on washwater consumption.

Fig. 7. The effect of feed slurry density on copper recovery. Solids feed rate is constant at 5.5 tph for both Test sets A and B. Other conditions for both Test sets: $H_t = 1.0$ m; $J_g = 0.9$ cm/s.

**Spray Head Depth**

The wastewater spray head was designed to be positioned at any vertical location between 100 cm below the lip and 10 cm above the lip. Its location influenced metallurgical performance as is shown in Figs. 8a and b. The major impact was on the concentrate quality, with relatively little effect on copper recovery. There is clearly a dramatic effect of spray head location upon concentrate slurry density (Fig. 9). The concentrate was about
Fig. 8a. Effect of spray head location on copper recovery. Conditions: feed rate 6.2 tph, $J_g = 0.9$ cm/s.

Fig. 8b. Effect of spray head location on concentrate Ni grade.

35% solids when the spray head was immersed in the froth, and below 20% solids when the spray head was above the froth. When water is added above the froth there is a strong tendency for bypassing of water directly to the column overflow.

**Column Optimizing Control Summary**

Gas rate is clearly a suitable parameter for grade-recovery manipulation when gas flow is less than about 1.3 cm/s (Fig. 3a and b). At a higher gas rate, the performance in recovery is not enhanced significantly, but the concentrate quality becomes inferior. The control strategy will be to optimize the recovery of copper by maximizing the gas rate to the point where the nickel content in the concentrate will become the constraining factor.
Fig.9. Effect of spray head location on concentrate slurry density.

CONCLUSIONS

Column flotation provides the potential to produce a more effective copper nickel separation than conventional flotation cells. The resulting improvement in concentrate quality, in terms of lower nickel content, is attained mainly through better washing of the -37 micron size fraction. Two plant scale columns have been constructed in the concentrator earlier this year based on the results of amenability tests done on 3.8 cm and 5.7 cm columns, and operation of a 0.91 m pilot column. The gas rate strongly affects the grade of copper concentrate. Above 1.3 cm/s gangue and nickel mineral particles smaller than 37 micron become entrapped in the froth. Washing is inadequate, despite the increase in washwater rate. The grade improves with a higher froth depth, but recovery drops; deeper froth resulted in a higher concentrate density.

When recovery is high it is affected slightly by slurry density. A low recovery often is the result of reagent conditions, and can be improved by an increase in feed density.

The best concentrate quality can be attained when water is sprayed on top of the froth. For a submerged spray head it is best to locate it as near to the lip as possible.
ACKNOWLEDGEMENTS

The authors are indebted to Strathcona Mill management who allowed the testwork to take place, to Larry E. Seeley, Manager of Metallurgical Technology at Falconbridge for his sponsorship, to McGill University for their contributions and to Falconbridge Limited, who gave permission for publication of this paper.

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2.9. BUBBLE GENERATION ASSESSMENT FOR AN INDUSTRIAL FLOTATION COLUMN

B.J. Huls, C.D. Lachance, G.S. Dobby

Minerals Engineering, in press.

ABSTRACT

A 0.9 m diameter pilot flotation column has been in operation at Falconbridge's Strathcona concentrator for the purpose of Cu-Ni separation. The results from this column have been used to design a full scale column circuit to treat all of the Cu-Ni separation feed. To design a bubble generation system for the full scale column circuit extensive tests have been run with the pilot column, which is instrumented to measure all process flows, froth depth, gas holdup and bubble diameter. Two sparging systems have been studied: a cloth sparger and an external contacting system. This paper reports the bubble generation characteristics and metallurgical performance of the two systems.

KEYWORDS

Column flotation, Cu-Ni separation, bubble generation, spargers, bubble size

BACKGROUND

Falconbridge Limited is the second largest world producer of nickel and a major producer of copper. The company operates mines, a mill and a smelter complex in the Sudbury, Ontario area. Five mines currently feed the mill at a rate of about 9000 tonnes per day of ore containing about 1.5 % Ni and 1.4 % Cu. The principal nickel bearing mineral is pentlandite, with a small but significant source of nickel occurring in pyrrhotite. Chalcopyrite is the copper bearing mineral. The mill flowsheet (Scales et al., 1986) reflects a combination of grinding, classification, flotation, magnetic separation and dewatering. Primary rougher concentrate is pumped to a Cu-Ni separation circuit, from which the nickel concentrate is combined with the secondary rougher concentrate. Magnetic separation splits the scavenger concentrate to allow individual flotation of the non-magnetic and magnetic fractions for optimum rejection of pyrrhotite. In 1987 a separate copper circuit (C-circuit) was commissioned for a high grade copper ore. Copper and nickel concentrates from this circuit are combined with those from the original circuit.

In 1986 column flotation was proposed as an alternative to the conventional method of separating copper and nickel minerals, which employed 1.4 m³ Denver flotation cells in a closed circuit rougher cleaner configuration. Cu-Ni separation is effected at high pH
(through lime addition) and with the addition of sodium cyanide. Historically, it has proven difficult to control the nickel grade in the copper concentrate to an acceptable level. The concentrate typically assays 0.8 % to 0.9 % Ni and 27 % to 29 % Cu. It is desired to maintain the nickel content below 0.6 %.

Column flotation in a 0.9 m diameter pilot unit proved successful. The effects of gas rate, froth depth and spray head depth on separation performance with this column have been discussed previously (Huls et al., 1989). As an outcome of the pilot column results, two 2.1 m diameter by 12 m high columns, one to treat primary rougher concentrate and one to treat rougher concentrate from C-circuit, were designed and commissioned in June, 1989.

This paper summarizes the sparger evaluation test program that was conducted on the 0.9 m pilot column.

EXPERIMENTAL

Sparger Designs

Two sparger designs were evaluated: a cloth sparger system and an external air-water contacting system originally developed by the U.S.B.M.

When originally constructed the pilot column was equipped with two air manifolds to accommodate two sets of six cloth spargers each. The layout and dimensions are shown in Figure 1. The two rows of sparging elements were separated by 25 cm. Each tube was 3.2 cm in diameter and was perforated with 0.94 mm diameter holes, with 26 holes per meter of tube length. The tubes were covered with two layers of cloth, a tightly woven T-100 Beane Bag as the inner cloth and a Neotex 3570 on the outside to provide strength. The Neotex 3570 was chosen for its resistance to scaling under conditions of high pH. The ratio of sparger surface area to column cross sectional area was 2.3; this value approaches the recommended ratio established by Manqui and Finch (1988).

The external contacting system was also installed. In this system, air and water (both at high pressure) were contacted in a chamber containing a bed of ground quartz or ceramic spheres (Foot et al., 1986; McKay et al., 1988). The discharge of the chamber was carried into the column through four steel rods, and was expelled through holes about 1 mm in diameter. The bubble size generated with this system was controlled through air and water addition rates only; no frother addition to the sparger water was made during the tests. The piping was also arranged to allow bypass of the contacting chamber for some of the tests.

Pilot Column Measurement and Control

Feed to the pilot column was pumped with a dedicated line from the sump that fed main circuit, and feed rate was controlled through a pinch valve. Feed and tailings flows and densities were measured with magnetic flow meters. Stabilizing control of the column level
Fig. 1. Cloth sparger assembly on the 0.9 m diameter pilot column. All dimensions are in millimetres.
(i.e. Froth depth) was achieved by the combination of: (a) maintaining an operator selected difference between the water flows in the tailings and the feed, and (b) automatic wash water addition to maintain a constant pressure at a point 3.1 m below the top of the column (this stabilizing control approach is known conventionally as bias control). A pressure measurement was also made 11.9 m from the top of the column (1.2 m above the spargers). Pressure measurements were performed with DP-cells.

Estimation of Bubble Size

Bubble size from each test was estimated using the drift flux approach (Dobby et al., 1988; Yianatos et al., 1988). To make this calculation the following information was required: gas flow rate, pressure drop in the collection zone, tailings slurry flow rate, and tailings slurry density. Gas holdup in the collection zone (averaged over the distance in between the two pressure taps) was calculated using the pressure drop and tailings slurry density readings (actual slurry density in the collection zone is lower than that in the tailings stream, due to settling of the solid particles; the correction for this has been described by Dobby et al. (1988)).

Testwork

With relatively little manipulation, the air sparger system could be switched from cloth to the external contacting system, with or without the contacting chamber, while the column was in operation. For each test the column was brought to steady state, waiting at least three retention times from the time of a parameter change to the start of sampling. Composite samples over a 30 minute period were taken, while feed flow rate and slurry density were controlled at the desired levels.

RESULTS AND DISCUSSION

More than 40 tests were conducted to assess the characteristics of the two sparging systems. A summary of the results is given in Tables 1 to 3. In these Tables, the following conventions are adopted:

- $\dot{J}_w$ is wash water superficial rate (volume flow rate divided by column cross-sectional area),
- $\dot{J}_g$ is gas superficial rate (at the top of the column),
- $E_g$ is gas holdup (at the middle of the collection zone), and
- $d_b$ is bubble diameter (at the middle of the collection zone) as estimated from the drift-flux calculations.

Table 1 summarizes the results, at a feed grade of about 9 % Cu and 12 % Ni, for three cases: cloth sparger, external contacting sparger with the contacting chamber in place, and external contacting sparger without the contacting sparger. Feed rate in all tests was 265 L/min slurry, and gas rate was also constant, at 0.86 cm/s. The external contacting sparger system produced gas bubbles slightly smaller than those produced by the cloth sparger. This resulted in a somewhat higher wash water addition rate for the external contacting sparger. By comparing copper recovery and nickel recovery it is seen that separation efficiency was
virtually identical. The external sparger yielded similar results with and without the contacting chamber in place.

The effect of varying system pressure (i.e. sparger water rate) for the external contacting sparger is summarized for two cases in Table 2. System pressure, measured after the gas-water contacting point, was varied from 30 to 60 psig. For both cases gas rate is constant,

<table>
<thead>
<tr>
<th>Sparger Type</th>
<th>Feed %Cu</th>
<th>Feed %Cu</th>
<th>Feed %Ni</th>
<th>Feed Cu Rec</th>
<th>Feed Ni Rec</th>
<th>$J_w$ (cm/s)</th>
<th>$J_q$ (cm/s)</th>
<th>$F_q$ (g)</th>
<th>$d_b$ (mm)</th>
</tr>
</thead>
<tbody>
<tr>
<td>cloth 9.8</td>
<td>30.6</td>
<td>0.44</td>
<td>68</td>
<td>0.79</td>
<td>0.15</td>
<td>0.86</td>
<td>6.5</td>
<td>0.84</td>
<td></td>
</tr>
<tr>
<td>cloth 9.5</td>
<td>30.2</td>
<td>0.43</td>
<td>60</td>
<td>0.78</td>
<td>0.16</td>
<td>0.86</td>
<td>6.5</td>
<td>0.85</td>
<td></td>
</tr>
<tr>
<td>cloth 9.5</td>
<td>30.0</td>
<td>0.45</td>
<td>62</td>
<td>0.72</td>
<td>0.20</td>
<td>0.86</td>
<td>6.1</td>
<td>0.90</td>
<td></td>
</tr>
<tr>
<td>Average 9.6</td>
<td>30.3</td>
<td>0.44</td>
<td>63</td>
<td>0.76</td>
<td>0.17</td>
<td>0.86</td>
<td>6.4</td>
<td>0.86</td>
<td></td>
</tr>
<tr>
<td>External 9.3</td>
<td>31.2</td>
<td>0.39</td>
<td>61</td>
<td>0.76</td>
<td>0.22</td>
<td>0.86</td>
<td>7.4</td>
<td>0.76</td>
<td></td>
</tr>
<tr>
<td>External 8.9</td>
<td>30.7</td>
<td>0.36</td>
<td>62</td>
<td>0.57</td>
<td>0.21</td>
<td>0.86</td>
<td>7.8</td>
<td>0.73</td>
<td></td>
</tr>
<tr>
<td>External 8.8</td>
<td>30.8</td>
<td>0.36</td>
<td>59</td>
<td>0.62</td>
<td>0.22</td>
<td>0.86</td>
<td>6.8</td>
<td>0.82</td>
<td></td>
</tr>
<tr>
<td>External 8.7</td>
<td>30.6</td>
<td>0.40</td>
<td>61</td>
<td>0.68</td>
<td>0.22</td>
<td>0.86</td>
<td>7.0</td>
<td>0.80</td>
<td></td>
</tr>
<tr>
<td>External 8.8</td>
<td>30.3</td>
<td>0.44</td>
<td>66</td>
<td>0.85</td>
<td>0.25</td>
<td>0.86</td>
<td>7.1</td>
<td>0.79</td>
<td></td>
</tr>
<tr>
<td>External 8.8</td>
<td>30.3</td>
<td>0.43</td>
<td>65</td>
<td>0.84</td>
<td>0.22</td>
<td>0.86</td>
<td>7.8</td>
<td>0.73</td>
<td></td>
</tr>
<tr>
<td>Average 8.8</td>
<td>30.7</td>
<td>0.44</td>
<td>63</td>
<td>0.81</td>
<td>0.25</td>
<td>0.86</td>
<td>7.0</td>
<td>0.80</td>
<td></td>
</tr>
<tr>
<td>External* 8.8</td>
<td>30.8</td>
<td>0.44</td>
<td>64</td>
<td>0.79</td>
<td>0.25</td>
<td>0.86</td>
<td>7.1</td>
<td>0.79</td>
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<td>31.0</td>
<td>0.43</td>
<td>65</td>
<td>0.83</td>
<td>0.19</td>
<td>0.86</td>
<td>8.0</td>
<td>0.72</td>
<td></td>
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<tr>
<td>External* 8.9</td>
<td>30.2</td>
<td>0.41</td>
<td>67</td>
<td>0.77</td>
<td>0.26</td>
<td>0.86</td>
<td>7.6</td>
<td>0.75</td>
<td></td>
</tr>
<tr>
<td>External* 8.8</td>
<td>30.2</td>
<td>0.43</td>
<td>63</td>
<td>0.78</td>
<td>0.24</td>
<td>0.86</td>
<td>7.1</td>
<td>0.79</td>
<td></td>
</tr>
<tr>
<td>Average 8.9</td>
<td>30.4</td>
<td>0.43</td>
<td>65</td>
<td>0.79</td>
<td>0.23</td>
<td>0.86</td>
<td>7.5</td>
<td>0.76</td>
<td></td>
</tr>
</tbody>
</table>

* Operated without contacting chamber

Table 1. Summary of sparger performances: Cloth Sparger and External Contacting Sparger.

except for the 60 psi test, where gas rate is slightly lower. Copper recovery increases with increasing system pressure. At the same time, bubble diameter decreases from about 1.25 mm at 30 psi to about 1.0 mm at 50 psi. The higher recovery, and corresponding lower grade, at the higher system pressure is therefore attributed to the generation of a smaller bubble.

Gas rate was varied for the filter cloth sparging system and the results from two cases are summarized in Table 3. Copper recovery increased with increasing gas rate up to about 1.2 cm/s and appears to plateau at about this level. Increasing gas flow above about 1.2 cm/s acted to decrease concentrate grade dramatically, likely because of a significant increase in entrainment recovery. In Case C the bubble diameter increased from about 0.5 mm to 0.6
mm when gas flow was increased from 0.7 to 1.1 cm/s; this is the expected trend. No trend is observed between bubble size and gas rate in Case D.

The separation performance from the tests reported in Tables 1, 2 and 3 is summarized in Figure 2, where nickel recovery is plotted versus copper recovery. Either an increase in gas rate or an increase in the system pressure (for the external contacting system) moves the separation from low to higher copper recovery. The clear point to note is that, over the range of performance studied here, neither sparger system produced superior separation.

<table>
<thead>
<tr>
<th>Pressure (PSI)</th>
<th>Feed %Cu</th>
<th>Feed %Ni</th>
<th>Conc Cu Rec</th>
<th>Conc Ni Rec</th>
<th>J_w (cm/s)</th>
<th>J_g (cm/s)</th>
<th>E_g (%)</th>
<th>d_b (mm)</th>
</tr>
</thead>
<tbody>
<tr>
<td>CASE A</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>30</td>
<td>9.0</td>
<td>32.0</td>
<td>0.29</td>
<td>57</td>
<td>0.40</td>
<td>0.15</td>
<td>1.01</td>
<td>4.7</td>
</tr>
<tr>
<td>40</td>
<td>8.2</td>
<td>30.6</td>
<td>0.38</td>
<td>70</td>
<td>0.85</td>
<td>0.16</td>
<td>1.01</td>
<td>5.3</td>
</tr>
<tr>
<td>50</td>
<td>7.8</td>
<td>29.2</td>
<td>0.56</td>
<td>80</td>
<td>1.90</td>
<td>0.18</td>
<td>1.01</td>
<td>5.3</td>
</tr>
<tr>
<td>60</td>
<td>8.9</td>
<td>29.8</td>
<td>0.47</td>
<td>79</td>
<td>1.13</td>
<td>0.18</td>
<td>0.72</td>
<td>5.3</td>
</tr>
<tr>
<td>CASE B</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
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<tr>
<td>30</td>
<td>20.3</td>
<td>30.5</td>
<td>0.41</td>
<td>58</td>
<td>8.6</td>
<td>0.25</td>
<td>1.08</td>
<td>5.5</td>
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<tr>
<td>40</td>
<td>20.9</td>
<td>29.7</td>
<td>0.49</td>
<td>67</td>
<td>14.6</td>
<td>0.32</td>
<td>1.08</td>
<td>7.2</td>
</tr>
<tr>
<td>50</td>
<td>21.1</td>
<td>29.1</td>
<td>0.67</td>
<td>76</td>
<td>21.6</td>
<td>0.41</td>
<td>1.01</td>
<td>7.5</td>
</tr>
<tr>
<td>60</td>
<td>21.1</td>
<td>26.5</td>
<td>1.21</td>
<td>87</td>
<td>42.8</td>
<td>0.65</td>
<td>0.86</td>
<td>7.4</td>
</tr>
</tbody>
</table>

Table 2. Effect of gas pressure on performance of External Contacting Sparger.

<table>
<thead>
<tr>
<th>Gas Rate (L/min)</th>
<th>Feed %Cu</th>
<th>Feed %Ni</th>
<th>Conc Cu Rec</th>
<th>Conc Ni Rec</th>
<th>J_w (cm/s)</th>
<th>J_g (cm/s)</th>
<th>E_g (%)</th>
<th>d_b (mm)</th>
</tr>
</thead>
<tbody>
<tr>
<td>CASE C</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>283</td>
<td>11.9</td>
<td>31.0</td>
<td>0.39</td>
<td>47</td>
<td>0.39</td>
<td>0.13</td>
<td>0.72</td>
<td>15.3</td>
</tr>
<tr>
<td>340</td>
<td>11.0</td>
<td>31.4</td>
<td>0.33</td>
<td>47</td>
<td>0.33</td>
<td>0.11</td>
<td>0.86</td>
<td>15.5</td>
</tr>
<tr>
<td>396</td>
<td>11.0</td>
<td>31.6</td>
<td>0.46</td>
<td>60</td>
<td>0.46</td>
<td>0.13</td>
<td>1.01</td>
<td>17.0</td>
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<tr>
<td>453</td>
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<td>31.0</td>
<td>0.81</td>
<td>72</td>
<td>0.81</td>
<td>0.14</td>
<td>1.15</td>
<td>17.9</td>
</tr>
<tr>
<td>CASE D</td>
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<tr>
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<td>0.43</td>
<td>59</td>
<td>0.7</td>
<td>0.16</td>
<td>0.65</td>
<td>4.0</td>
</tr>
<tr>
<td>340</td>
<td>8.8</td>
<td>30.0</td>
<td>0.45</td>
<td>64</td>
<td>0.8</td>
<td>0.18</td>
<td>0.86</td>
<td>5.9</td>
</tr>
<tr>
<td>396</td>
<td>7.7</td>
<td>28.2</td>
<td>0.68</td>
<td>74</td>
<td>1.35</td>
<td>0.26</td>
<td>1.01</td>
<td>5.6</td>
</tr>
<tr>
<td>510</td>
<td>7.6</td>
<td>26.4</td>
<td>1.32</td>
<td>80</td>
<td>2.84</td>
<td>0.41</td>
<td>1.29</td>
<td>8.1</td>
</tr>
<tr>
<td>566</td>
<td>7.2</td>
<td>22.0</td>
<td>2.85</td>
<td>76</td>
<td>7.71</td>
<td>0.84</td>
<td>1.44</td>
<td>9.3</td>
</tr>
</tbody>
</table>

Table 3. Effect of gas rate on performance of Cloth Sparger.
performance. This is evidence that either bubble size or gas rate can be manipulated to move a separation system along its grade-recovery curve. When gas rate has been increased to the point of maximum recovery ($J_g$ approximately 1.2 cm/s in this case) then decreasing bubble size may help to increase recovery further. There is a limit to this, however; calculations have indicated that there is little increase in recovery expected when gas rate is already at its maximum and when bubble diameter is decreased below about 1 mm (Dobby and Finch, 1986).

When gas is an expensive consumable, such as when using nitrogen, producing a smaller gas bubble obviously will be advantageous.

![Graph](image)

**Fig.2.** Nickel recovery versus copper recovery for the data in Tables 1, 2, and 3 (Case B of Table 2 not included).
CONCLUSIONS

1. In the application considered here, increasing gas rate and decreasing the bubble size had similar effects on separation performance.

2. The contacting chamber of the external contacting sparging system did not play a significant role in bubble generation.

3. The external contacting system provides the plant operator with the flexibility to move along the grade-recovery curve by modifying either gas rate or system pressure, i.e. bubble size.

ACKNOWLEDGEMENTS

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REFERENCES


SUMMARY

In the introductory section mineral sources were divided into categories that are distinguished by their concentration process. For sulphides, to which this dissertation is restricted, the typical beneficiation steps are crushing, grinding, classification, flotation, dewatering, and sometimes, magnetic separation. This section alludes to the progress made in improving equipment to separate minerals more effectively and economically. Specifically, the focus is on changes in philosophy in operating unit processes, emphasizing process control.

The first chapter discusses both the external factors and those factors inherent to the very ore body, each determining which equipment or process parameters should be selected. In processing sulphide ores, the interaction between grinding and flotation is of utmost importance. Both the degree of liberation required and the occurrence of preferential grinding may dictate the product size to be attained in the grinding process. Size-by-size recovery data aid into identifying that particular product size. In addition to these parameters, the essence of stability within a grinding circuit also is expressed. Paper One in Part Two describes a mechanism and method by which stability in a grinding circuit can be achieved, namely the Particle Size Monitor. In the second paper this Monitor is shown to become the central point in a stabilizing and optimizing grinding control strategy.

In Chapter 1.2 the fundamentals of grinding are extended to the application of models, their deficiencies being highlighted. These include sampling methods, which form the basis of model development, the lack of information about slurry rheology, and the complexity of models. Furthermore, the conventional operating philosophy of a grinding circuit is discussed and is contrasted against process control techniques, which are elaborated upon in Chapter 1.4. Optimization of a grinding circuit through the implementation of control is preceded by elimination of bottlenecks to render the most expensive piece of equipment, the mill, as the sole restricting element. For example, one way to expand classifier capacity in a confined area could be by installing cyclones horizontally. This route should only be taken for cyclones in excess of 0.5 m in diameter that are subjected to a pressure drop of 70 kPa. In this position, the size separation curve becomes sharper, resulting in a drop in circulating load of at least 200% absolute.

In discussing the theory of flotation, a correlation is presented between the wettability of a solid, the surface tension of a wetting agent, and the natural hydrophobicity found for many sulphide minerals. Zisman defined the critical surface tension of wetting below which spontaneous spreading of a liquid across a surface occurs. This value for galena was found to be nearly equivalent to those of both allotropic forms of sulphur, suggesting the presence of elemental sulphur on the surface of galena due to oxidation. Most sulphide minerals were found to exhibit some form of elemental sulphur, a metal-deficient sulphide layer, or a poly-sulphide on their surface. Such presence leads to the natural floatability of these minerals as sulphur in whichever form is thought to be the hydrophobic entity, and presumably plays a role in the adsorption of collectors.
Besides particle size, pulp density is important in flotation. By diluting the pulp the flotation response of coarse particles is improved. With less particles present, the probability of detachment decreases, because the buoyancy of a particle-bubble aggregate is determined by the amount of particles attached to the bubble. Fewer coarse particles must attach if reaching the froth layer, let alone the cell lip, is to be ensured. For particles larger than 210 microns the recovery was shown to improve by at least 10%, when diluting the pulp a few percent.

To determine the adequate pulp density and the desired pulp potential to float at, represent steps in the managerial philosophy proposed in Chapter 1.4. An incremental approach is emphasized, starting at the beginning of the process to define the real maximum capacities of each unit process at optimum operating conditions. Most often, this will materialize in expanded use of existing equipment and facilities.

Column flotation is highly amenable to process control, as is outlined in the third paper. The extent of control employed depends on the application. A particular subject of continued investigation is the froth zone. Related is the froth drop back of solids into the upper portion of the collection zone, which is a measure of selectivity. This chapter brings forward a proposed mechanism to deal with consistent recovery levels, up to a point independent of changes in floatable feed mass. The philosophy is to slowly alter the interface position depending on the floatable mass that enters the column. It is known that the heavier the froth, the more material will return to the collection zone.

Papers 4 to 7 discuss the development of a probe to measure the height of the interface, based on conductivity. The text in Chapter 1.5 expands on conductivity measurements to infer from these gas hold-up and bias as well. This device has the potential to greatly simplify the required instrumentation around a column, when enhanced control is desired.

Direct scale-up from flotation rate constants did not prove to be satisfactory; better results were obtained after calibrating the simulator to the experimental results through slightly adjusting the flotation rate constants obtained in the test work. The new constants were then applied to predict the performance of the industrial size columns.
Samenvatting

De inleiding behandelt de verscheidene categorieën waarin de mineralen zijn onderverdeeld. Elke categorie onderscheidt zich van de andere door zijn karakteristieke concentratie proces. Ertsscheidingsmethoden voor de sulfiden groep bestaan uit breken, malen, classificeren, floteren, ontwateren en, in bijzondere gevallen, magneetscheiding. Dit gedeelte geeft aan dat in de laatste tijd belangrijke verbeteringen zijn aangebracht om mineralen beter en ekonomischer te scheiden. Met name op de manier tot beheersing van het proces wordt de aandacht gericht.

De selektie van de afzonderlijke processtappen is onderhevig aan de verschillende invloeds factoren die van buitenaf zijn opgelegd, maar ook bepaald worden door het ertsvoorkomen zelf. Het maalproces wordt gekarakteriseerd door de eisen die flotatie aan malen stelt, maar ook andersom beoefent het malen een invloed uit op de manier waarop flotatie wordt verricht. De mineralen moeten in zekere mate vrij gemaakt worden, terwijl sommige mineralen gemakkelijker verkleinen dan andere. Door tijdens de flotatie de winning per korrelklasse te bepalen kan men aangeven tot hoeveel gemalen moet worden. Daar komt bij dat ook de stabilitéit in het maalcircuit van groot belang is. Het eerste artikel beschrijft dan ook een apparaat en methode waarmee zulk een stabiliteit verkregen kan worden, namelijk de Particle Size Monitor. Dit instrument vormt het centrale punt in een strategie tot beheersing van het maalproces ter verkrijging van stabilitéit en optimalisatie, zoals in het tweede artikel is aangegeven.

De grondbeginselen van malen worden in hoofdstuk 1.2 aan de kaak gesteld en uitgebreid tot de toepassing van modellen, die echter onderhevig zijn aan ontoereikendheden; bijvoorbeeld de manier waarop wordt gemonsterrd, hetgeen een basis vormt voor de modellen, verder het gebrek aan informatie over de rheologie van de vloeistofmassa, en de complexiteit van de modellen zelf. De gebruikelijke manier waarop in het maalcircuit wordt gewerkt, wordt getoetst aan de toepassing van procesbeheersing door middel van instrumentatie, waarop in hoofdstuk 1.4 verder wordt ingegaan. Voordat echter een maal circuit kan worden geautomatiseerd, zullen er verscheidene veranderingen moeten worden aangebracht om alle belemmeringen te verhelpen, zodat slechts de staaf- of kogelmolen, als duurste gereedschap, de beperkende eenheid in het circuit wordt. In beperkte ruimte, kunnen cyclonen bijvoorbeeld horizontaal worden gemonteerd, hoewel deze ingreep alleen zin heeft voor cyclonen groter dan 0.5 m in doorsnede en met een drukval geringer dan 70 kPa.

De discussie over de flotatietheorie legt een verband tussen de bevochtiging van een deeltje, de oppervlaktespanning van de vloeistofmassa, en de natuurlijke hydrophobiciteit van de sulfidische mineralen. Zisman definiëerde de kritieke oppervlaktespanning tot bevochtiging als die oppervlaktespanning waaronder een vloeistof zich spontaan uitspreidt over het oppervlakte van een vaste stof. De waarde van deze variabele voor galeniet bleek ongeveer dezelfde te zijn als die voor beide allotropische vormen van zwavel, hetgeen de veronderstelling wekte dat op het oppervlak zwavel wordt gevormd door enige oxidatie. Het merendeel van de metaalsulfiden blijkt bedekt te zijn met een vorm van een zwavel verbinding die ófwel bestaat uit elementaire zwavel, ófwel uit een metaalarme zwavelver-
rijkte laag, óf wel uit een polysulffide. Elk dezer verbindingen zou flotatie kunnen veroorzaken en de adsorptie van een collector kunnen bevorderen.

Van verder belang in de flotatie is de deeltjesgrootte, maar ook de dichtheid van de vloeistofmassa. In hoofdstuk 1.4 wordt aangetoond hoedanig de flotatie van de grootste korrelklassen wordt beïnvloed door verdunding van deze massa; minder deeltjes behoeven zich aan een luchtbel te hechten, waardoor de algemeene dichtheid niet het stijgen van de bel in gevaar brengt. Als de deeltjesdichtheid te groot zou zijn kan een bel nooit de schuimlaag bereiken. In de tekst is er een voorbeeld gegeven waar de winning in flotatie voor deeltjes groter dan 210 micronen met zeker 10 % absoluut toenam zodra de vloeistofmassa met enkele procenten werd verlaagd.

Het bepalen van de juiste verdunning en van de optimale potentiaal die zouden moeten worden toegepast tijdens het flotatieproces, vormen onderdelen van een filosofie ter behandeling van een capaciteitsverruiming. Met name wordt er nadruk gelegd op een stapsgewijze benadering om aan de wens tot uitbreiding te voldoen, waarbij men bij de eerste processtap moet aanvangen. Voor elke stap wordt dan onder optimale bedrijfsovergangen en de maximale vermogen bepaald. Over het algemeen zal dit leiden tot beter gebruik van bestaand materieel, voordat toevloed van grootkapitaal noodzakelijk wordt.

Vooral in kolomflotatie kan van procesbeheersing gebruik worden gemaakt, zoals in het derde artikel is aangegeven. De mate waartoe daarvan gebruik wordt gemaakt hangt af van de toepassing. Dit hoofdstuk behandelde de theorie van kolomflotatie en stelt dat er nog maar weinig bekend is over de schuimlaag. Hiermee staat de terugval van deeltejes naar de collection zone in verband, welke de selectiviteit verhoogt. In het geval dat het floteerbaar gedeelte in de voeding toeneemt, wordt een methode geopperd die binnen zekere grenzen de winning in de kolom maximaal houdt. Door de diepte van de schuimlaag te ver- minderen, kan namelijk de opbouw van het gewicht, dat op de onderste zone in die schuimlaag zou worden uitgeefend, worden beperkt. Zo niet, dan verhoogt de terugval van deeltjes uit de schuimlaag naar de collection zone.

Artikels 4 tot 7 behandelen de ontwikkeling van een meetapparaat voor het niveau van het scheidingssvlak in diepe schuimlagen, dat gebaseerd is op geleidingsvermogen. Vergeleken met de gebruikelijke methode om de hoogte van het scheidingssvlak te meten, door middel van drukverschil, is de nieuwe methode niet onderhevig aan veranderingen in bedrijfsovergangen, zoals verschil in dichtheid. Dit werk wordt ook uitgebreid behandeld in hoofdstuk 1.5, waar verder wordt ingegaan op toepassing van geleidingsvermogen voor de berekening van bias en gas hold-up. Uiteindelijk kan dit apparaat leiden tot een verminderd aantal instrumenten om het kolomflotatieproces volledig te beheersen.

Opschalen van een laboratorium kolom tot industrieformaat met behulp van de theorie zoals deze huidig bekend is, moet met zorg gebeuren. Het rechtstreeks toepassen van de flotatiekonstanten die op kleine schaal zijn bepaald kan tot niet realiseerbare produktie verwachtigen leiden. Beter is het om het simulatie pakket te toetsen aan verkregen resultaten, waarbij de flotatiekonstanten enigszins worden gewijzigd. De nieuwe waarden worden gebruikt om de produktie op grote schaal te voorspellen.
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CURRICULUM VITAE


In augustus, 1978 begon ik mijn loopbaan op de Research Afdeling van Sherritt Gordon in Fort Saskatchewan, Alberta, Canada, waar ik als research metallurgist te werk gesteld werd. Bij Sherritt hield ik me principieel bezig met de hydrometallurgie van nikkel en zink. In 1982 ging ik over naar Billiton in Leidschendam, die mij in 1983 uitzond naar de Cuajone Concentrator van Southern Peru Copper Corporation. Daar begon ik als metallurgist en klom op tot Technical Assistant of the Superintendent. In die positie had ik de leiding over alle metallurgische projecten alsmede de bedrijfsmetallurgie en het analytisch laboratorium. Als voornaamste projecten golden in die tijd kolomflotatie en procesbeheersing in het maalcircuit.

In 1987 keerde ik terug naar Canada waar ik verantwoordelijk werd gesteld voor de ertstechnologie binnen Falconbridge. In mijn huidige functie als Superintendent Mineral Processing draag ik zorg voor de modernisatie van de concentrator en voor de beperking van zwaveldioxide, die door de schoorsteen wordt uitgestoten. Dit laatste heeft verband met verminderde winning van pyrrhotien, een ijzer-zwavel verbinding dat ongeveer 0.6 % nikkel bevat. Het modernizeren van de concentrator gaat gepaard met de installatie van kolomflotatie, flotatie cellen van groot volume, proces beheersings methoden, en verbetering van het magneetscheidingscircuits. Ik heb me met name toegelegd op kolomflotatie en op procesbeheersing. Dit jaar ontving ik een onderscheiding voor uitnemende werkresultaten over het jaar 1989, de Golden Falcon Award.
Notes on joint authorship

For the papers with joint authorship the contribution of the author is indicated below.

B.J. Huls and E. Socolich. - Automatic Grind Control at the Cusione Copper Concentrator.
The first author wrote the paper. Design of test work and data evaluation was also done by the first author.

B.J. Huls, P.L. Dunstan and S. Donges. - Distributed Digital Control at Cusione.
The first two authors have written the paper jointly. The original design and engineering was performed in cooperation with Armaco Autometrics. The first author configured the system and redeveloped and implemented the process control for the grinding circuit.

B.J. Huls proposed the theory on which the paper is based and designed and supervised the test work. The paper was written after he had left this company, but his comments were included.

C.O. Gomez, A. Uribe-Salas, J.A. Finch and B.J. Huls. - Level Detection in Flotation Columns using Conductivity.
The last author was actively involved in the discussions and commentary on the paper, and participated in the test work.

The last author was actively involved in the preparation and supervision of the test work. The paper was written by McGill personnel after many discussions with the last author on its contents.

The last author had outlined the preferable shape of the probe, but details were worked out by McGill personnel. He also supervised the test work conducted in the pilot column at Falconbridge. Again the input into the paper of the last author was in the form of comments to the text proposed by McGill.

B.J. Huls, C.O. Gomez, A. Uribe-Salas, J.A. Finch. - Level Detection in Flotation Columns at Falconbridge.
This paper was written jointly by all four authors. The first author also supervised the test work conducted in both the pilot and plant scale columns at Falconbridge.

The first author designed, supervised and evaluated all experiments. The paper was written with the last author, while the second author performed the experiments.

The first author designed, supervised and evaluated all test work. The paper was written in conjunction with the last author. The second author conducted the experiments.