COLUMN FLOTATION
MODELLING AND TECHNOLOGY

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ABSTRACT

A brief review of the art of column flotation technology is presented. The flotation column characteristics are described, followed by an analysis of the principal variables involved in the process and the impact they have on column performance. Discussion is given of both industrial experience and fundamental studies.

Attempts to model and scale-up flotation columns are reviewed with emphasis on the key factors affecting column scale-up, i.e. rate constants, mixing characteristics, particle residence time and carrying capacity.

Existing control schemes for flotation columns are discussed, and a summary of column flotation applications is given illustrating the wisdespread use of this technology.

1. INTRODUCTION

An inherent limitation with the flotation of fine partcles in conventional cells is recovery of hydrophilic (gangue) particles by mechanical entrainment in the water reporting to the froth. The method of minimizing entrainment is to create a 5-30cm thick froth at the slurry surface. The froth permits the gangue to drain back to the pulp while retaining the hidrophobic particles which are eventually discharged over the cell lip. Another possibility is to dilute the pulp, with the consequent loss of capacity. These cleaning actions are seldom sufficient and sequential stage flotation is necessary.

An alternative approach to the cleaning of entrained particles has been realised with the introduction of column flotation. Flotation columns differ radically from conventional mechanical flotation units both in design and operating philosophy. This has been a cause of the reluctance to test these units industrially since their invention (in Canada) in the mid-60's by Boutin and Tremblay. This situation has changed dramatically since 1981. Columns are now standard in Mo upgrading, and testing and application on other circuits are becoming widespread (eg. bulk Cu/Mo cleaning at Gilbraltar, Cu/Ni separation at Inco, Cu/Mo cleaning at Magma Copper, Cu cleaning at Los Bronces, Exxon, Chile).

Because of the efficient cleaning action, flotation columns can upgrade a fine sized concentrate in a single step, where conventional flotation machines would require several sequential stages. Flotation columns offer improved metallurgy, simplified circuits and easier control compared to conventional cells (Coffin and Miszczak, 1982; Wheeler, 1985; Feeley et al., 1987).

Development and industrial applications of flotation columns have been described in a large number of references. An extensive review has been recently presented by Finch and Dobby (1989).

Fundamental studies were slow to commence. Sastri and Fuerstenau (1970) and Rice et al. (1974), analysed the collection and mixing characteristics of flotation columns. Since 1984, however, the number of column flotation studies has dramatically increased. In 1981 a program to study column flotation fundamentals started at McGill University. The initial thrust was development of a scale-up methodology as it was felt that the inability to estimate capacity from pilot unit data had contributed to the reluctance to install columns.

Scale-up of conventional, mechanically agitated flotation machines has proved to be a complex procedure. Design of plant circuits from laboratory data usually relies upon a large safety factor based upon personal experience. (Even scale-up from small to large plant flotation machines is not a straightforward task). However, the effort to scale-up flotation columns should be considerably more fruitful, for the following reasons:

1. There is no mechanical agitation;
2. Close control is maintained over pulp rates and air rate;
3. Bubble size distribution is more uniform;
4. Particle entrainment into the froth is negligible;
5. Steady state laboratory testing is feasible.

Recently, the first International Symposium on Column Flotation was held at Phoenix, Arizona. The event was organized to review and exchange the engineering and technology of column flotation (Sastri, 1988).

2. COLUMN CHARACTERISTICS

A flotation column is shown schematically in Figure 1. Basically, the column consists of two distinct zones, the collection zone (also known as the bubbling zone and slurry or pulp zone) below the interface, and the cleaning zone (also known as the froth zone) above the interface.

In the collection zone, particles from the feed slurry are contacted countercurrent with a bubble swarm produced by a gas sparger at the bottom of the column. Hydrophobic particles collide with and attach to the bubbles and are transported to the cleaning zone. Hidrophilic and less hydrophobic particles are removed from the bottom of the column. In the cleaning zone, water is added near the top of the froth, thus providing a net downward liquid flowrate called a positive bias. The existence of a positive bias prevents the hydraulic entrainment of the fine particles into the concentrate (Dobby and Finch, 1985; Yianatos et al., 1987).

The column has proved particularly attractive for cleaning applications and can achieve in a single stage upgradings comparable to several stages of mechanical cells, often with improved recoveries (Coffin and Miszczak, 1982; Amelunxen and Redfearn, 1985; Wheeler, 1985).

Three principal design features distinguish the column from a mechanical cell: firstly the wash water (added at the top of the froth), secondly the absence of mechanical agitation and thirdly the bubble generation system (gas sparger).

Finally, the existence of a single three-dimensional froth in the column makes the study of the process simpler than that for a bank of mechanical cells, where the froth characteristics change from cell to cell.

Industrial flotation columns are square or circular, typically 0.5-3m in diameter or side, and 12-15m height. Since their invention, square columns have been marketed commercially by the Column Flotation Company of Canada Ltd. The apparent simplicity of the column has led others to copy the original idea and make their own "homemade" columns. Usually, homemade columns are made of common large diameter pipes (ie. Gilbraltar, Inco).

3. COLUMN VARIABLES

3.1 Air Rate (also referred to here as gas rate)

Floatable (hidrophobic) minerals are selectively collected in the recovery zone. The basic mechanism of particle collection consists of collision and attachment of the particles on the interfacial gas (bubble) surface. Subsequently the air is required to transport the floatable solids to the overflow.

Superficial air rate, J_g (cm/s), is commonly use to describe bubble column operations because it can be compared for different column sizes. Superficial air rate is defined as the volumetric air flowrate per unit cross-section. Typical operating conditions are J_g = 1-3 (cm/s). To measure the actual average air rate in columns, reference must be made to standard conditions. A good estimation is obtained using the following relationship:

$$J_g = \frac{P_c \, J_g^{*} \ln (P_t/P_c)}{P_t - P_c} \qquad [1]$$

where J_g^* is the superficial gas rate at standard (atmospheric) conditions, ie., essentially that at the column overflow, P_c is the absolute pressure at the overflow and P_t the absolute pressure at the bottom of the column. For example, for a 10m high column, $P_t \sim 2P_c$, and therefore $J_g = 0.69 \, J_g^*$. A particular case is found in plants located at great heights above sea level, where the atmospheric pressure can be as low as 0.7 atmospheres (eg., La Disputada and Andina in the Chilean Andes at approximately 3,000m above sea level).

The medium of particle transport is the bubble surface. Consequently, the rational approach to estimate the air consumption is to estimate the bubble surface requirement. This can be extended to give a preliminary estimate of column capacity based on particle transport limitations.

Fundamental studies have shown that there is an optimun gas rate to maximise collection rate (Dobby and Finch, 1986a,b), as well to maximise the mineral transport by the gas, namely carrying capacity (Xu et al., 1987). It seems the gas rate is not critical over a wide range (1.5-3.5 cm/s) (Espinosa et al., 1988a,c; Feeley et al., 1987). Industrial experience has shown that excess of gas causes grade recovery losses (Coffin and Miszczak, 1982; Clingan and McGregor, 1987).

3.2 Bubble Size and the Gas Sparger

The bubbles are generated in the flotation columns using two main types of sparger:

* internal porous spargers, perforated rubber or filter cloth
* External generators, e. g. turbo-air, Mott, Deister

The first type includes porous materials made of steel, ceramic or fritted glass with pore size up to 300μm and sieve plates with hole diameter up to 300μm; and rubber or filter cloth spargers, made of different materials, which are the most commonly used in column flotation applications. External generators use special devices, installed outside the column, to generate the bubbles that are carried into the column by means of an additional pulp or water stream (Yianatos et al., 1989b).

Clingan and McGregor (1987) have found that "in general the bubble size generated through an internal flotation column sparger, be it fabric, rubber, or ceramic, is a function of air flowrate for a given surface area of sparger material". This result is in agreement with that reported by Xu and Finch (1989). The other important variable affecting bubble size is the chemical conditioning of the pulp.

Feeley et al.(1987) reported the evaluation of filter cloth, rubber and sintered metal as sparging media for the Inco's Matte separation plant. They found "the filter cloth gave large irregular bubbles while the sintered metal sparger produced the smallest bubbles; however, the latter was susceptible to plugging in an alkaline slurry. Rubber sleeves provided the most satisfactory overall results; they generate an even distribution of fine bubbles (~1-2mm diameter) and do not plug". Similar results have been reported by Wheeler (1985), using rubber spargers.

Fundamental studies (Dobby and Finch, 1986b) have shown there is an optimum bubble size to maximise collection rate (eg., db = 0.4-0.8mm). The reason is that for a certain bubble size there is a maximum gas rate at wich the column can operate normally. The maximum gas rate becomes even smaller at higher downward liquid rates. Similar results have been found regarding the maximum carrying capacity of the bubbles (Xu et al., 1987). Physical and capacity constraints confirm that bubble sizes in the order of 1-2mm are optimal in plant operation. Frother concentration has a significant effect on bubble size. However, no extra consumption seems to be required in plant operations, and in general there have been claims of savings in reagent consumption. Recently, a study on characterization of gas sparging media showed that the gas permeability of the filter cloth is a critical parameter affecting bubble size (Yianatos et al., 1989b).

3.3 Gas Holdup

The gas holdup in the collection zone is an important variable which can affect significantly particle residence time and mineral collection. Gas holdup depends on gas rate, bubble diameter, slurry rate, slurry density and bubble loading, among other variables.

Fundamental studies have shown the large impact the bubble loading can have upon the gas holdup (Yianatos et al., 1988b; Yianatos and Levy, 1989a). Thus, the observed gas holdup, typically about 15%, can possibly increase up to 20-25%, because of the bubble loading. Unfortunately, the gas holdup is difficult to measure in a typical flotation column operation. Thus, as far as is known, no data have been reported on gas holdup measurement under controlled conditions of pulp density and bubble loading. This lack of information prevents a more accurate estimation of the effective volume of collection zone available for particle retention time.

3.4 Wash Water

Wash water increases froth stability, and allows a deep froth bed to develop (Yianatos et al., 1986a). Through the froth bed, a net downward flow of water is maintained which prevents hydraulic entrainment of non-floatable minerals. A rule of thumb to estimate the wash water requirement, J_W (cm/s), is

$$J_w = \frac{J_g^* \varepsilon_c}{1 - \varepsilon_c} + J_b \qquad [2]$$

where

$$\varepsilon_c = \frac{J_c}{J_c + J_g^*} \qquad [3]$$

and represents the fractional holdup of concentrate at the top of the column. In plant operation typical ε_c values are in the range 0.1-0.2. J_b is the superficial bias rate (net downward liquid rate) and J_c the superficial concentrate flowrate.

The effect of the superficial bias rate, J_b, seems not to be significant over a wide range, the minimum being $0 < J_b < 0.1$ cm/s (Clingan and McGregor., 1987; Feeley et al., 1987; Espinosa et al., 1988a). The use of negative bias $J_b < 0$ deteriorates grade, while larger bias rates, $J_b > 0.4$ (cm/s), increases mixing (Yianatos et al., 1987, 1988a) and decreases retention time (Yianatos et al., 1988c; Espinosa et al., 1988a).

It has been experimentally observed (Yianatos et al., 1986a) that a strong increase in superficial wash water rate can drastically change the near plug flow regime of the bubble bed to a more heterogeneous behaviour including severe channeling and recirculation. This effect is directly related to the wash water sparger design and location. The correct practice is to maintain sprinkling water for the whole range of water rates and to avoid jetting.

3.5 Froth Depth

Froth depths in plant operation are around 0.5-1.5m. No general rule has emerged regarding froth depth. It seems the froth depth has little effect over a relatively wide range.

If hydraulic entrainment is the only problem then quite shallow depths may suffice (~0.5m or less), as entrainment appears to be eliminated close to the bubbling zone/froth zone interface when operating at moderate gas rate ($J_g^* < 1.5$ cm/s) (Yianatos et al., 1987). If selectivity between hydrophobic species is required or if high gas rates are used ($J_g^* > 2$ cm/s), deep froths are desirable (Yianatos et al., 1987, 1988a; McKay et al., 1987; Espinosa et al., 1988a).

3.6 Column Height

The collection zone height, H, and the column height to diameter (H/D) ratio have a significant impact on column performance (Clingan and McGregor, 1987; McKay et al., 1987; Yianatos et al., 1988c).

Mineral recovery increases with increasing H/D ratio for a constant column volume and feed flowrate, while the concentrate grade decreases to a minor extent. Columns 10-15m in height with H/D ~10 are well suited for feed rates 150-600 L/min. Maximum H/D is determined by the gas carrying capacity, an important consideration for pilot units where H/D ≈200 (Yianatos et al., 1988c; Espinosa et al., 1988a,b,c).

The total column height, L_C, required for a plant operation can be estimated as,

$$L_C = H + 2 \qquad (4)$$

where L_C includes the required length for collection (H), as well as a provision of 1.5m for froth depth and 0.5m below the gas sparger, to decrease shortcircuiting of fine bubbles into the tailing.

3.7 Feed Solids Percent

High solids percent (30%-50%) can be used in the feed without affecting the concentrate grade, because of the combination of the efficient cleaning action and the positive bias (Mauro and Grundi, 1984; Wheeler., 1985; Feeley et al., 1987). However, limitations of bubble carrying capacity will arise especially in systems operating with fine particles (< 20μm) requiring a high solids recovery into the concentrate, eg., in cleaning operations (Espinosa et al., 1988a,b,c).

3.8 Summary of Typical Design and Operating Conditions

The following values correspond to design and operating conditions typically observed at industrial and pilot scale operations. These values are recommended as a good starting point for new research or production column flotation investigations:

Superficial Gas Rate	: 1-3 cm/s
Superficial Pulp (Slurry) Rate	: 1-2 cm/s

Superficial Wash Water Rate	: 0.3-0.5 cm/s
Superficial Bias Rate	: 0.1-0.2 cm/s
Froth Depth	: $\geq$ 100 cm
Average Bubble Size	: 0.05-0.2 cm
Height/Diameter Ratio	: $\geq$ 10/1

4. PROCESS MODELLING FOR DESIGN AND SCALE-UP

The first attempt to model flotation columns was presented by Sastri and Fuerstenau (1970). They developed a mathematical model based on the assumption of axially dispersed plug flow of both the liquid and air-bubble phases. Also an expression for flotation rate, taking into account the changing concentration of solid particles on the bubble surface, was applied. In this model the fresh water feed to the top of the column was omitted as well as the froth zone. The solution of the equations was only possible for specific limiting cases, and no experimental results were provided.

Dobby and Finch (1986a) developed a flotation column scale-up methodology. The scale-up model utilises kinetic data that are obtained from a series of laboratory column experiments. For modelling purposes the column is considered to consist of two zones: the collection zone, where particle recovery occurs, and the cleaning zone, a packed bubble bed generated by downward flowing wash water. The model explicitly accounts for both the collection zone and cleaning zone recoveries and allows for the effects of bubbles overloading. Results of column scale-up experiments at Gilbraltar Mines are presented. Clingan and McGregor (1987) reported their experience on column flotation design and scale-up at Magma Copper, San Manuel, Arizona. In this project two flotation columns replaced a two stage conventional copper flotation cleaner circuit, and they concluded "the Dobby and Finch column flotation scaling model does provide reasonable sizing parameters for column flotation installations".

From the above experience it is clear that both the mixing conditions and the kinetic process (rate constants) are key factors to be considered in order to model and scale-up flotation columns.

In the remainder of this section some basic aspects of the scale-up methodology developed at McGill University are described (Dobby and Finch, 1986a; Del Villar et al., 1988).

4.1 Kinetic Approach (Rate Constant Measurement)

Firstly an attempt was made to estimate pure collection zone rate constants (Dobby and Finch, 1986a). The approach was to modify the laboratory column by operating at a very high bias, (close to 100%) and moving the wash water addition downward to a point halfway between the column top and the feed entrance. The effect of this was to eliminate the cleaning zone, i.e., there was no packed bubble bed. Short circuit of feed to the concentrate should be prevented then by a sufficiently high bias and no drop back to the recovery zone was expected. Thus, recovery due to flotation in the collection zone was estimated. Total column recovery was evaluated in terms of the estimated collection zone recovery, assuming a value for the cleaning zone recovery. Unfortunately, the use of the "high bias" approach was not enough to prevent entrainment, and at high gas rates recovery by entrainment was higher than obtained in normal column operation. The cleaning zone recovery was unknown and by inspection was fitted to 100%.

An attempt to model the cleaning (froth) zone recovery in a moly upgrading circuit at Gaspé Mines has been reported by Yianatos at al (1988a). Results confirm that floatable minerals show high recoveries (about 90%), but less hydrophobic or depressed minerals are preferentially rejected (e.g., depending on the froth depth), resulting in much lower recoveries (50-60%). The cleaning zone recovery is difficult to evaluate separately, especially at the pilot unit scale. An alternative approach to estimate rate constants from pilot unit experiments has been used by Espinosa et al (1988a). It consists of determining the rate constants based on the total column recovery (including the collection and froth zones). Thus, the column is operated under normal conditions of bias rate and froth depth. This approach gives conservative estimates for the collection zone rate constants. Scale-up of the froth zone is still arbitrary. Recently, Falutsu and Dobby (1989) developed a modified column flotation (pilot unit) that allows the measurement of the recovery of both zones. Their results show that cleaning zone recovery is about 40-50%, and it is supposed that large size columns (2-3m diameter) should have recoveries even lower (10-20%).

4.2 Mixing Characteristics

The mixing in the collection zone is described in terms of the vessel dispersion number, N_D:

$$N_p = \frac{E_p}{(U_L + U_p)\ H} \qquad [5]$$

The dispersion coefficient, E_p, has been related to D, the column diameter, by (Dobby and Finch., 1985; Laplante et al., 1988):

$$E_p = 2.98\ D^{1.31}\ J_g^{0.33}\ \exp\ (-0.025\ S) \qquad [6]$$

Equation (6) was empirically determined from plant operation under the following conditions: column diameter 0.5-1.8m and column height 10-12m. Recently, it has been reported that equation (6) gives a good prediction for columns of 2.5m in diameter (Espinosa et al., 1989).

4.3 Particle Residence Time

Solids residence time governs recovery (Dobby and Finch., 1985; McKay et al., 1987). In a flotation column particle residence time τ_p, increases with decreasing particle size, to approach the liquid residence time. τ_p is a function of the particle settling velocity U_p and interstitial liquid velocity U_L (Dobby and Finch., 1985; Yianatos et al., 1986b):

$$\frac{\tau_p}{\tau_L} = \frac{U_L}{U_L + U_p} \qquad [7]$$

where U_p is the particle settling velocity in a swarm of bubbles and particles. To estimate U_p the equation of Masliyah (1979) for hindered settling in a multi-species particle system can be used, considering bubbles as particulates of zero density (Yianatos et al., 1986b):

$$U_p = \frac{g.d_p^2(1-\varepsilon_g)^{2.7}[\rho_p-\rho_{susp}]}{18\mu[1+ 0.15\ R_e^{0.687}]} \qquad [8]$$

The mean residence time of liquid (slurry), τ, can be estimated as the ratio between the effective volume of recovery zone and the volumetric tailing flowrate, T:

$$\tau_L = \frac{A_c H (1-\varepsilon_g)}{10T} \qquad [9]$$

where A_c is the column cross section, H represents the distance between the air input level and the froth/bubbling zone interface (typically H = 10-12m), and ε_g is the gas holdup. Notice that some workers define H conservatively as the distance between the feed input and air input. Typical values of liquid residence time in plant operations are 10-40 min (Feeley et al., 1987; Yianatos et al., 1987; Espinosa et al., 1989).

Figure 2, from Dobby and Finch (1985), shows the typical trend of the ratio τ_p/τ_L vs U_L with particle size as parameter. As an example, for the usual range of liquid downward velocity U_L = 1-2cm/s, the residence time of 120μm particles (s.g = 4.0) is only 60% of that of the liquid. For further discussion on estimating particle residence time see Dobby and Finch (1985), Yianatos et al. (1986b).

4.4 Recovery Estimation

Assuming flotation is a first order process, recovery can be estimated in terms of the mixing characteristics (N_p), the particle residence time (τ_p), and the rate constant (k) (Dobby and Finch., 1986a):

$$R = \frac{4\,A \exp(0.5/N_p)}{(1+A)^2 \exp(0.5A/N_p) - (1-A)^2 \exp(-0.5A/N_p)} \qquad [10]$$

where

$$A^2 = 1 + 4\,k\,\tau_p\,N_p \qquad [11]$$

4.5 Carrying Capacity Limitation

Maximum loading of the bubbles is a possible limitation in flotation columns, especially in systems operating with fine particles (less than 20μm) requiring a high solids recovery into the concentrate, e.g., in cleaning operations.

The maximum carrying capacity can be determined from pilot unit experiments (Espinosa et al., 1988a), in terms of mass of solids per unit time per unit column crosssectional area (C_A). Alternatively, C_A can be estimated from the following semi-theoretical relationship (Espinosa et al., 1988c):

$$C_a = \frac{60\,\eta\,\pi\,d_p\,\rho_p\,J_g}{d_b} \qquad [12]$$

where η is a combined efficiency factor to account for unknowns such as bubble coverage and solids dropback from the froth, as well as to adjust changes in bubble size. Average values for η of about 0.6 were obtained from pilot unit experiments.

Some practical observations of carrying capacity constraints, from Espinosa et al., (1988c), are as follows: "As operators seek to gain the maximum productivity from a column, the device will be pushed towards its carrying capacity. It is probable then that many columns operate at this maximum production rate. This has some important consequences. A column at its carrying capacity is sensitive to feed grade fluctuations. An increase in feed grade will increase the feed rate of floatable mineral, and since the column is at its carrying capacity this will mean a recovery loss. Operating changes such as increasing gas rate which may work in a mechanical cell will not work in a column (if carrying capacity is increased by gas rate this probably means moving into negative bias and thus loss of grade). The required operating manoeuvre is to change feed solid rate in inverse proportion to the feed grade, to maintain a constant feed rate of mineral, a manoeuvre not readily achieved. As an aside, an increased feed grade requires a decreased solids feed rate which gives a longer retention time. This should not be interpreted as reduced kinetics, however, since the column is not kinetically controlled in this case. This lack of ready accommodation of feed changes through operating variable changes throws more emphasis on the design stage. Probably the best option is series columns, with the downstream column retreating the tailings. The circuit should be designed for the maximum projected mineral feed rate to ensure the circuit's total carrying capacity is not exceeded. The next question is what to do with downstream column concentrates, especially when mineral feed rates drop and concentrate grades too low for final circuits concentrate are produced. There is no point in recycling these to a column already at its carrying capacity. This suggests that column circuits should have at the least 3 stages so that the option exists to recycle the nth column concentrate to the (n-1)th (underloaded) column. Since a column at its carrying capacity is producing final circuit grade concentrate it may be necessary only to monitor the final one or two column stage concentrates and make control decisions as to where to send these concentrates". Experience at Mount Isa Mines confirmed these comments (Espinosa et al., 1989).

4.6 Column Flotation Simulator

A simulator was developed at McGill University, based on the procedure of scale-up and modelling proposed by Dobby and Finch (1986a) with some modifications (Del Villar et al., 1988). This program simulates a circuit of flotation columns, arranged as cleanears or scavengers (Figure 3). A variable number of stages of columns can be simulated. Each stage or unit consists of a feed box and a flotation column. A final box is used to merge all the concentrates in a scavenger circuit or all the tailings in a cleaner circuit to provide an estimate of the overall circuit product. The simulator calculates mineral flowrates to concentrate from the flotation rate constans (input data), and particle retention time and mixing characteristics (calculated). Mineral recoveries and grades for each stream are then calculated. Water recovery to the concentrate is estimated from the gas flowrate assuming a liquid fraction at the lip level; tailings water content is calculated from the feed water and the bias rate. Wash water addition is then calculated by balance. Carrying capacity, C_A, is entered as an experimentally determined value. The simulator "collects" particles up to the C_A value and then stops. If no experimental data are available, equation (12) is used as a first guess. During program execution, mass balance results for the different components and some mixing characteristics are displayed at the end of each iteration to monitor the progress of the simulation.

Complete simulation results (flowrates, recoveries and grades) at any intermediate stage of the simulation can by displayed on the screen or printer. The final printout is organized in two sections. The first contains most of the information used for the simulation (coming from a datafile or from the modifications entered by the user). The second section gives the simulation results. Four types of information are provided: mixing parameters for all column stages, plus recoveries, flowrates and grades for all column stages, species and streams.

The simulator has proved to be very useful to predict circuit performance as well as to make sensitivity analyses upon those critical parameters affecting column flotation performance (i.e., rate constants, feed solids percent, bias rate, holdup, recycles, etc.)

5 BIAS AND CONTROL

The cleaning action (rejection of hydraulically entrained particles) is based on the existence of a net downward flowrate in the froth zone. This is achieved by having a tailings volumetric flowrate, T, greater than that of the feed, F. One way to control this is to maintain a constant difference or bias, B, i.e.,

$$B = T - F > 0 \qquad [13]$$

A second way is to maintain a ratio of tailings to feed volumetric flowrate constant at some value greater than one. This is called a "bias ratio", BR, i.e.,

$$BR = \frac{T}{F} > 1 \qquad [14]$$

BR values from 1.01 to 1.15 are typically recommended (Dobby et al., 1985; Clingan and McGregor, 1987; Feeley et al., 1987). However, plant operation sometimes shows a wider range (BR = 1.01-1.50). This spread in BR can be explained in terms of the column flotation control system.

Figure 4 shows two different control modes used in plant practice. Alternative (a) corresponds to the original column (Coffin and Miszczak, 1982), where the volumetric difference B is maintained, while an independent loop for level control manipulates the wash water. The same control strategy can be used with a set on bias ratio BR (Amelunxen and Redfearn, 1985; Clingan and McGregor, 1987; Feeley et al., 1987). Alternative (b) shows a different way to operate, by fixing the wash water flowrate and using a level control loop linked to the tailing flowrate (Mauro and Grundi., 1984; Feeley et al., 1987; Espinosa et al., 1989).

Both schemes accomplish the objetive of positive bias. However, alternative (a), operating with a constant difference B, has the advantage that feed perturbations do not directly affect the wash water flowrate, and thus more constant froth characteristics are maintained. In alternative (a), with a set on bias ratio, a change in feed volumetric rate will cause a consequent change in wash water rate, and thus the system operates at variable bias. This corresponds to an interactive (coupled) control system which can be more difficult to tune. Case (b) remains an alternative when feed flowrate cannot be measured easily (Lornex experience) or when on-stream analysis (OSA) is available to adjust the proper wash water addition (experiences at Inco and Mount Isa Mines).

Feeley et al. (1987) reported the experience at Inco's Matte separation plant, where they found that similar unit recoveries were achieved using alternatives (a) and (b). However, alternative (b) using OSA was simpler and involved less instrumentation; also it required lower air and wash water rates. This condition, of course, is much closer to the optimal operation (minimum net downward flowrate in the froth).

At the laboratory/pilot scale, the use of a constant bias or wash water rate is recommended. This practice is particularly useful in situations where low feed flowrates are necessary due to residence time limitations. In these cases the wash water rate is determinad by the requirements of constructing an adequate froth zone rather than being related to the feed rate. Consequently, bias ratios will often be high (e.g., BR=2), but this will not necessarily reflect the BR to be expected at the plant scale.

All the control schemes presently in use are limited because no direct measurement of bias flowrate (in the froth) is available. The inaccuracy of bias prediction from feed and tailing flowrate prevents the application of optimal control polices, i.e. minimum bias. It must be noticed that both equations (13) and (14) are quite conservative, because the volume of floated minerals is not considered (Finch and Dobby, 1989). In other words, an excess of wash water over that required to generate a positive downward flowrate in the froth will always be obtained, thus increasing mixing in the froth and decreasing collection zone residence time. Excess water also costs more, and the dilution of tailings downstream may also be underisable.

A method of measurement of the net downward flowrate in the froth has been devised (Moys and Finch, 1988), which employs a principle involving the measurement of the temperature profile.

6 INDUSTRIAL APPLICATIONS AND TESTING

Industrial use of column flotation is becoming widespread around the world. Some typical applications in occidental countries are listed below. The asterisk(*) means column testing reported.

6.1 Summary of Plant and Pilot Operations

AUSTRALIA

AMDEL, Adelaide	* (several)
BHP Central Research Laboratory	* (coal)
Blue Spec/Golden Spec, W.A.	(gold)
Harbour Lights, W.A.	(gold)
Hellyer, Tasmania	(lead, zinc, silver)
Kambalda Nickel Operations, W.A.	(nickel)
Mount Isa Mines, Queensland	(copper, lead, zinc)
Paddington, W.A.	(gold)
Renison Ltd., Tasmania	* (tin)
Riverside Coal Preparation Plant	(coal)
Telfer, W.A.	(copper)
Woodcutters, N.T.	(zinc)

CANADA

Geco Mines, Ontario	(copper, lead)
Gibraltar Mines, B.C.	(copper)
Inco, Ontario	(copper)
Inco, Thompson	(copper)
Lornex Mining Co., B.C.	(copper, moly)
Mines Gaspé, Québec	(moly)
Cominco, Polaris	(lead, zinc)
Niobec, Québec	*(carbonates, niobium)
Noranda, New Brunswick	* (copper, moly)
Noranda, Mattabi, Ontario	* (copper, lead)

CHILE

Cía. Minera del Pacífico	* (Phosphates)
Codelco, Chuquicamata	(copper,moly)
Codelco, Andina	* (copper, moly)
Codelco, El Teniente	* (copper)
Exxon, Disputada, Los Bronces	(copper)
Exxon, Disputada, El Soldado	* (copper)

La Escondida	* (copper)
Mantos Blancos, Antofagasta	(copper)
Soquimich	* (astrakanite)

PAPUA NEW GUINEA

Bougainville Copper, Panguna	(copper)

PERU

Cuajone, Tacna	(copper)

U.S.A.

Cominco, Alaska (projet Red Dog)	* (lead, zinc)
Cyprus Minerals, Sierrita, Arizona	(copper, moly)
Magma Copper Co., Pinto Valley	(moly)
Magma Copper Co., San Manuel, Arizona	(copper, moly)
Kennecott Copper Co., New Mexico	(moly)
U.S. Bureau of Mines, Salt Lake	* (fluorite, chromite)

6.2 New Projects and Expansions

Several plants are testing columns for expansion or new applications, e.g. Chuquicamata, Andina, Disputada, El Teniente, etc.

New projects in development have also considered the use of columns, e.g. La Escondida project (Chile) and Red Dog project, Alaska, (U.S.A.). La Escondida is a large high grade copper deposit in Chile. Huggins et al (1987) reported a comparative pilot plant test with 35 tons of ore. They found that one column can replace three stages of cleaning and result in a signicantly higher concentrate grade with no loss in recovery.

The Red Dog project in Alaska (Giegerich, 1986) is the second largest zinc deposit ever discovered. It is expected that at full production Red Dog will be the largest zinc mine and of the lowest cost producers in the western world.

7 GENERAL REMARKS

a) Cleaner/Scavenger vs. Rougher: Column Application.

Column flotation has proven to be successful in most cleaning and scavenger applications. However, references to rougher uses are scarce. Wheeler (1985) reported a successful application of columns on a rougher/scavenger stage in a cooper ore in Peru. Mauro and Grundi (1984) described the unsuccessful experience at Lornex Mining Corp. Ltd., when columns were tested in a Cu/Mo bulk rougher stage. They found that a froth column could not be established that was stable enough to allow washing in any of the test runs. Without washing, the column was not able to attain the grade. Also, it was found that there was no recovery at all below 65 mesh and hardly any recovery below 100 mesh. In general, difficulties in rougher applications seem to be related to froth stabilization, which implies low recoveries, and then a potential increase in circulating load. Similar reasons can explain the failure in other unreported attempts to use columns in rougher stages.

Lane and Dunne (1987) reported the succes of a particular application of columns in a rougher/scavenger circuit at Harbour Lights (Australia). The column circuit separates iron and arsenic minerals, and this simplifies the following cyanidation of gold.

b) Use of Pilot Units in Plant Testing.

The pilot unit typically used in a first step of plant testing is a 50mm diameter by 10m height column (Espinosa et al., 1988a,c; Del Villar, 1989). Experience at Inco's Matte separation plant showed that the 50mm column testwork gave a good indication of the grade/recovery relationship to be expected with the larger columns despite the 1000:1 volume scale-up. A second level in plant testing is to use columns of 0.5-1.0m in diameter by 10-12m height (i.e., Gilbraltar, Inco, Mount Isa Mines, Chuquicamata, Andina, Disputada).

c) Use of Baffles in Large Size Columns.

Large size columns (diameter greater than 1m) should be baffled because of the increased in mixing and then in shortcircuiting. Wheeler (1985) uses baffles in the 1.8 x 1.8 x 14m square column. He also cites structural reasons to justify the use of baffles. At Gilbraltar Mines, homemade columns of 2.1m in diameter were baffled to decrease mixing. Inco installed a 1.8m column without baffles and they were able to get similar results to the 1.1m column. However, they have claimed (Feeley et al., 1987) that the apparent requirement in the larger column for increased retention time to achieve the same recovery can be attributed to a change in feed quality as well as due to short circuiting (increased mixing). Measurement of the mixing charactericts has been recently done at Inco, and clearly shows an increase in mixing if compared with performance of 1m diameter columns (Laplante et al, 1988). Similar results have been observed at Mount Isa Mines in columns of 2.5m diameter (Espinosa et al., 1989).

d) Considerations on Gas Spargers.

It seems there is no room to improve flotation column performance by decreasing bubble size bellow 0.5-2mm, normally used (Dobby and Finch, 1986b; Xu et al., 1987; Yianatos and Levy, 1989a). The use of smaller bubble sizes can cause troubles in operation and decreases capacity. Consequently, research on gas spargers should mainly be addressed to finding suitable designs and materials for each application, regarding lifetime and maintenance. As far as is known, the only rule that appears to be widely accepted for scaling up gas spargers is to maintain the same superficial gas rate (J_g) per unit area of sparger (Xu and Finch, 1989).

e) Advantages of Flotation Columns vs. Conventional Mechanical Cells.

The following list summarizes the main advantages that have contributed to justifying the use of flotation columns in plant conditions:

* Improved metallurgical performance
* Operating Costs: lower power consumption
  lower reagent consumption
  saving in nitrogen if used instead of air
* Control : easy and stable operation
* Space : less space is required
* Simpler in design and contruction: savings in capital investment
* Ability to work with very fine particles at high solids percent while maintaining good cleaning action
* Potential for decreasing circulating load.

e) Disadvantages: Wash water costs, tailings dilution, gas sparger maintenance, and requirements of space in height.

8 FUTURE TRENDS IN RESEARCH

Most of the research work in column flotation has been addressed to a better understanding of the variables affecting the process, in order to improve its operation and control. Special attention has also been given to the development of a process model to predict metallurgical performance for design and scale up purposes. Nevertheless, much remains to be done regarding optimality in design, operation and control.

a) **Modelling**

Collection Mixing (general correlation for dispersion coefficient)
Rate constants (standard method of measurement).
Gas holdup (measurement and modelling)
Particle residence time (testing and validation of correlations).

Froth Zone: Selectivity (impact of particle size and liberation)
Recovery (measurement and modelling, effect of mineral drop back and column size).
Carrying capacity (measurement and modelling, effect of column size).

b) **Design**

Gas Sparger Selection, design and scale-up (methodology for testing and evaluation)
Wash Water Distributor: Design and location (methodology for evaluation)

c) **Control**

Bias: minimun water addition by direct measurement or OSA analysis
Air: as a function of the amount of floatable minerals in the feed.

d) **Circuit Configuration**: via simulator

Column arrangements (series or parallel, cleaner or scavenger).
Recycle streams
Integration of columns with conventional cells

NOMENCLATURE

A_c	column cross-section, cm^2
B	bias flowrate, L/min
BR	bias ratio
CA	carring capacity, $g/min/cm^2$
D	column diameter, m
d_b	bubbles diameter, cm
d_p	particle diameter, cm
E_p	dispersion coefficient of particles, m^2/s
F	Feed flowrate, L/min
g	gravity acceleration, cm/s^2
H	height of the bubbling zone, m
J_b	superficial bias rate (net downward rate), cm/s
J_c	superficial concentrate flowrate, cm/s
J_g	superficial gas rate, cm/s
J_g^*	superficial gas rate at standard conditions, cm/s
J_w	superficial wash water rate, cm/s
k	rate constant, min^{-1}
L_c	total column height, m
P_c	absolute pressure at the overflow (top) level, kPa
P_t	absolute pressure at the bottom level, kPa
R	mineral recovery
R_e	Reynolds number
S	feed solids percent
T	tailings flowrate, L/min
U_L	interstitial liquid (slurry) velocity, cm/s
U_p	particle settling velocity, cm/s

Greek Symbols

ε_c	fractional concentrate holdup at the top of the column
ε_g	fractional gas holdup
μ	liquid viscosity, g/cm/s
η	efficiency factor in equation [12]
π	number pi
ρ_p	particle density, g/cm^3
ρ_{sus}	suspension density, g/cm^3
τ_L	mean residence time of liquid (slurry), min
τ_p	mean particle residence time, min

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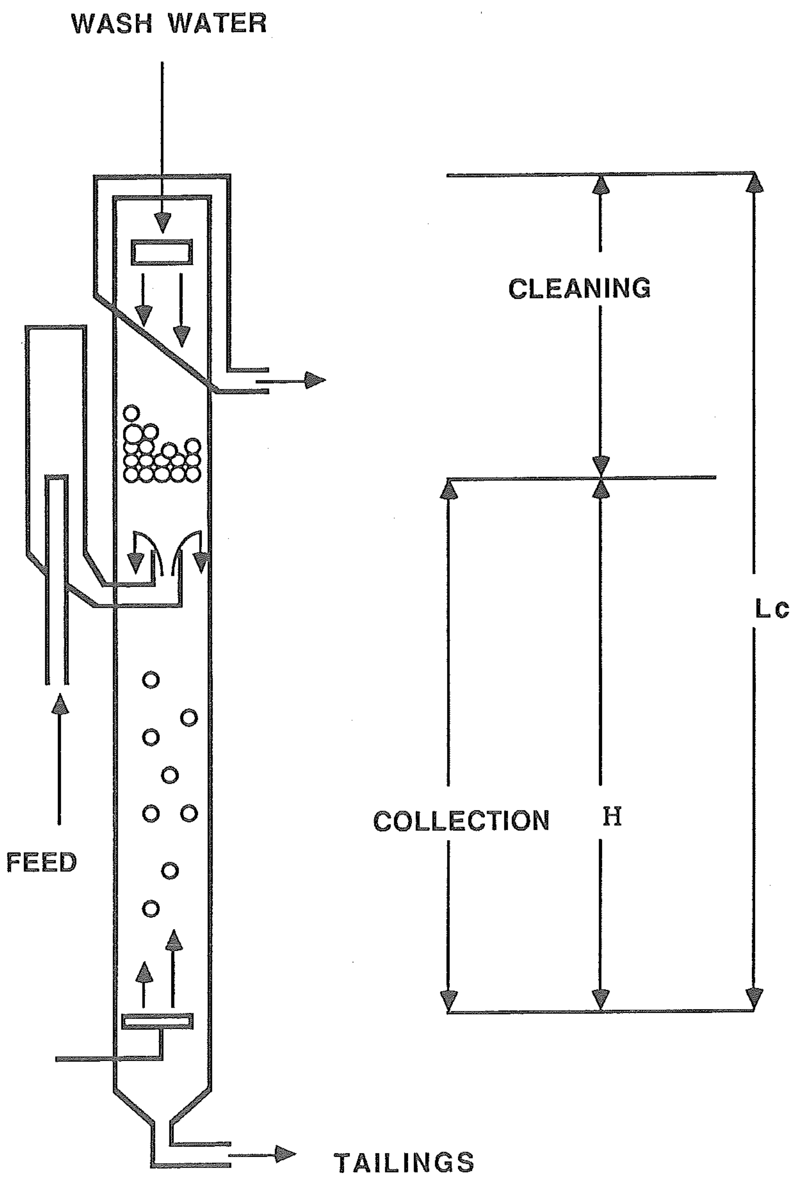

FIGURE 1 Flotation Column.

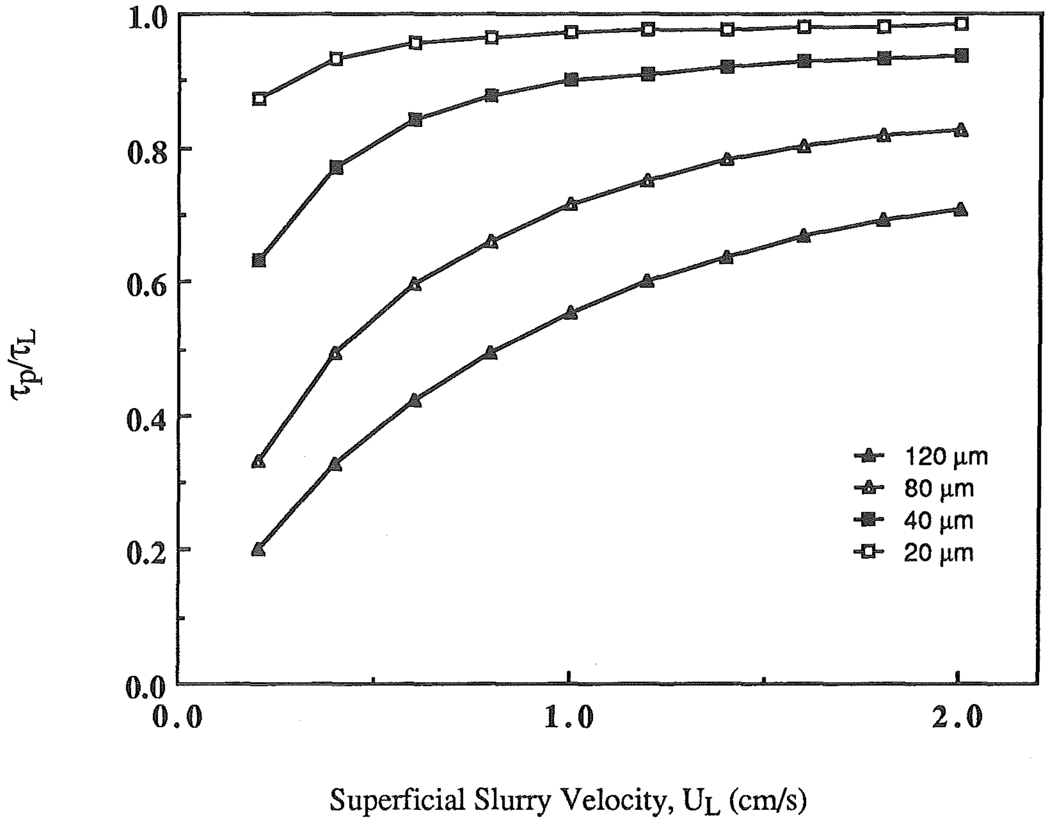

Figure 2 Effect of Paticle Size on Residence Time
(From Dobby and Finch, 1985)

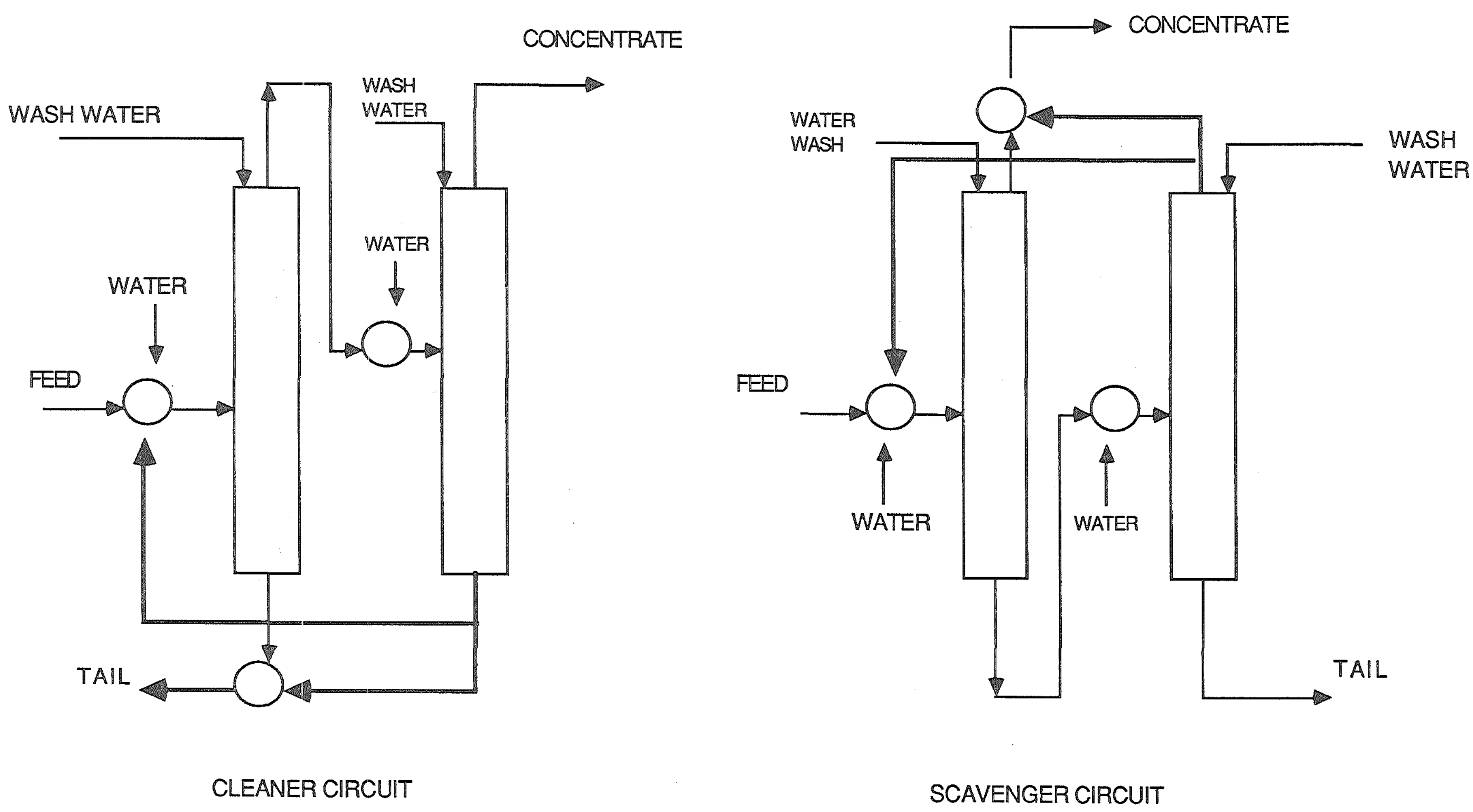

FIGURE 3 CIRCUIT SIMULATION

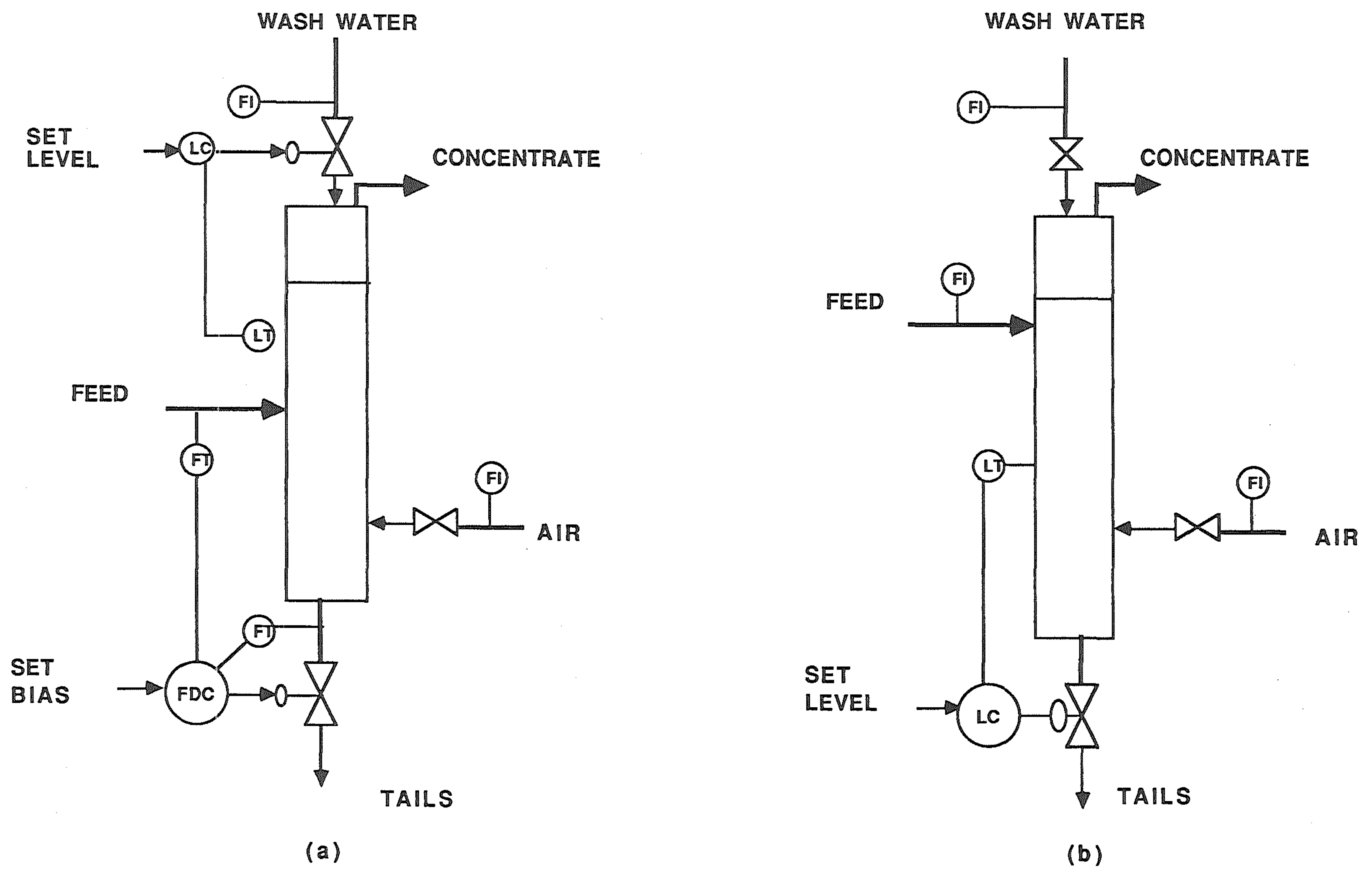

Figure 4 Control Schemes in Plant Flotation Columns.