A BENCHMARKING TOOL FOR ASSESSING FLOTATION CELL PERFORMANCE

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ABSTRACT

Bubbles are unquestionably the heart of the flotation process. Their size and combined surface area are largely what drive the recovery process. This suggests that plant engineers should be aware of where their plants are operating with respect to what is achievable for these key parameters. The paper presents a model of how the key operating variables of frother type and concentration, gas rate, altitude (above sea level) and viscosity affect the Sauter mean bubble size ($D_{32}$) and bubble surface area flux ($S_b$). The results are compared to a set of plant measurements that demonstrate the benchmarking capability of the approach. A case study is used to link hydrodynamic change to improved metallurgical performance.

KEYWORDS

Gas Dispersion, Flotation, Hydrodynamics, Bubble Size, Modeling
INTRODUCTION

At the beginning of the 20th Century the early practitioners of flotation were aware of the importance of understanding bubble behaviour (Rickard, 1916) but measurement of individual bubble and gas behaviour in the industrial environment was not possible due to a lack of adequate sensors. This situation was not rectified until three quarters of a century later with the advent of industrially-robust sensors developed, notably, at the University of Queensland’s Julius Kruttschnitt Mineral Research Centre (JKMRC) (Schwarz & Alexander, 2006) and McGill University (Gomez & Finch, 2002). What emerged was not the chaotic environment perhaps expected, but defined relationships between gas rate, gas holdup, bubble size and the total surface area of bubbles. These relationships could be defined by mathematical expressions and predicted from an understanding of the key process variables (Gorain, Franzidis & Manlapig, 1997; 1999) (Hernandez-Aguilar, Gomez & Finch, 2002) (Gomez, Cortes-Lopez & Finch, 2003) (Nesset, Gomez & Finch, 2007). The early JKMRC work focused on a large pilot cell (3 m$^3$) installed in plant environments, while the McGill work was initiated on flotation columns (Finch & Dobby, 1990) (Xu, Finch & Uribe-Sales, 1991) and later adapted to industrial mechanical cells (Cooper, Scott, Dahlke, Finch & Gomez, 2004) (Dahlke, Gomez & Finch, 2005) (Nesset, Gomez, Finch, Hernandez-Aguilar & Difeo, 2005). Yianatos of Santa Maria University (Chile), a graduate of the McGill research team, carried the work to industrial settings in South and Central America (Yianatos, Bergh, Condori & Aguilera, 2001). The methods of gas dispersion measurement are now well accepted and most major research centres and flotation cell manufacturers have adopted the technology (Pyecha, Sim, Lacouture, Hope & Stradling, 2006) (Yañez et al., 2009) (Collins, Schwarz & Alexander, 2009).

The drivers behind the desire to better characterise and understand the industrial flotation environment are undoubtedly economic in origin; lower-grade deposits and higher tonnage operations put larger capital expenditures at risk; tightening profit margins demand continuous improvement in process performance. One measure of this change in expectation is, arguably, the increase in mechanical flotation cell size over time as illustrated in Figure 1. Understanding the processing, as well as mechanical, scale-up criteria is essential in order to avoid disaster. The gas dispersion measures have, over the past 15 years, been added to the traditional scale-up parameters thereby enhancing the tool kit available to manufacturers, operators and researchers. A brief review of the traditional criteria and the gas dispersion parameters follows.

Figure 1 - Increase in maximum flotation cell size (mechanical cells) over the past 100 years (adapted from Lynch et al., 2007)
HYDRODYNAMIC SCALE-UP PARAMETERS

The Classic Parameters

These parameters are essentially those of fluid mixing and gas-liquid reactor design in chemical engineering (Oldshue, 1983). Their applicability to flotation has been reviewed by Arbiter, Harris and co-workers (Arbiter & Steininger, 1965; Arbiter, Harris & Yap, 1976; Harris, 1976; Arbiter, 1999) and more recently by Deglon, Egya-Mensah and Franzidis (2000). The most relevant of these are:

Power intensity, $P_I$

The net power per unit volume of the cell. Typical ranges are 1-3 kW/m$^3$ but can have a range of 0.8 to 9.5 (Deglon et al., 2000). Values tend to decrease with increasing cell volume. A more important variant for larger cells is the power intensity of the highly-turbulent near-impeller region, and manufacturers typically keep these values constant as a scale-up criterion for large cells;

$$P_I = \frac{P_{\text{net}}}{V}$$

(1)

Impeller tip speed, $S$

Tip speed is considered important since blade velocity affects pulp circulation rate, fluid/gas shear and therefore gas dispersion. Typical values are 5-7 m/s but values as high as 9 m/s have been reported (Deglon et al., 2000). Note that consumed power will increase as $N^3$ so motor size and cost increase rapidly with increasing $S$. Manufacturers still consider tip speed an important scale-up criterion and keep it relatively constant with increasing impeller size, depending on duty (coarse/fines, roughing/cleaning). Relatively constant tip speed (above a critical minimum value of around 4 m/s) maintains air dispersion capability and keeps the volume of pulp pumped per unit of energy the same as cell size is increased.

$$S = \pi ND$$

(2)

Air flow velocity, $U_Q$

Air flow velocity, $U_Q$ (Equation 3), relates the ability of an impeller of diameter $D$, to accommodate a volumetric flow rate of gas, $Q$. Industry values vary greatly, from 0.02 – 0.5 m/s, so like $N_Q$, it is more of a guide in terms of a maximum operating value, than a scale-up criterion. It links to the superficial gas velocity ($J_g = Q/A_{\text{cell}}$) since there is typically a close relationship between impeller diameter ($D$) and cell size ($A_{\text{cell}}$).

Other measures such as the Power number, $N_P$ (Equation 4), (ratio between actual (net) power and the theoretical power), Air Flow number, $N_Q$ (Equation 5), (ratio between the air flow rate and the theoretical impeller pumping rate), Froude number, $N_F$ (Equation 6), (ratio of inertial (or centrifugal) to gravitational force), and Weber number, $N_{We}$ (Equation 7), (ratio of inertial (or centrifugal) to surface tension force), serve more as guides than fast scale-up criteria and will not be covered here. The impeller mixing Reynolds number, $N_{Re}$ (Equation 8), is a measure of inertial to viscous forces and will increase in proportion to the rotor diameter if tip speed remains constant. $N_{Re}$ for flotation cell impeller design is always well into the turbulent region ($>>10^3$) thus $N_P$ remains constant for a particular rotor design, and so $N_{Re}$ does not impact directly as a scale-up criterion.
These were the initial hydrodynamic parameters used for flotation cell scale-up. Parameters linking the ability of a cell to effectively remove the recovered solids, such as tonnes/m$^2$ of cross-sectional area, and tonnes/m of launder lip length, are also used for scale-up but are not considered to be hydrodynamic parameters directly.

**Gas Dispersion Parameters**

One of the obvious shortcomings of the traditional hydrodynamic parameters is that bubble size is not accounted for in spite of its key role. Harris (1976), in his classic work on scale-up parameters, comments on the critical importance of bubble size in driving recovery, and the strongest influence on bubble size being frother addition. Figures 2 and 3 illustrate the influence on bubble size of increasing frother addition of DF-250 (DowFroth 250) (Nesset, Finch & Gomez, 2007). Note the bimodal bubble size distribution (BSD) without frother and trend toward a narrower BSD and smaller bubble size with increasing frother addition, ultimately reaching a limiting bubble size. The concentration at which this limiting size is reached has been termed the *Critical Coalescence Concentration* or CCC by Laskowski (2003).
Figure 2 - Images of bubble size as frother concentration (DF-250) is increased for a mechanical forced-air machine. Images correspond to BSD’s in Figure 3

Figure 3 - Changes in bubble size distribution (inset frequency distributions) as frother concentration (DF-250) is increased in a mechanical forced-air machine (Nesset et al., 2007)

The importance of bubble size and the total surface area of gas in driving particle, and hence flotation, recovery had been established some time ago (Pogorely, 1962; King, 1972, 1973; Laplante, Toguri and Smith, 1983) but it was not until the work in the late 1980’s and 1990’s of Finch and co-workers at McGill on columns, and Gorain and co-workers at the JKMRC on a pilot-scale mechanical cell, that the industrial link between the gas dispersion variables and flotation kinetics was firmly established.

These groups developed the techniques and robust sensors for measuring what have become known as the gas dispersion parameters; Superficial gas velocity \((J_g)\) representing the volumetric gas flow per unit of time \((Q_g)\) exiting the pulp per unit cross-sectional area of the cell \((A_{cell})\); the Sauter mean bubble size \((D_{32})\) representing the BSD (of measured individual size \(d_i\)) for use in calculating the bubble surface
area flux \((S_b)\), the total surface area of bubbles exiting the cell per unit cross-sectional area of the cell per unit of time. These parameters are linked by the following relationships:

\[ J_g = \frac{Q_g}{A_{cell}} \]  
\[ D_{32} = \frac{\sum d_i^3}{\sum d_i^2} \]  
\[ S_b = \frac{6J_g}{D_{32}} \]

A third measured variable (given that \(S_b\) is calculated from two others), the volumetric gas hold-up \(\varepsilon_g\), expressed as a fraction or %, also forms one of the gas dispersion metrics and relates the volume of gas \(V_g\) to the volume of the cell \(V_{cell}\):

\[ \varepsilon_g = \frac{V_g}{V_{cell}} \]

Gorain and co-workers (1997, 1999) showed that bubble surface area flux \((S_b)\) is a key driver of flotation recovery through the \(k-S_b\) relationship:

\[ k_{pulp} = PS_b \]  
\[ k = PS_bR_f \]

where \(k_{pulp}\) represents the rate constant in the pulp phase, \(k\) represents the overall flotation rate constant (pulp and froth), \(P\) is a floatability parameter for the mineral particles to be floated, and \(R_f\) is recovery of floated particles across the froth phase. Observing that \(S_b\) is inversely proportional to \(D_{32}\) it is seen that the rate of flotation, and hence recovery, is inversely linked to bubble size and directly proportional to the volumetric flow of gas, \(Q_g\). Studies of flotation performances require, therefore, accurate and precise measurements of both the \(J_g\) and \(D_{32}\) parameters in order to calculate \(S_b\). These needs were fulfilled by the developments in sensor technology and process measurement methodology of McGill and the JKMRC. Interestingly, the earlier work of Laplante et al. (1983) predicted the same relationship as Equation 14 for what they termed the “free-floating mode”, that is, the condition where particles are free to collect on the bubble surface as opposed to their described “hindered floating mode” where the bubble surface becomes loaded with collected particles and the direct link between \(k\) and \(S_b\) is no longer valid. The majority of base metal rougher and scavenger flotation can be considered to be in the free-floating mode.

**Gas Dispersion Measurements**

Measurement of \(J_g\), \(D_{32}\) and \(\varepsilon_g\) utilizing the McGill gas dispersion sensors has been well documented (Gomez & Finch, 2002, 2007; Nesset et al., 2005, 2006) and will not be covered in detail here. Each sensor relies on a different physical parameter of the system: relative conductivity of aerated and non-aerated pulp for determining gas holdup, pressure change (in a tube) due to volumetric displacement of pulp by gas for establishing superficial gas velocity, and digital imaging of captured bubbles in an exterior chamber for determination of the BSD and calculation of \(D_{32}\). Pulp density changes are also typically monitored using a pressure sensing tube to permit corrections to the \(J_g\) measurement due to cell level/density changes. Data are logged to computer files for subsequent analysis. The sensors are well-
suited for plant testing and have been used world-wide in dozens of campaigns over the past decade by the McGill team and other groups (Harbort & Schwarz, 2010). Figure 4 illustrates a typical sensor grouping in an industrial cell showing cell-characterisation measurements being taken as close together in the cell as practical.

![Figure 4 - Example of multiple sensors performing a single cell characterisation](image)

**DEVELOPMENT OF A D_{32} MODEL**

Somewhat surprisingly, given its importance in flotation, there existed no mathematical relationship (model) linking bubble size created in mechanical flotation machines to the key process variables of gas rate, frother concentration (and type), energy (impeller) input, fluid viscosity and altitude (above sea level). A program of investigation to develop such a model was undertaken at McGill utilizing an available 0.8 m$^3$ Metso RCS™ tank cell that offered variable impeller speed and was sufficiently large to permit distinct separation of turbulent and quiescent zones, enhanced by the presence of an internal “shelf-baffle” as illustrated in Figure 5. The cell was sufficiently large to permit a full suite of gas dispersion sensors to be installed. It needs to be noted that all testing was performed in the two-phase, air-water system without the presence of solids. The details of the approach have been previously presented (Nesset et al., 2007) and this paper serves to update the reporting of results and to present the final mathematical relationships and model. Note that the diameters of cell and rotor are 1.15m and 0.21m respectively.
Variables Investigated

Five variables were selected for investigation, being key process parameters that are either manipulatable (frother type and concentration, gas rate, impeller speed), or determined by virtue of plant location or slurry feed (gas density, liquid viscosity). BSD’s were determined by imaging at least 10,000 bubbles using commercial software (Northern Eclipse™, Empix Imaging) and calculating $D_{32}$ using Equation 10. Superficial gas velocity was determined by averaging 15 measurements. Measurement precision of the reported mean values at the 95% confidence level based on repeat testing was 3% for $D_{32}$ with frother and 5.5% without frother, and 4.9% and 7.2% for $J_g$ at 0.5 and 1 cm/s respectively.

Frother Concentration and Type

Table 1 lists the five frothers used, their chemical descriptions, manufacturer and molecular weights. All are commercial frothers with the exception of Pentanol which is a component of some commercial frothers. They were selected to cover a broad spectrum in type and molecular weight. Note that two can be considered alcohols and three as polyglycols. Composition of commercial frothers can vary, particularly blended types such as F-140, which is a mixture of aldehydes and ketones.
Table 1 - Frothers used for testing in the Metso RCS™ 0.8 m³ cell

<table>
<thead>
<tr>
<th>Frother</th>
<th>Type</th>
<th>Supplier</th>
<th>Molecular Weight</th>
</tr>
</thead>
<tbody>
<tr>
<td>1-Pentanol</td>
<td>Simple alcohol</td>
<td>Fisher</td>
<td>88</td>
</tr>
<tr>
<td>MIBC</td>
<td>Methyl isobutyl carbinol</td>
<td>Dow</td>
<td>102</td>
</tr>
<tr>
<td>DF-250</td>
<td>Polypropylene glycol alkyl ether</td>
<td>Dow</td>
<td>235 – 265</td>
</tr>
<tr>
<td>F-140</td>
<td>Mixture aldehydes and ketones</td>
<td>Flotttec</td>
<td>Mixture of C8 – C22, Typical 200-250</td>
</tr>
<tr>
<td>F-150</td>
<td>Polypropylene glycol</td>
<td>Flotttec</td>
<td>410 - 440</td>
</tr>
</tbody>
</table>

Characterisation was by adding frother to the 700 litres of water in the Metso test cell and measuring the BSD at \( J_g \) 0.5 and 1 cm/s. A sufficiently wide range in frother concentration was covered and data analysed by fitting an exponential decay function (as suggested by Comley, Harris, Bradshaw & Harris, 2002) of the type illustrated in Figure 6. CCC95 values were determined for each frother type at the two \( J_g \) values and are listed in Table 2. Repeat testing in 2010 of the Pentanol with extended concentrations provided a better CCC95 value. The CCC95 concept has been previously introduced (Nesset et al., 2007) and can be extended to define the CCCX concept, where X represents the % reduction in A as described in Figure 6. Of note is that CCC95 values increase somewhat with increasing \( J_g \). This will be further discussed under Interaction Effects.

Plotting \( D_{32} \) values against normalized froth concentration, defined as ppm/CCC95, as illustrated in Figure 7 (upper plot), reduces all data to a single curve suggesting that any frother can be uniquely characterised by its CCC95 value. The expanded view in the lower plot in Figure 7 does show that the curves for the two \( J_g \) data sets are slightly different. As noted, these interaction effects will be covered later, however as a first approximation, Equation 9 serves as a good model having a 95% confidence interval of ±0.24 mm compared to ±0.12 mm and ±0.22 mm for the \( J_g = 0.5 \) and 1 cm/s data respectively.

\[
D_{32} = 0.874 + 2.26 \cdot \exp \left[ -3.09 \cdot \frac{ppm}{CCC95} \right]
\]  

\[
D_{32} = D_{limiting} + A \cdot \exp[-B \cdot ppm]
\]

Figure 6 - Illustration showing the exponential model used to fit the \( D_{32} \) versus ppm frother data. The CCCX lines indicate the ppm levels for X% reduction in A. B represents the decay constant in the exponential term and \( D_{limiting} \) is the limiting bubble size for infinite frother addition. CCC95 correlates well with Laskowski’s CCC values.
Table 2 - CCC95 values (ppm) for the five frothers determined for J\(_g\) of 0.5 and 1 cm/s. The 2010 value for Pentanol is considered a better value

<table>
<thead>
<tr>
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</tr>
</thead>
<tbody>
<tr>
<td>J(_g) = 0.5 cm/s</td>
<td>30.5</td>
<td>23.7</td>
<td>12.4</td>
<td>10.1</td>
<td>9.0</td>
<td>4.2</td>
</tr>
<tr>
<td>J(_g) = 1 cm/s</td>
<td>34.8</td>
<td>-</td>
<td>14.0</td>
<td>16.8</td>
<td>13.0</td>
<td>6.0</td>
</tr>
</tbody>
</table>

Figure 7 - The normalized plots of D\(_{32}\) versus ppm/CCC95 for four frothers (not including Pentanol). Note the bottom plot is an expanded view of the 0-2 ppm/CCC95 range and that the model fits are slightly different for J\(_g\) = 0.5 and 1 cm/s

Superficial Gas Velocity, J\(_g\)

The effect of J\(_g\) on D\(_{32}\) was explored using MIBC and DF-250 frothers as these two are arguably, standards for the industry with “equivalent” types available from various manufacturers. J\(_g\) was varied across a wide range, 0.05 to 1.5 cm/s, at five different frother concentrations, including zero frother addition. The results for MIBC are shown in Figure 8 and results for DF-250 were very similar (not shown). Building on work reported earlier (Nesset et al., 2006, 2007) the data at each frother concentration were fitted to models of the form;

\[ D_{32} = D_0 + a \cdot \left(100 \cdot J_g\right)^{0.5} \] (16)
where $D_n$ represents the notion of a finite “creation size” at zero gas rate, and $a$ is a constant of the system that depends on the frother concentration and the bubble generation method. The square root relationship between $D_{32}$ and $J_g$ implies that as $J_g$ is increased both the size and number of bubbles also increases.

![Figure 8: Curves for MIBC showing the fit of $D_{32}$ versus $J_g$ data to Equation 10 at various frother concentrations. The plots for DF-250 were similar](image)

**Viscosity**

The (dynamic) viscosity of the flotation pulp is affected by different factors including solids content, particle size and shape, chemical content and temperature of the liquid. This study was limited to investigating the effect of water viscosity due to temperature on $D_{32}$ since testing was in the two-phase air-water system. The use of sucrose and polyacrylamide solutions to introduce viscosity changes was attempted but proved unsuitable (frothing and shear effects) so the temperature of the water itself was varied as a method for introducing significant viscosity change. The viscosity-temperature effect is illustrated in Figure 9 (left plot) and shows that dynamic viscosity can change by a factor of three over the range 4 to 40 °C, a not uncommon range in flotation circuits. Winter testing (March) using cold to hot Montreal tap water (700 litres) in the Metso cell permitted testing of this range and results for four tests are shown in Figure 9 (right plot). The data were fitted to a power relationship for the ratio of dynamic viscosity ($\mu$) to dynamic viscosity at 20 °C ($\mu_{20}$) producing an exponent of 0.776. The viscosity correction factor ($f_v$) for $D_{32}$ can therefore be defined as:

$$f_v = \left(\frac{\mu}{\mu_{20}}\right)^{0.776}$$  \hspace{1cm} (17)
Altitude

Many mines in mountainous regions operate flotation plants at altitudes of 3000 to 5000 m above sea level (o.s.l.) where air density is considerably less than closer to sea level. For example, at sea level dry air has a density of 1.293 kg/m$^3$ while at 4.25 km it is only 0.736 kg/m$^3$, about 57% of the sea level value. The Antamina copper-zinc mine in Peru and the Collahuasi copper-molybdenum mine in Chile are at 4.2 and 4.1 km o.s.l. respectively.

The effect of gas density on flotation has not been well studied. Hales (1998) reported some initial testing but bubble size measurements were not included in the work. Work by Wilkinson and van Dierendonck (1990), on aerated bio-reactors at pressures from one to twenty atmospheres using different gas mixtures to vary density, suggested that only density and not gas type impacted bubble size. Mixtures of helium (density 0.1786 kg/m$^3$) and air were therefore used to simulate the effect of altitude on bubble size (at constant $J_g$) using a 5.5 litre Denver flotation machine instead of the much larger Metso unit. The results showing the effect on $D_{32}$ of gas density ($\rho_g$) relative to standard conditions, defined as one atmosphere and 20 °C ($\rho_o$), are shown in Figure 10 (left plot). The boxed area indicates the density region of practical interest. Use of the smaller Denver cell resulted in more data scatter since bubble size measurements were in the turbulent, near-impeller, and not quiescent region.

Figure 10 - The relationship between relative gas density ($\rho_g/\rho_o$) and $D_{32}$ (left plot). Standard conditions are for one atmosphere and 20 °C (box indicates region of practical interest). Comparison of $D_{32}$ versus $D_{10}$ for 100% air and 100% helium (right plot)
Fitting the data to a power relationship as was done for viscosity yielded an exponent of 0.132, indicating a weak effect of gas density on $D_{32}$ for the density range representing sea level to approximately 5 km o.s.l. The $D_{32}$ correction factor ($f_d$) for gas density effect is therefore:

$$f_d = \left( \frac{\rho_o}{\rho_g} \right)^{0.132} \quad (17)$$

As a check on the effect of gas density on the bubble size distribution a plot of $D_{32}$ versus $D_{10}$ (the arithmetic mean bubble size) is shown in Figure 10 (right plot). The overlapping of data sets for pure air and pure helium indicates the shape of the BSDs is not influenced by the type of gas thus supporting the conclusions of Wilkinson and van Dierendonck. Equation 12 can be used to show that at a maximum practical elevation of 5km, the effect is a 10% increase in $D_{32}$ and, conversely, a 10% decrease in $S_b$.

**Energy Input (Impeller Speed)**

Plant operators often consider the impeller rotational speed of the flotation machine as an adjustment that can be used to change the pumping rate of slurry, impact particle suspension and bubble-particle contact, or change bubble-size, and it was therefore included in the study. Energy drawn by the impeller is proportional to $N^3$ where $N$ is the rpm of the impeller, and as expressed by Equation 2 the tip speed varies linearly with $N$. Typically, industrial units have tip speeds of 5 to 7 m/s however, values up to 10 m/s have been noted in attempts to improve fine Pt recovery in South African plants (Deglon et al., 2000). Testing was conducted across the range 3.4 to 9.2 m/s, representing a 20-fold increase in power, for Jg values of 0.5 and 1 cm/s and at frother concentrations (DF-250) of 0, 2.5, 5 and 10 ppm. The effects on $D_{32}$ of the changes in impeller tip speed appear in Figure 11 and, perhaps surprisingly, show that between 4.6 and 9.2 m/s there is little if any appreciable effect at all frother levels and for both $J_g$ conditions. As the impeller reaches its lower operating limit at around 3.5 m/s there is a noted increasing trend in $D_{32}$.

![Figure 11 - D_{32} versus impeller tip speed for 0, 5, and 10 ppm frother (DF-250) at Jg of 0.5 and 1 cm/s.](image)

Error bars are for 95% confidence limits but only visible for 0 ppm frother.

**Interaction Effects**

The work has demonstrated that each frother type can be characterised by its CCC95 value, however, Figure 12 showing $D_{32}$ versus DF-250 concentration for increasing $J_g$ also shows that CCC95 increases linearly with $J_g$ as detailed in Figure 13 (left plot). Consequently, the metric for frother
characterisation is chosen as the CCC95 at $J_g = 0.5$ cm/s. Specific CCC95 at other values of $J_g$ can be determined from the following relationships.

\[ \text{CCC95} = \text{CCC95}_{J_g = 0.5} \cdot (0.6736 + 0.6528 \cdot J_g) \]  \( \text{(18)} \)

For alcohols (based on MIBC data):

\[ \text{CCC95} = \text{CCC95}_{J_g = 0.5} \cdot (0.8639 + 0.2723 \cdot J_g) \]  \( \text{(19)} \)

The notion of a limiting $D_{32}$ bubble size as illustrated in Figure 6 also appears to be frother-type dependent as shown in Figure 13 (right plot) with $D_{\text{limiting}}$ decreasing for frothers with higher CCC95. The factor ($f_l$) for $D_{\text{limiting}}$ to be incorporated into the overall $D_{32}$ relationship was determined to be:

\[ f_l = 0.0176 \cdot (0.06 - \text{CCC95}_{J_g = 0.5}) \]  \( \text{(20)} \)
The Overall D_{32} Model

An overall model for predicting $D_{32}$ was assembled from Equations 15, 16, 17, 18 and 20 and fitted to the data of the type shown in Figures 7 and 8. The global equation can be expressed as:

$$D_{32} = f_v \cdot f_d \cdot \varphi \cdot \left[ J_g, ppm, CCC95 \right]$$  \hspace{1cm} (21)

Where the function $\varphi$ is determined from:

$$\varphi = 0.267 + f_f + 0.064 \cdot (100 \cdot J_g)^{0.5} + \left[ 2.316 - (f_f + 0.267) + 0.0619 \cdot (100 \cdot J_g)^{0.5} \cdot \exp \left[ -3.09 \cdot \frac{ppm}{CCC95} \right] \right]$$  \hspace{1cm} (22)

Minimum $D_{32}$ curve (CCC99)  \hspace{1cm} Maximum $D_{32}$ curve (CCC0)  \hspace{1cm} Exponential decay

Figure 14 - Illustration of how the components of Equation 17 are used to construct the $D_{32} - J_g$ relationship for varying frother concentrations (DF-250 in this example)

The factors $f_v$, $f_d$ and $f_f$ are determined from Equations 18, 19 and 20 respectively. Figure 14 shows how the components of Equation 22 are used to construct the $D_{32} - J_g$ relationship for various frother concentrations (DF-250 in the example) from CCC0 (maximum $D_{32}$) to CCC99 (minimum $D_{32}$). Goodness-of-fit based on residuals analysis was performed between the $D_{32}$ model predictions and the 199 measured data points and are shown in Figure 15. Overall the model predictions at the 95% confidence level are ±0.30 mm for $D_{32}$ and data is reasonably well distributed on both sides of the 1:1 line as evident in Figure 15.
Figure 15 - Comparison of measured versus predicted D$_{32}$ for 199 tests

The D$_{32}$ model can similarly be used as a predictor for S$_b$ using Equation 11. The goodness-of-fit in this case, shown in Figure 16, results in 95% confidence limits of ±21% on a relative basis for S$_b$ prediction.

Figure 16 - Comparison of measured versus predicted S$_b$ for 199 tests

**THE BENCHMARKING “ROADMAP”**

The D$_{32}$ and S$_b$ models defined here can be used to assess plant performance by comparing measured plant values to those predicted. Model plots such as the one in Figure 14 can be prepared for a given frother type and plant conditions of altitude and pulp temperature, and the measured data then plotted against this background. To validate this approach, data from circuits at five operating plants using mechanical forced-air cells are plotted against a model based on DF-250. The five plants are listed in Table 3 and the comparisons are shown in Figure 17 for D$_{32}$ (upper plot) and S$_b$ (lower plot) as a function of J$_g$. 

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Cell volume varied from 16 to 130 m$^3$ for rectangular and tank cell designs and applications included copper, nickel, platinum and palladium circuits on four continents.

Table 3: Data from five plant sites were used to validate the $D_{32}$ and $S_b$ models as shown in Figure 17

<table>
<thead>
<tr>
<th>Company →</th>
<th>NA Palladium</th>
<th>BHP-Billiton</th>
<th>Xstrata Ni (Escondida)</th>
<th>Impala Platinum</th>
</tr>
</thead>
<tbody>
<tr>
<td>Operating Site →</td>
<td>Lac des Iles</td>
<td>Leinster (LNO)</td>
<td>Raglan</td>
<td>Los Colorados</td>
</tr>
<tr>
<td>Manufacturer, cell type</td>
<td>Outotec TC</td>
<td>Outotec 16U</td>
<td>Outotec 28U</td>
<td>Outotec TC</td>
</tr>
<tr>
<td>Cell volume, m$^3$</td>
<td>130</td>
<td>16</td>
<td>28</td>
<td>100</td>
</tr>
<tr>
<td>Circuit duty</td>
<td>R/S</td>
<td>R/S</td>
<td>R/S</td>
<td>R</td>
</tr>
<tr>
<td>Location</td>
<td>Ontario</td>
<td>Australia</td>
<td>Quebec</td>
<td>Chile</td>
</tr>
<tr>
<td>Metal/Mineral</td>
<td>Palladium</td>
<td>Nickel</td>
<td>Nickel</td>
<td>Copper</td>
</tr>
</tbody>
</table>

R=rougher, S=scavenger, C=cleaner, TC=tank cell, U=rectangular cell with U-bottom

Plant measurements in different cells within a circuit will obviously show scatter due to variations in mechanical condition, frother concentration and other factors, however the comparison between the plant data and model curves in Figure 17 is remarkably consistent considering the model was developed in the two-phase air-water system. The plant data follow the shape of the model curves at given frother levels, and all data fall within the minimum and maximum curves defined by the model. What is evident is that two plants, Lac des Iles and Escondida Los Colorados, are clearly operating with bubble size that is larger and surface area flux that is lower than might be considered optimum for recovery.

The potential for increasing $S_b$ by either reducing bubble size by increased frother addition, or increasing gas rate, or both, can be mapped by such an analysis. This, then, defines the “benchmarking road-map” tool referred to in this publication. Although the five plants listed here do not all use DF-250 as their frother the data can nevertheless be compared on an “equivalent DF-250” basis. Also of interest is that two plants are operating close to the minimum bubble size achievable, Raglan and Leinster, both in highly saline pulp environments where the high salt content (exceeding sea water salinity) contributes strong frother characteristics (Quinn, Kracht, Gomez, Gagnon & Finch, 2007) effectively producing an environment close to CCC99.
A CASE STUDY: LAC DES ILES

The Lac des Iles palladium mine in Ontario had poorer flotation recovery of Pd than projected from initial pilot plant studies (Martin, Ohrling, Olsen & Taggart, 2003). An investigation in 2003 by a joint SGS-McGill team studied the gas dispersion characteristics and metallurgical performance of the rougher-scavenger circuit and concluded the deficient recovery was due to poor recovery of minus 10 μm Pd (Nesset et al., 2005), likely the result of lack of small, <1 mm, bubbles compared to the pilot circuit. The plant cells averaged 10% minus 1mm bubbles while the pilot plant cells averaged 50%. A follow-up study by SGS in 2005 (Hernandez-Aguilar, Thorpe & Martin, 2006) tested this premise by introducing a stronger, blended alcohol-polyglycol frother to the last two cells in the circuit in addition to the standard MIBC used by the plant, and again measuring gas dispersion parameters and metallurgical behaviour.

Figure 18 illustrates the gas dispersion data from the 2003 and 2005 studies showing that the plant was operating under the same hydrodynamic conditions during both studies with frother concentration (MIBC) effectively being at 5 ppm, well below the CCC95 value of 12.4 to 14 ppm for \( J_g \) range 0.5 to 1 cm/s (Table 2). The 2003 pilot plant data showed significantly smaller \( D_{32} \) (upper plot) and higher \( S_b \) values (lower plot), equivalent to 10 ppm of MIBC on these gas dispersion “roadmaps”. Addition of the stronger blended frother in the 2005 study achieved the desired result of having \( D_{32} \) and \( S_b \) match the pilot
study data as evident in Figure 18. Note that $S_b$ increased by ~40% (at the same $J_g$) through this frother change.

![Figure 18](image)

**Figure 18** - The Lac des Iles case study gas dispersion “roadmaps”. Circles indicate 2003 and 2005 plant data. Squares indicate the 2003 pilot plant data which was matched in 2005 by the use of a stronger, blended frother than MIBC.

Table 4 presents a summary of the 2005 Lac des Iles metallurgical results obtained by Hernandez-Aguilar et al. (2006) when reducing bubble size by the addition of stronger frother to the last two cells in the rougher-scavenger circuit. Reducing the $D_{32}$ of the BSD to below 1.2 mm improved Pd recovery in all three size fractions, but most notable in the minus 10 μm fraction for which recovery improved by +40% (abs). Continuing on this approach of combining gas dispersion and metallurgical studies at the Highland Valley copper-molybdenum operation, Hernandez-Aguilar, Basi and Finch (2010) have demonstrated that reducing bubble size in molybdenum cleaner flotation columns by changing sparger technology in this case (instead of by a frother change) improved circuit Mo recovery by 4% (abs) with increased concentrate quality as well. The proportion of minus 1 mm bubbles was increased from less than 2% to over 20%. The authors argue that the improved metallurgical recovery within the individual columns is too large to be explained by $k$ being inversely proportional to $D_{32}$ in the $k$-$S_b$ relationship and suggest that the relationship may be an inverse square if one assumes that the $P$ term in Equation 13 is inversely related to $D_{32}$ and not a constant. In practical terms this means that recovery benefits, of very fine particles for example, by decreasing bubble size could be significantly greater than projected by the Gorain $k$-$S_b$ model. The large recovery improvement (+40% abs) in the minus 10 μm fraction with decreased bubble size at Lac des Iles would tend to support this hypothesis.
Table 4 - Impact on palladium recovery at Lac des Iles (last two rougher-scavenger cells) by changing bubble size distribution through additional frother (from Hernandez-Agular et al., 2006)

<table>
<thead>
<tr>
<th>Particle Size Fraction (microns)</th>
<th>Pd Recovery for D32 in Specified Bubble Size Range</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>&lt; 1.2 mm</td>
</tr>
<tr>
<td>+45</td>
<td>21.3</td>
</tr>
<tr>
<td>-45 +10</td>
<td>15.1</td>
</tr>
<tr>
<td>-10</td>
<td>69.1</td>
</tr>
</tbody>
</table>

**DISCUSSION**

Although the traditional flotation scaling parameters of power intensity, impeller tip speed and air flow velocity, have and continue to, serve the industry well, they are complimented by the addition of the gas dispersion parameters introduced over the past two decades. These parameters link performance and the importance of bubble size and frother effects in a manner that relate directly to the kinetics of flotation recovery. As cell size has steadily increased and new designs are proffered, these gas dispersion parameters increasingly serve as complementary scale-up criteria (Grönstrand & Niitti, 2007). It is noteworthy that most of the major cell manufacturers have obtained a set of the McGill gas dispersion sensors over the past decade (personal communication, C.O. Gomez).

Given the importance of bubble size in flotation, acknowledged since the technology was first introduced commercially a century ago, it is perhaps past due that a model of bubble size linking to the key operating variables be available. This communication updates an earlier one (Nesset et al., 2007) and presents the final D32 model that incorporates the important effects of frother concentration and type, as well as air rate, viscosity and altitude. Surprisingly, the impeller tip speed does not significantly impact the bubble size (or BSD) over a wide operating range of 4.6 to 9.2 m/s, represents an eight-fold increase in power intensity. This finding contradicts some earlier literature (Gorain et al., 1999; Grau & Heiskanen, 2003, 2005) but has received recent support (Grano, 2006; Grönstrand & Niitti, 2007). The key was to isolate the size measurements from the highly-turbulent near-impeller region where increasing impeller speed will increase fluid re-circulation thereby entraining a larger population of smaller bubbles. The Metso RCS™ cell with its internal shelf baffle proved to be a highly suitable design for the purpose of isolating and negating this effect by creating distinct turbulent and quiescent zones. By their definition as parameters linking to flotation recovery, the gas dispersion measurements relate to gas and bubbles that are exiting the pulp zone to enter the froth, and are not meant to measure near impeller conditions. It is beyond the scope of this paper to detail the hydrodynamic reasons why an eight-fold increase in power intensity does not decrease bubble size, however it is postulated that the high void fraction (25 to 50%) and free-stream turbulence within the impeller region could well lead to conditions where the disruptive forces acting on bubbles are less dependent on velocity (form drag) and more on viscosity (friction drag). This could explain the observed relationships relating to impeller speed and viscosity. Note that standard corrections for hydrostatic pressure as detailed by Gomez and Finch (2007) may still need to be applied to D32.

A significant and unique outcome of the model is the ability to predict the effect of frother type and concentration on D32, facilitated by the CCCX concept. Given a frother’s CCC95 value (characterised at Jg = 0.5 cm/s) and concentration relative to its CCC95 value, it has been shown that D32 for all frothers reduces in an exponential fashion as concentration is increased, reaching a limiting bubble size. This limiting size itself has a lower value for frothers having higher CCC95 (typically lower molecular weight alcohols). Using the D32 model, some important relationships emerge as evident from Figure 19. The left plot showing Sb as a function of frother concentration at various Jg, tells us that the desired frother concentration should be where the slope is flattening in order to avoid the sensitive steeply-sloped region where small changes in concentration will have a larger impact on Sb. Since CCC95 increases with Jg (Equations 18 and 19) a higher frother concentration will be required as Jg in a cell is increased. In this example, for DF-250, frother concentration ~12 ppm is indicated for Jg of 0.5 cm/s but increases to ~15 ppm for a Jg of 1.5 cm/s in order to maintain a small D32 and high Sb. The significance of achieving a lower limiting D32 by selecting frothers with higher CCC95, illustrated in the right plot of Figure 19, is that
smaller bubbles and hence higher $S_b$ can be achieved, conditions that may be advantageous for recovery kinetics of smaller particles.

![Diagram of frother concentration vs. Sb](image)

Figure 19 - Model predictions. The left plot shows that the optimum frother concentration (the oval overlay) increases with increasing $J_g$. The right plot shows that frothers with higher CCC95 (e.g. Pentanol) have lower $D_{\text{limiting}}$ and will produce higher maximum $S_b$.

Operators are still left with the requirement for having CCC95 values available for use in the model. Work is currently underway at McGill to predict CCC95 values from the structure of the frother molecule thus eliminating the need for testing time and equipment currently required. It turns out that the CCC95 for a frother (either alcohol or polyglycol) can be accurately determined from its molecular structure alone through the HLB parameter (Hydrophile-Lipophile Balance), a measure of its solubility in water. Figure 20 illustrates the relationships that have been established for polyglycol and alcohol-based frothers from data available in the literature (Grau & Laskowski, 2006) and from the current research (Nesset PhD Thesis, 2010). Laskowski’s CCC concept has been deemed equivalent to our CCC95. HLB can be empirically determined by applying the Davis equation (Davis, 1957; Davis & Rideal, 1961):

$$HLB = 7 + \sum \text{Hydrophilic group numbers} + \sum \text{Hydrophobic group numbers}$$  \hspace{1cm} (23)

The various hydrophilic (e.g. OH\(^-\)) and hydrophobic (lipophile) (e.g. CH\(_3\)) groups are assigned numbers that reflect their relative solubility contribution to the molecule. Details are beyond the scope of this paper but HLB values have been documented by Rao and Leja (2004) and Laskowski, Thone, Williams & Ding (2003). Typical values of HLB for frothers are in the range 4 – 10 with higher values indicating more soluble frothers. Work is ongoing on a broader spectrum of frothers and other surfactants to more closely link the HLB parameter to CCC95 and the limiting bubble size and will be the subject of future communications. What has been established by this initial work is that frother behaviour in terms of its effect on $D_{32}$ and $S_b$ can be predicted by the HLB-CCC95 relationship if the molecular structure of the frother is known. The on-going work also includes blended frothers.
Figure 20 - The relationship between CCC95 and HLB for polyglycol and alcohol frothers (MW is the molecular weight)

The more practical and beneficial (to operators) use of the model may well be as a benchmarking tool for assessing their plant’s gas dispersion parameters on the framework of the \( D_{32} \) and \( S_{b} \) “roadmaps” as illustrated by the data of the five plants shown in Figure 17. The resulting data, imposed on the model curves for the frother and other conditions appropriate to the plant, will highlight the potential for recovery improvement (increased kinetics) via bubble size reduction and increased air rate. The Lac des Iles study has served as a compelling example of the importance of considering and manipulating the gas dispersion parameters within flotation circuits. Downstream effects, such as concentration build-up and froth persistence, will also need to be considered in frother selection and dosage.

Although the \( D_{32} \) model has been developed within the 2-phase water-air system, the excellent correlation with 3-phase plant data lends validity to its application within the operating plant environment. One can assume that the effect of solids is, in most cases, a secondary one compared to the predominant effects of frother, gas rate and liquid viscosity. There may well be conditions of high solids content, very fine particle size and high bubble loading for which the assumption of secondary effects becomes invalid. The reader also needs to be reminded that frother concentration should always be considered as the residual concentration in solution since there are several situations under which frother can be removed from solution, such as adsorption on mineral surfaces (e.g. talc), preferential fractionation to the froth phase, and “stripping” from solution by air/gas. Recycled water containing organics to the flotation pulp, and high salinity can also add frothing properties (Quinn et al., 2007; Azgomi, Gomez & Finch, 2009). There are new methods for assessing frother concentration in solution (Gélinas & Finch, 2007; Zangooi, Gomez & Finch., 2010).

CONCLUSIONS

- A robust model for the prediction of \( D_{32} \), and by extension \( S_{b} \), has been developed for mechanical forced-air flotation machines within the 2-phase air-water system, however, the model seems suitably robust for application to many 3-phase systems based on validation data from operating plants.
- The model incorporates the role of both frother type and concentration and introduces that the parameter CCC95, which is characteristic of frother type, can be established from the molecular structure of the frother through the HLB concept.
- The model has established that frother concentration is the most important variable affecting bubble size, followed by air rate and liquid viscosity. The effect of altitude was simulated by using a mixture.
of air and helium, and was judged to be of secondary importance increasing $D_{32}$ by about 10% at the maximum altitude of currently operating plants (4 to 5 km above sea level).

- The effect of impeller speed was tested over a wide range and was found to have no appreciable effect on $D_{32}$ (or BSD) over a broad operating range of 4.6 to 9.2 m/s, representing an eight-fold increase in power intensity, and was not included in the model.
- The model is seen as a powerful benchmarking tool that permits operators to map measured gas dispersion parameters in their plants in order to assess potential for improving recovery through bubble size reduction and increased air rate.

### NOMENCLATURE

<table>
<thead>
<tr>
<th>Symbol</th>
<th>Units</th>
<th>Definition</th>
</tr>
</thead>
<tbody>
<tr>
<td>$A$, $a$, $B$</td>
<td>Various constants and parameters in the $D_{32}$ model</td>
<td></td>
</tr>
<tr>
<td>$J_g$</td>
<td>cm/s</td>
<td>Superficial gas velocity, gas rate, air rate</td>
</tr>
<tr>
<td>$d_i$</td>
<td>mm</td>
<td>Individual bubble diameter</td>
</tr>
<tr>
<td>$D$</td>
<td>m</td>
<td>Impeller diameter</td>
</tr>
<tr>
<td>$D_{32}$</td>
<td>mm</td>
<td>Sauter mean diameter of the BSD, total volume of bubbles divided by total surface area of bubbles assuming spherical shape</td>
</tr>
<tr>
<td>$D_{10}$</td>
<td>mm</td>
<td>Arithmetic mean diameter of the BSD, total diameter of bubbles divided by total number of bubbles</td>
</tr>
<tr>
<td>$D_s$</td>
<td>mm</td>
<td>“Creation” bubble size at $J_g = 0$ cm/s</td>
</tr>
<tr>
<td>$D_l$, $d_l$, $d_{limiting}$</td>
<td>mm</td>
<td>Limiting bubble size as frother concentration $\to \infty$</td>
</tr>
<tr>
<td>$S_b$</td>
<td>s$^{-1}$</td>
<td>Bubble surface area flux, total surface area of bubbles per unit area of cell per unit of time</td>
</tr>
<tr>
<td>$Q$</td>
<td>m$^3$/s</td>
<td>Volumetric flow rate of gas</td>
</tr>
<tr>
<td>$A_{cell}$</td>
<td>m$^2$</td>
<td>Cross-sectional area of the cell</td>
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<tr>
<td>$V_g$</td>
<td>m$^3$</td>
<td>Volume of gas in cell</td>
</tr>
<tr>
<td>$V_{cell}$</td>
<td>m$^3$</td>
<td>Volume of cell</td>
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<tr>
<td>$k$, $k_{pulp}$</td>
<td>s$^{-1}$</td>
<td>Flotation rate constant, overall and in pulp</td>
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<tr>
<td>$R_f$</td>
<td>fraction</td>
<td>Froth phase recovery factor</td>
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<tr>
<td>BSD</td>
<td></td>
<td>Bubble size distribution</td>
</tr>
<tr>
<td>HLB</td>
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<td>Hydrophilic - Lipophilic Balance, empirical relationship measuring contribution of polar and non-polar groups for a frother in aqueous solution</td>
</tr>
<tr>
<td>RCSTM</td>
<td>Reactor Cell System, Metso Minerals patented cell design</td>
<td></td>
</tr>
<tr>
<td>$N$</td>
<td>rpm</td>
<td>Impeller rotational speed/minute</td>
</tr>
<tr>
<td>$P$</td>
<td>fraction</td>
<td>Floatability parameter</td>
</tr>
<tr>
<td>$P_{net}$</td>
<td>W</td>
<td>Power input</td>
</tr>
<tr>
<td>$P_I$</td>
<td>W/m$^3$</td>
<td>Power intensity or specific power, net power divided by the cell, liquid or impeller-swept volume</td>
</tr>
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<td>$S$</td>
<td>m/s</td>
<td>Impeller tip speed</td>
</tr>
<tr>
<td>CCC</td>
<td>ppm</td>
<td>Critical coalescence concentration</td>
</tr>
<tr>
<td>CCC95</td>
<td>ppm</td>
<td>Concentration for which 95% reduction in $D_{32}$ has been reached</td>
</tr>
<tr>
<td>CCCX</td>
<td>ppm</td>
<td>Concentration for which X% reduction in $D_{32}$ has been reached</td>
</tr>
<tr>
<td>PPM, ppm</td>
<td>g/t</td>
<td>Concentration expressed as parts per million</td>
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<td>Viscosity correction factor in $D_{32}$ model</td>
</tr>
<tr>
<td>$f_d$</td>
<td></td>
<td>Altitude (gas density) correction factor in $D_{32}$ model</td>
</tr>
<tr>
<td>$f_l$</td>
<td></td>
<td>Limiting bubble size correction factor in $D_{32}$ model</td>
</tr>
<tr>
<td>$\sigma$</td>
<td>N/m</td>
<td>Surface tension</td>
</tr>
<tr>
<td>$\rho$</td>
<td>kg/m$^3$</td>
<td>Density</td>
</tr>
<tr>
<td>$\mu$</td>
<td>Pa-s</td>
<td>Dynamic viscosity</td>
</tr>
<tr>
<td>$\varphi$</td>
<td></td>
<td>Symbol for mathematical function in $D_{32}$ model</td>
</tr>
<tr>
<td>$\xi$</td>
<td>%, fraction</td>
<td>Gas holdup or void fraction of gas in cell volume</td>
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ACKNOWLEDGEMENTS

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