AN INVESTIGATION OF THE GAS DISPERSION PROPERTIES OF MECHANICAL FLOTATION CELLS: AN IN-SITU APPROACH

by

Sanja Miskovic

Dissertation submitted to the faculty of the Virginia Polytechnic Institute and State University in partial fulfillment of the requirements for the degree of

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Gerald H. Luttrell, Chairman
Roe-Hoan Yoon
Gregory T. Adel
Dariusz Lelinski
Bartosz Dabrowski

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ABSTRACT

Bubble size is considered to be one of the most important parameters affecting the performance of froth flotation cells. However, monitoring, controlling and predicting bubble size is a very challenging task. This dissertation presents results obtained from a comprehensive pilot- and industrial-scale experimental investigation of gas dispersion performance of two commercially available flotation cells. To facilitate this investigation, a continuous pilot-scale flotation system was developed and tested. The results of the hydrodynamic and metallurgical testing conducted on the pilot-scale flotation circuit are presented. In addition, an assessment of the impact of two commercially available rotor/stator mechanism designs on bubble generation was performed under non-coalescing conditions. Based on obtained results, the mechanisms of gas dispersion throughout the flotation cell and gas cavity formation behind the impeller blades have been presented and discussed.

A new in-situ optical bubble sampling method was also developed as part of this investigation. The new system allowed an accurate estimation of local bubble sizes and determination of overall gas dispersion patterns within the cell. The new method was compared to the existing ex-situ bubble sampling method commonly used in industry. Two image analysis techniques were also evaluated, i.e., a template matching BubbleSEdit technique and the edge detection Northern Eclipse technique. Significant variations in bubble size as a function of the sampling method, sampling location, operating condition, machine type and image analysis method were observed. Generally, bubbles observed with the in-situ sampling method appeared to be larger than bubbles recorded with the ex-situ method. Furthermore, the mean bubble size determined by the Northern Eclipse bubble sizing method was smaller than the BubbleSEdit value. The experimental tests also revealed that sampling location had a strong effect on measured local mean bubble size and bubble size distribution in both vertical and horizontal directions. In addition, aeration rate was found to have a profound impact on the gas dispersion pattern in the cell and on local bubble size. Agitation rate also had a significant effect on bubble size, although the degree of impact strongly depended on the agitation level, chemical conditions in the cell and the machine type.
DEDICATION

For Rada and Slavica
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CHAPTER 1:
INTRODUCTION

The importance of bubble size on flotation efficiency has been long recognized. Using high-speed cinematography in his experiments (Bennett, 1958) concluded that, for a constant air supply rate, flotation rate increases by reducing the bubble size and increasing the number of bubbles. Recent intensification in research efforts to better understand gas dispersion and hydrodynamics in flotation cells has come about because of the growing need to design larger and more efficient flotation cells to treat the lower grade and more finely disseminated ores that are currently being mined (Sawyerr et al., 1998). Therefore, among hydrodynamic parameters, gas dispersion is considered to be the key. In this context, gas dispersion is defined as the dispersion of air into bubbles. It is well documented that the gas dispersion properties in the flotation process have a direct effect on the process performance (Schwarz and Alexander, 2006).

Accurate data about gas dispersion are required for understanding of the physical processes that govern the flotation. Among all gas dispersion parameters, the bubble size and the bubble size distribution are the most important. Bubble size governs the surface area over which solid particles and bubbles interact, which contributes significantly to system hydrodynamics and overall flotation performance. Local bubble size and bubble size distribution in the flotation cell strongly depends on various operational, technical, and chemical factors whose effects on bubble size should be taken into account while designing or modeling a flotation process.
Turbulent flotation models developed to date cannot predict bubble size in the flotation system from commonly measured flotation parameters such as pulp surface tension, aeration rate, energy input, particle size, and particle surface characteristics. Due to this limitation, the bubble size and bubble size distribution must be obtained experimentally and subsequently used as an input for current flotation models. On the other hand, bubble sizing in the pulp phase is difficult because of the high concentration of bubbles and solid particles in standard flotation cells (Nelson and Lelinski, 2000). The sizing technique must be robust and rugged to withstand the environmental conditions wherever it might be installed. This is especially important if equipment has to be deployed in industrial environments, where erosive/corrosive conditions are present and flotation cells may be dusty and vibrating and difficult to access.

There are many bubble sizing systems currently available, most of which are design to measure bubble size in two-phase laboratory conditions. Only few bubble sizing methods have been successfully employed in real industrial settings. Although they are typically very easy to operate (bubble sample is drawn from the top, quiescent zone of the cell to the external viewing box where bubbles are viewed) the results obtained are not particularly reliable. Only few studies on spatial distribution of bubble sizes in the stirred vessels have been published so far. On the other hand, a large amount of data is necessary for the successful assessment, support, and validation of the current and future models. Therefore, there is a great necessity for a rugged, reliable, and efficient bubble sampling system that can be used for bubble sizing in different regions of flotation cell.

The gas-liquid hydrodynamics in the system will also strongly depend on the type of bubble generator used, so the choice of a proper impeller/stator assembly to satisfy the
necessity of optimal gas dispersion is the key for the success and economy of the process. However, the mechanism of bubble generation in flotation systems has not yet been explored in depth and the differences between the bubble generation mechanisms of different flotation cells have not yet been well addressed in the literature.

1.1. OBJECTIVES

The main objective of this dissertation is to generate a better understanding of gas dispersion properties of different mechanical flotation cells. Also, the mechanisms affecting and controlling bubble size, and thereby the phenomena of bubble generation in different mechanical flotation cells, have been investigated. Specifically, this study primarily deals with the effect of hydrodynamic variables on bubble size in pilot-scale flotation cells, although some measurements were also performed on full-scale industrial machines. The results of this study provides ample data necessary to support and validate current flotation models, which are relevant to the design, scale-up, optimization and control of mechanically agitated flotation cells.

The main goals of this research are:

- to design and construct a new in-situ bubble sizing method for accurate measuring of gas dispersion properties in pilot-scale flotation cells;
- to compare the performance of the new in-situ method with another commercially available bubble sizing method;
- to determine operating limits of the new system in both two- and three-phase flotation environments;
to design and construct a new, automated, modular, pilot-scale flotation circuit, which can accommodate different impeller/stator mechanisms and allow continuous testing of multiple process parameters while the system is operated as a batch reactor or continuously;

• to use developed systems to study gas dispersion properties of different mechanical flotation cells over various operating conditions;

• to investigate the effect of sampling location, aeration rate, agitation rate and image analysis technique on estimated bubbles size;

• to study gas dispersion efficiency of different mechanical flotation cells (forced-aerated vs. self-aerated flotation systems);

• to evaluate and characterize bubble generation mechanisms of different rotor/stator mechanisms;

• to perform hydrodynamic and metallurgical performance investigation of the pilot-scale flotation cell; and

• to provide data for validation and support of current flotation models.

1.2. OUTLINE

This dissertation is composed of three papers, which have been published or submitted to a peer-reviewed journal or are ready for submittal (Chapters 3 to 5).

A literature review pertaining to fundamental concepts of the flotation process are given in the Chapter 2. Also included in the Chapter 2 is the overview of the current research on bubble formation, gas dispersion and bubble size measurement techniques.
Chapter 3, *A New Modular Pilot-Scale Setup for Hydrodynamic and Metallurgical Flotation Performance Evaluation* has been presented during the Flotation ’11 Conference, Cape Town, South Africa, and is to be submitted for a review for the special issue of the Minerals Engineering by December 2011. Dr. Gerald Luttrell and Mr. Robert Bratton served as co-authors in this work, and are acknowledged as such at the beginning of the chapter. This chapter also presents details of the pilot-scale flotation circuit design and describes results obtained during hydrodynamic and metallurgical investigation of the pilot-scale flotation cell.

Chapter 4, *Comparison of Two Bubble Sizing Methods for Performance Evaluation of Mechanical Flotation Cells* has been presented during the Roe-Hoan Yoon International Symposium on Advanced Separation Processes and Resource Engineering, SME Annual Meeting, 2011, Denver, Colorado, and is published as a part of the conference proceedings. Dr. Gerald Luttrell also served as co-author in this work. This chapter presents details of the new in-situ bubble sizing system design and presents the results of detailed investigation of effects of bubble sampling method, sampling location, operating conditions, and image analysis technique on estimated bubble size.

Chapter 5, *Comparison of gas dispersion mechanisms for forced-air and self-aerated mechanical flotation cells* has also been presented during the Flotation ’11 Conference, Cape Town, South Africa, and is to be submitted for a review for the special issue of the Minerals Engineering by the December 2011. Dr. Gerald Luttrell, Dr. Saad Ragab, and Mr. Hassan Elhady Fayed served as co-authors in this work. Gas dispersion patterns within two different mechanical flotation cells are presented in this chapter.
Also, mechanisms of bubble generation in the impeller/stator zone of the flotation cell are discussed.

1.3. REFERENCES


CHAPTER 2:
LITERATURE REVIEW

2.1. HISTORY OF FLOTATION

Flotation is the most widely used separation process in the mineral processing industry today. The importance of flotation technology in the global economy is significant. A rough estimate of the quantity of crushed ore treated by flotation is about nine billion tons per year, and the portion of base metals processed using this method is approximately 95% (Brewis, 1996).

The idea of flotation dates back to 1860 when William Haynes claimed that fine sulfide ore could be agglomerated by oil and separated from the gangue material by washing (Fuerstenau et al., 2007). Even though Haynes’s idea had no known commercial application at the time, his process, originally named bulk-oil flotation, is considered to be the first patent in the field of flotation. The first commercial flotation process was designed and successfully tested by Adolf Bessell in his factory in Dresden, Germany, in the late 1860s (Lynch, 2010).

Today flotation is used for separation of almost all sulfide and many non-sulfide metallic minerals, industrial minerals, and energy minerals such as coal and bitumen. Though mainly used in the mineral processing industry, the flotation process has also been used in other industrial fields such as wastewater treatment and paper recycling, for the removal of organic contaminants from effluents in the diary and beer industry, and for remediation of contaminated soil (Brewis, 1991).
2.2. MACROSCOPIC DESCRIPTION OF FLOTATION PROCESS

Froth flotation is a complex physico-chemical process that utilizes natural and induced hydrophobicity to separate and collect valuable mineral particles from slurry (Malati, 1984). In flotation, hydrophobic minerals suspended in the aqueous phase are collected by air bubbles from a solid-liquid suspension, as illustrated in the Figure 2.1. Due to bubble buoyancy, bubble-particle aggregates are transported from the pulp to the top of the flotation cell where they accumulate as froth. Accumulated froth is removed and valuable mineral concentrate is recovered. The unattached hydrophilic particles remain in the pulp, and are either discarded or reprocessed separately. Flotation is, therefore, a heterogeneous, multiphase, and multi-component separation process.

Figure 2.1. Selective attachments of hydrophobic particles to air bubbles.

Efficiency of the flotation process is directly related to the number of collisions between particles and bubbles, which are strongly dependent on the ratio of particle diameter to bubble diameter. In a flotation system where bubbles are much larger than particles, a flow streamlines around the bubble sweep particles near the bubble surface and prevents attachment of valuable mineral particles to the bubble. Hence, in order to
provide optimal conditions for the flotation, it is necessary to generate bubbles with sizes similar to the size distribution of particles in the pulp.

Prior to the flotation stage, all crushed and ground material is conditioned with various reagents, including collectors, frothers, regulators (activators, pH regulators, and depressants), and other surface modifying agents. The role of collectors is to form a hydrophobic surface film on a given mineral and to increase its hydrophobic affinity, thus allowing better conditions for recovery of the processed material by attaching valuable particles to air bubbles. Frothers are surface-active chemicals used to reduce water surface tension, enabling generation of smaller bubbles and formation of a stable froth phase. The main purpose of regulators is to alter the action of the collector and enable adsorption of the collector on targeted particles (Bulatovic, 2007).

The three distinct zones within a flotation cell are the turbulent zone, quiescent zone, and the froth zone, as shown in Figure 2.2. The rotating action of the impeller in the turbulent zone (Zone I) provides the energy necessary to keep particles in suspension, enables the generation of small bubbles, and maintains the hydrodynamic conditions needed for efficient bubble-particle interaction. The second zone (Zone II) is the quiescent zone. This region is less energy intensive than the turbulent zone, and provides conditions for detaching entrained or entrapped gangue particles from created aggregates. This zone also helps maintain a quiescent pulp-froth interface, which stabilizes the froth phase. The froth phase (Zone III) is the upper cleaning zone of the process.

There are three major processes that can be identified within the froth zone of the flotation cell. Generated dense bubble-particle aggregates are carried vertically from the pulp-froth to the air-froth interface and horizontally toward the lip of the flotation cell,
and then collected from the launder as a mineral concentrate. Less hydrophobic particles, loosely attached to the bubble surface, detach and drop back to the collection zone. Fine particles can be entrained in the water contained within the Plateau borders, in a capillary channel between three adjacent bubbles in the froth phase.

Figure 2.2. Flotation cell and its basic components.

Finch defines flotation as a two stage process each consisting of two operations, reaction and separation, as shown in Figure 2.3 (Finch, 1995). Slurry is the medium that carries mineral particles that are fed into the flotation cell and reaction stage. In the
reaction stage, different chemical reagents are added into the system in order to create
differences in particle surface properties, and in that way promote selective loading of
particles on the bubble surface. Typical particle sizes in a flotation system range from a
few microns to several hundred microns in 10% to 40% solid suspensions (Yianatos,
2007).

![Conceptual flotation design.](image)

**Figure 2.3.** Conceptual flotation design.

Additional energy has to be added to the flotation system to keep the solids in a
suspension and to disperse the large volume of induced air into fine bubbles. For
industrial size flotation cells, a specific power introduced into a system typically goes
from 0.5 to 1 kW/m³. Bubbles sizes generated in the flotation are normally in the range
from 0.2 to 3 mm (Fuerstenau et al., 2007).

Due to the extensive particle size distribution present in the pulp, it is not possible
to achieve and maintain optimum hydrodynamic conditions for all particle sizes
simultaneously. For optimal flotation of large, coarse particles, the specific power input should be kept at the lowest level necessary to keep particles in suspension. On the other hand, a much higher power input is required for fine particles (Schubert, 1999).

2.3. Flotation Equipment

There are four main types of flotation cells used in the mineral processing industry: mechanical flotation cells, pneumatic flotation cells, froth separators, and flotation columns.

Based on the method of air introduction, flotation equipment can also be divided into the following flotation groups: mechanical, pneumatic, dissolved air, vacuum, and electroflotation (Brewis, 1991; Brewis, 1996; Young, 1982). Bubbles are generated and dispersed by forced introduction of the air through a deeply submerged rotating impeller (forced-air mechanical cells), by self-aeration of shallow rotating impellers (self-aerated mechanical cells), by self-aeration through an orifice (Jameson cell) (Evans et al., 1992); by various spargers (flotation columns), and by parallel introduction of air and slurry through an in-line mixer (Microcel) (Yoon, 1987).

From the beginning, mechanical flotation cells have been the most widely used flotation cells in the mineral industry. Mechanical flotation cells consist of a tank, typically cylindrical shaped, fitted with an impeller drive assembly, and a stator. A main function of the stator, which is positioned around the impeller, is to transform tangential flow of the pulp in the cell in the radial direction. The impeller, on the other hand, provides the energy necessary for successful flotation operation and is therefore considered to be the heart of the flotation cell. Bubbles are generated and dispersed by
forced introduction of the air through a deeply submerged rotating impeller in forced-air mechanical cells and by self-aeration of shallow rotating impellers in self-aerated mechanical cells.

Schematic representation of the two main types of mechanical flotation cells is shown in Figure 1.4. In the forced-aerated cell (Figure 2.4.a), the agitator mechanism is typically positioned at the bottom, but sometimes in the center, of the cell (Lelinski, 2005). As shown in Figure 2.4.b, the agitator mechanism in self-aerated cells is located near the top of the cell. For both cell types, the impeller/stator assembly has to be designed and installed so as to allow re-circulation of pulp through the agitator zone in order to keep particles in a suspension and to provide good conditions for bubble-particle interaction (Yianatos, 2007).

![Figure 2.4. Two basic types of conventional flotation cells with flow patterns: a) Forced-aerated cell; b) Self-aerated cell. Red zone – high intensity turbulent zone (after Yianatos, 2007)).](image)

Most of the mechanical flotation cells used in the mineral processing industry today are manufactured and sold by three leading suppliers of mineral processing...
equipment: FLSmidth Dorr-Oliver Eimco (Dorr-Oliver, WEMCO, XCell), Outokumpu (OK FreeFlow, OK Multi-Mix, FloatForce™), and Metso (RCS, DR).

The main goal in the process of designing of any flotation cell is to maximize gas-liquid interface in the pulp (particle-water-air mixture) and hence to increase the probability of collisions between air bubbles and hydrophobic particles. Furthermore, every flotation cell should be designed and built to provide all of the following performance functions:

- to generate adequate turbulent conditions for successful bubble-particle attachment in the contact zone,
- to allow sufficient solids suspension,
- to perform efficient gas dispersion,
- to allow unhindered particle transfer from the pulp to the froth zone,
- to allow proper froth removal, and
- to provide optimal retention time for all three phases (gas, liquid, and solid) necessary for achieving maximal material recovery. (Lelinski, 2005).

The total energy introduced by mechanical agitation is, therefore, spent to perform three main functions: particle suspension and transport, gas breaking into bubbles and their dispersing throughout the flotation cell, and generation of micro-turbulences necessary to facilitate bubble-particle collisions (Figure 2.5).
Figure 2.5. Distribution of the total power consumed during the flotation process.

A fraction of the energy supplied to the system is transported throughout the cell by means of kinetic energy contained in the generated turbulent eddies of different scales. Most of this energy is dissipated through micro-interactions of all three phases, which include:

- bubble-particle collisions, attachment, and aggregation (Luttrell and Yoon, 1992a),
- liquid-particle viscous friction and lubrication (Eskin et al., 2005),
- particle-particle partially inelastic collisions.
2.4. FLOTATION CIRCUITS

Slurry is a water mixture of particles of different composition, size, shape, and density. Due to the complex nature of the slurry, the most efficient and selective way to separate valuable minerals from slurry is to arrange individual flotation cells in series, or banks. In this way, both residence time of particles and the number of bubble–particle collisions in the system are increased. Typical residence time for particles in a bank of cells ranges from 5 to 15 min. (Wills, 2006).

To achieve better grades of recovered concentrate from a single stage flotation circuit, re-floating is required in one (cleaner) or more (recleaner) additional stages. A series of cells producing a primary concentrate is called the rougher stage. The process of retreatment of the rougher tailings is called a scavenging stage. In order to reduce entrainment in the froth and recover only high grade particles, pulp in the cleaner stage is treated with lower reagent concentrations and generally has lower density. In order to maximize the recovery of valuable minerals, higher dosages of reagents and longer residence time are needed during the scavenger stage of the process. Mass flow rate and material properties through the flotation circuit are major factors determining the number and size of flotation cells that need to be installed in a processing plant (Gupta and Yan, 2006). Several examples of flotation circuits are presented in Figure 2.6.
Figure 2.6 Basic types of flotation cell arrangements.
2.5. FLOTATION SCALE-UP AND DIMENSIONLESS NUMBERS

Only recently were the basic microscopic processes of flotation and development of scale-up procedures necessary for successful design and operation of full-scale industrial cells from the lab- and pilot-scale data identified (Schubert and Bischofberger, 1998).

Flotation is governed by a number of different process variables that can be classified into three general groups, as shown in Figure 2.7: chemical, operational, and cell factors. (Harris, 1976; Smar et al., 1994). In the past, the role of cell variables has received less attention than chemical factors, even though they are a vital part of the flotation process.

Dimensionless numbers are often used to quantify the effect of different cell factors on metallurgical performance such as grade and recovery. Brandeer (Brander, 1993), claims that the appropriate application of dimensional similitude can lead to the discovery of forgotten or excluded variables during the scale-up process.

The advantage of using dimensionless numbers is reflected in the following (Ruzicka 2008):

a) dimensionless numbers reduce the number of variables needed to describe the process, simplifying the correlation of the experimental data;

b) dimensionless numbers allow simplification of the governing equations;

c) dimensionless numbers provide valuable scale estimates of key physical quantities;

d) dimensionless numbers contribute to the physical understanding of certain phenomena since they typically have a clear physical interpretation.
Figure 2.7 Variables affecting the flotation process.

There are many publications available relating the flotation metallurgical performance and dimensionless numbers, such as the Reynolds number, the Froude number, the Power number, the Air Flow number, and the Weber number (Deglon et al., 2000; Deglon et al., 1999; Mavros, 1992; Nelson and Lelinski, 2000; Rodrigues et al., 2001). Some of the most commonly used dimensionless numbers for the hydrodynamic evaluation of flotation processes are presented in Table 2.1. Typical number ranges found in the literature are also given here.

The correlation of dimensionless numbers with metallurgical performance of flotation was discovered by many authors. Deglon found a large variation of impeller
Reynolds numbers, aeration numbers, Power numbers, and Froude numbers during his tests on industrial-scale cells (Deglon et al., 2000). Although he didn’t provide an explanation for these variations, Mavros suggested that rotor aspect ratio and ratio of particle size to tank diameter might be the reason for these variations (Mavros, 1992). Rodrigues provided the best evidence of the dependence of the Reynolds number, the Froude number, and the Power number with recovery in a laboratory-scale mechanically agitated flotation cell (Rodrigues et al., 2001). For self-aerated, WEMCO, industrial flotation cells, Nelson and Lelinski provided data to support effective usage of hydrodynamic design criteria for scaling-up of cell volumes by a factor of 300, while still maintaining expected performance levels (Nelson and Lelinski, 2000).

Table 2.1. Dimensionless groups used to characterize flotation cell hydrodynamics (Gorain et al., 1996).

<table>
<thead>
<tr>
<th>Group</th>
<th>Definition</th>
<th>Description</th>
<th>Typical range</th>
</tr>
</thead>
<tbody>
<tr>
<td>Reynolds number</td>
<td>$Re = \frac{\rho_f ND^2}{\mu}$</td>
<td>gives a measure of the ratio of inertial forces to viscous forces</td>
<td>$10^6 - 7 \times 10^6$</td>
</tr>
<tr>
<td>Froude number</td>
<td>$Fr = \frac{DN^2}{g}$</td>
<td>defined as the ratio between the centrifugal or inertial forces due to the pumping action of the impeller and the gravitational forces due to hydrostatic head in the flotation cell</td>
<td>0.1 - 5</td>
</tr>
<tr>
<td>Power number</td>
<td>$N_p = \frac{P_0}{\rho_f N^3 D^5}$</td>
<td>defines a critical ratio between the absorbed power and the theoretical power draw which should be independent of cell size</td>
<td>0.5 - 5</td>
</tr>
<tr>
<td>Weber number</td>
<td>$We = \frac{\rho_f N^2 D^3}{S}$</td>
<td>represents the ratio between inertial forces to surface forces and plays an important role in bubble generation and distribution within an agitated vessel</td>
<td></td>
</tr>
</tbody>
</table>
Air flow number  \( Q_l = \frac{Q}{ND^3} \) defines a critical ratio between the aeration rate and the impeller pumping rate which should be independent of cell size 0.01 - 0.25

where:  
- \( N \) - impeller speed;  
- \( D \) – impeller diameter;  
- \( Q \) – volumetric air flow rate;  
- \( P \) – net power input;  
- \( \rho_f \) - slurry density;  
- \( g \) - gravitational constant;  
- \( \gamma \) - surface tension;  
- \( \mu \) - slurry viscosity.

After many years of research there is no general agreement on how to correlate dimensionless numbers with overall flotation scale-up process(es), and each equipment manufacturer emphasizes different aspects of the scale-up procedure and uses those arbitrarily for flotation cell design and scale-up (Yianatos, 2007).

Despite the great advances of general flotation process knowledge, the mechanisms, design, and scale-up of industrial flotation cells are still not fully understood and developed. Due to the complexity and multifaceted nature of the problem, development of a comprehensive theoretical model is required, which would include both hydrodynamic and chemical factors and provide a reliable scale-up approach.

2.6. FLOTATION KINETICS AND MODELING

Flotation is a process which occurs in time and its results are strongly determined by a series of random effects and factors. Unlike any other separation process, increased duration of the separation process does not necessarily improve the metallurgical performance. Increase in bubble-particle interaction time typically increases the recovery of the valuable mineral but leads to product grade deterioration.

A stable attachment between a particle and a bubble is formed when a large enough kinetic energy of colliding species is present at the moment of their collision. On the one hand, if the kinetic energy exceeds the potential energy of the bubble-particle
interaction, it can result in particle detachment from the bubble. Both the particle-bubble collisions and particle kinetic energy are random. The higher probability of bubble-particle connection, the faster is the flotation process. Additionally, the kinetics of the flotation process depicts not only the statistical character of the phenomena on the phase boundaries, but also depends on the continuous inflow of free surface into the flotation system. The fresh surface that continuously enters the flotation system comes both in the forms of fresh mineral particles (slurry feed) and continuously generated air bubbles.

As a result of the analogy between the bubble mineralization occurring during the flotation process and the chemical reaction, flotation kinetics is typically described with the equation for the kinetics of chemical reaction. Zuniga was the first author to describe the kinetics of batch flotation by using of the differential equation for the kinetics of chemical reaction (Zuniga, 1935). After Zuniga, many other authors worked on the problem of the flotation kinetics (Beloglazov, 1939; Bushell, 1962; Gaudin et al., 1942; Gorain et al., 1998a; Lazić and Ćalić, 2000; Sutherland, 1948; Szatkowski, 1987; Volin, 1964; Yianatos, 2007).

The general form of n-order differential equations describing the kinetics of the flotation process is (Derjaguin and Dukhin, 1961; Sutherland, 1948):

\[
\frac{dN_1}{dt} = -kN_1^n
\]  
(Eq. 2.1)

where \( N_1 \) is a concentration of floatable particles remaining in the flotation chamber up to the moment \( t \); \( k \) is the flotation rate constant; and \( n \) is a constant characterizing the order of flotation kinetics. The flotation rate constant, \( k \), is a macroscopic parameter containing information about the factors affecting the process. Flotation rate depends on such factors
as particle composition, size, surface properties, and reagent adsorption. Equation 2.1 describes the flotation kinetics of particles homogeneous in their surface properties that possess the same value of the flotation rate constant. Consequently, each floatable component may be represented by its continuous flotation rate (Polat and Chander, 2000).

Assuming identical chemical environments, the variable characteristics of this set of distributions during scale-up can be attributed to the differences in aeration and hydrodynamics between cells of different scale.

It is widely recognized that accuracy of the flotation kinetic model strongly depends on the initial assumptions for the mixing conditions in the cell (Harris, 1978; Yianatos, 2007). First order kinetics, originally introduced by Gaudin and Schuhmann (Gaudin et al., 1942) and elaborated by Garcia-Zuniga (Garcia-Zuniga, 1970), has been used extensively in the past. It is based on the assumption that the bubble concentration in the pulp remains constant (Sutherland, 1948).

Therefore, a solution of Equation 2.1 for \( n=1 \) and plug-flow reactor design is given as follows:

\[
R = 1 - e^{-kt} \quad \text{(Eq. 2.2)}
\]

where \( R \) represents the fractional recovery of the floatable species; and \( t \) is mean residence time of particles in the cell. This model provides a simplistic way for fractional recovery calculation of particles in the froth phase of the flotation system (Moys, 1978). Koh and Schwarz in their recent work (Koh and Schwarz, 2003; Koh and Schwarz, 2006)
concluded that the froth phase contributes significantly to overall recovery and that, due to its intrinsic complexity, it cannot be modeled straightforwardly as previously thought.

For the case of a perfectly mixed reactor design, fractional recovery derived from Equation 1, is given by:

\[ R = \frac{kt}{1 + kt} \]  
(Eq. 2.3)

This model has been typically used for calculation of the fractional recovery of particles in the pulp phase of a single flotation cell (Lynch, 1981; Weber, 1935)

The following expression for the overall flotation recovery calculation was recommended by Finch and Dobby (Finch, 1990):

\[ R = \frac{R_p R_f}{R_p R_f + 1 - R_p} \]  
(Eq. 2.4)

where \( R \) is overall fractional recovery; \( R_p \) is fractional recovery in the pulp phase; and \( R_f \) is fractional recovery in the froth phase, defined as the fraction of particles recovered after they are transferred into froth (Finch, 1990). Expression for the overall fractional recovery for a bank of \( n \) flotation cells can be easily derived (Finch, 1990; Harris, 1976):

\[ R = 1 - \left(1 - \frac{R_c R_f}{R_c R_f + 1 - R_c}\right)^n \]  
(Eq. 2.5)

In 2010, Do gave an expression for the fractional recovery in the froth phase, counting both recovery due to attachment and recovery due to entrainment (Do, 2010):

\[ R_f = \frac{d_{2-0}}{d_{2-f}} e^{-N_h f \left(1 - \frac{d_{2-0}}{d_{2-f}}\right)\left(\frac{d_1}{d_{2-0}}\right)^2} + R_w e^{-0.0325\left(\frac{p_2}{p_1} - 1\right) - 6.3 \times 10^4 d_1} \]  
(Eq. 2.6)
where \( d_{2.0} \) presents the diameter of the bubbles entering the froth phase, \( d_{2.f} \) presents diameter of the bubbles at the froth phase top, \( N \) is the number of particles attached to a single bubble; \( h_f \) is the froth height, \( R_w \) presents the maximal theoretical water recovery, \( \rho_3 \) is density of the water, and \( \rho_1 \) is particle density.

According to Sutherland, flotation rate can be expressed as a product of collision frequency between particles and bubbles \( (Z_{12}) \) and probability of flotation \( (P) \) as presented here (Sutherland, 1948):

\[
k = -Z_{12}P
\]

(Eq. 2.7)

Assuming independence of particle velocities from fluid flow, Abrahamson derived a turbulent collision model whose simplistic form is presented here (Abrahamson, 1975):

\[
Z_{12} = 2^{3/2} \pi^{1/2} N_1 N_2 d_{12}^2 \sqrt{\bar{u}_1^2 + \bar{u}_2^2}
\]

(Eq. 2.8)

where \( N_1 \) is a number of particles, \( N_2 \) is a number of bubbles, \( d_{12} \) is collision diameter or sum of radii of one bubble and one particle, \( \bar{u}_1 \) is the RMS velocity of particles, and \( \bar{u}_2 \) is the RMS velocity of bubbles.

As recommended by Tomlinson and Fleming, the total probability of flotation, \( P \), can be written as (Tomlinson, 1963):

\[
P = P_a P_c (1 - P_d) P_f
\]

(Eq. 2.9)

where \( P_a \) presents the probability of attachment (Ralston et al., 2002; Yoon and Mao, 1996), \( P_c \) presents the probability of collision (Luttrell and Yoon, 1992b), \( P_d \) is the probability of detachment (Honaker and Ozsever, 2003), and \( P_f \) presents the probability
of the successful transfer of bubble-particle aggregates to the froth phase (Do, 2010). Probabilities from Eq. 8 can be easily estimated by knowing several basic hydrodynamic and surface chemistry parameters for a given system. Some of these properties include particle size, density, zeta potential (a function of pH), contact angle (a function of mineral type and collector dosage), surface tension of the liquid phase (a function of frother concentration), pulp viscosity, and energy dissipation rate. This newly developed model was found to successfully simulate flotation results reported in the literature in both laboratory and industry (Do, 2010).

Computational Fluid Dynamics (CFD) modeling is another way to model the flotation process, and it is particularly useful for simulation of flotation systems in a state of local dynamic equilibrium between attachment and detachment processes (Koh and Schwarz, 2003). CFD modeling allows modeling of transport processes of free and attached particles in the flotation cell by combining source-sink balance equations with fundamental multi-phase flow equations.

For a flotation cell of known geometry and operating conditions, development of scale-up simulation from first principles is feasible when dependency between attachment and detachment rates and physical parameters are known (Koh and Schwarz, 2008). Koh and Schwarz found that the limiting factor is often not the collision or attachment rate, but insufficient bubble surface area for the attachment of all valuable particles present in the pulp.
2.7. HYDRODYNAMIC CHARACTERIZATION

As can be concluded from the previous section, any flotation process could be successfully predicted provided that fundamental chemical properties of the system and hydrodynamic conditions in the cell are known. Two of the most critical hydrodynamic properties of any flotation system are power input (or energy dissipation) and bubble size (or bubble size distribution).

2.7.1. Power consumption

Experimental determination of the spatial distribution of energy dissipation of three phase flotation systems proved to be a very challenging task (Yang, October 6-9, 2007). Several methods of flow velocity measurement of highly turbulent fluid flows have been noted in the literature, some of them being the Hot-Wire Anemometry Method, Electrochemical Method (Rubinstein, 1995), Laser-Doppler Velocimetry Method (Tiitinen, 2003), Particle Image Velocimetry Method (Brady et al., 2006; Zachos et al., 1996), and Multi-Hole Probe Method (Telionis, 2009). In general, most of the methods listed previously cannot accurately quantity fluid fields in the pulp with particle concentration higher than 10% . Even though energy dissipation varies throughout the flotation cell, the mean value of energy dissipation has been typically used for the evaluation of the flotation performance and as an input for the most recently developed flotation models (Do, 2010).

Newell and Grano conducted an experimental analysis of the influence of energy dissipation in the cell on the undistributed flotation rate constant for a given size fraction. For laboratory and pilot scale flotation systems, they have found that the $N^3D$ product,
which describes energy dissipation, can be used as a good flotation scale-up criterion. $N$ value in the product represents a rotational speed of the impeller and $D$ is an impeller diameter (Newell and Grano, 2006).

A number of researchers performed extensive studies on the effect of hydrodynamics, specifically power consumption, on the flotation process (Oyama, 1955), (Arbiter, 1965), (Arbiter et al., 1976), (Harris et al., 1983) and (Harris and Khandrika, 1985b). In their research they were considering the following variables: tank diameter, impeller diameter, liquid depth, distance of the impeller from the tank bottom, length and width of impeller blades, width of baffles, and size of solid particles. Flotation systems were described by dimensionless groups.

Arbiter 1965 found a decrease of aerated power consumption to non-aerated power consumption ratio with an increase of Air flow number. Based on Oyama and Endoh, the decrease in the power consumption in air-liquid systems is due to the lower density of the air-liquid mixture. However, the Power number calculated from the apparent density of the composite mixture was found to be higher than expected if the air were evenly dispersed throughout the system (Oyama, 1955).

Arbiter concluded that the air-liquid mixture in the vessel was not homogeneous, and in particular that the density in the zone around the impeller was lower than overall average density. One explanation for this phenomenon is a diminishing air concentration gradient of the air-liquid stream, flowing radially outwards from the impeller, with the distance from the impeller. Another mechanism could be the air re-entrainment, which is specific to the impeller-stator agitating mechanisms of standard conventional flotation.
cells. For mixers operating without vortexing in the fully turbulent regime, the power consumption is directly proportional to the liquid density:

\[ P = N_p \rho_l N^3 D^5 \]  

(Eq. 2.10)

where \( P \) is the overall power consumption, \( \rho_l \) is liquid density, \( N \) is impeller rotational speed, and \( D \) is impeller diameter.

Arbitrre 1968 found that, at fixed impeller speed, increase of impeller submergence of self-aerated flotation cells leads to increased aeration rate (Arbitrre et al. 1968). Experiments were conducted on a Fagergren laboratory flotation cell. For the same cell, he also found an increase of aeration rate with an increase of tank diameter to impeller diameter ratio. This relationship is attributed to a lower proportion of fresh air being drawn in the impeller zone, since more air is being recirculated in the larger tank. Also, in the same study, Arbitrre explored the difference in impeller blade design on power consumption. A wedged shaped impeller had a Power number twice that of an impeller with round posts.

Harris 1985 explained the increase of a degree of self-aspiration with the increase in impeller speed by the following proportion (Harris and Khandrika, 1985a):

\[ \text{Air suction} \propto N^2 \]  

(Eq. 2.11)

For self-aerated cells, a decrease in the power consumption was also found with the addition of the frother. It is known that frother addition reduces surface tension of the liquid phase and, in that way, enhances generation of smaller bubbles. Smaller bubbles are more readily entrained in the water and the fraction of recycled air increases. Therefore, one part of the air dispersed by the impeller is returned to the impeller zone by
the flow circulation that reduces the overall capacity of impeller to entrain fresh air into this zone. This consequently reduces the density in the impeller zone and lowers power consumption.

2.7.2. Gas dispersion

Among the hydrodynamic parameters, gas dispersion is considered to be the key. The intensification in research efforts to better understand gas dispersion and hydrodynamics in flotation cells has come about because of the growing need to design larger and more efficient flotation cells to treat the lower grade and more finely disseminated ores that are currently being mined (Sawyerr et al., 1998).

In this context, gas dispersion is defined as the dispersion of air into bubbles. It is well documented that the gas dispersion properties (e.g. bubble size distribution) in the flotation process have a direct influence on flotation performance (Schwarz and Alexander, 2006). This is understandable since the amount of created gas-liquid interfacial area affects particle collection kinetics.

Pursuing this notion, Gorain found that the flotation rate constant was not readily related to bubble size, gas holdup, or gas rate individually, but it was related to bubble surface area flux (Gorain et al., 1996). Bubble surface area flux is defined as the surface area of bubbles per unit time per unit cross-sectional area of a flotation cell.

2.7.3. Main variables for gas dispersion characterization

In the figure below (Figure 1.8), typical parameters defining gas phase flow in the flotation system are presented and described briefly.
Figure 2.8 Definition of gas dispersion parameters: a) gas superficial velocity, b) gas holdup, and c) bubble size and bubble surface area flux (n – number of bubbles per unit time; S – surface area per bubble), (after (Gomez and Finch, 2007)).

Generally, three hydrodynamic variables used for characterization of gas dispersion are superficial gas velocity ($J_g$), bubble size ($D_b$), and gas holdup ($\varepsilon_g$), which are used to obtain the derived parameter, bubble surface area flux ($S_b$). In case of mechanical cells, the Power number and Airflow number should also be included (Harris, 1976).

Superficial gas velocity, or simply gas rate, is defined as the volumetric flow rate of air per cross sectional area of the flotation cell ($J_g = Q_g/A$). Typical ranges in flotation systems are 0.5-2.5 cm/s depending on factors such as bubble size and slurry rheology.
Bubble size distribution is an important factor determining how well air is dispersed in a liquid phase. The Sauter mean diameter \( (D_b \text{ or } D_{32}) \) has been commonly used to describe the size of bubble population. The typical bubble size in flotation ranges from 0.5 to 2.5 mm (Gorain et al., 1995b).

Mean bubble size is usually used to describe a flotation system with a large number of bubbles of different sizes and can, in general form, be expressed as:

\[
D_{pk} = \sqrt[p=k]{\frac{\sum_{j=1}^{j=n} D_{ij}^p}{\sum_{j=1}^{j=n} D_{ij}^k}}
\]

For: \( pk=10 \) Mean number diameter
For: \( pk=20 \) Root mean square diameter
For: \( pk=30 \) Mean volume diameter
For: \( pk=32 \) Sauter diameter
For: \( pk=43 \) Mean mass diameter

(Eq. 2.12)

where \( n_i \) is the number of bubbles in the size class \( i \), \( D_i \) is a bubble diameter of certain bubble size class, \( f_i \) is a volumetric fraction of the size class \( i \). Similarly, \( n \) is the number of all bubbles analyzed in an image (sample size), \( D_j \) is measured equivalent bubble diameter, \( f_j \) is a volumetric fraction of a measured bubble \( j \).

Frothers have a pronounced impact on reducing bubble size up to a certain concentration, above which further addition of frother appears to have no effect (Finch, 1990). A decrease in coalescence is the accepted explanation of the decrease in bubble size with increasing frother concentration. After a certain concentration, which has recently been referred to as the critical coalescence concentration (CCC), a constant bubble size is obtained, implying that coalescence is now minimal. Different frothers reach CCC at different concentrations (Sweet, 1997), (Laskowski, 1998). A number of authors have tried to link the ability to reduce bubble size to surface tension. Lower
surface tension values are usually associated with higher frother concentration, resulting in smaller bubbles. On the other hand, Aldrich and Feng found that MIBC solutions with higher surface tension values than Dowfroth 200 solutions, produced smaller bubbles, which is counter-intuitive (Aldrich and Feng, 2000). From this it can be concluded that the bubble coalescence rate is not simply related to surface tension.

Gas holdup is the volume fraction of the mixture occupied by gas at any point in a flotation cell. It is the simplest gas dispersion parameter which combines the influences of both bubble size and gas rate. Gas holdup is directly influence by most of the flotation parameters. For the same gas flow rate, bubble population with larger number of smaller bubbles result in higher gas holdup, and vice versa. In 1993, Zhou found that frother type might also have an effect on bubble size and bubble size distribution (Zhou et al., 1993a). Similar conclusion was given by In 1996, Sam who measured the terminal velocity of a single bubble in different frother systems. Sam found that the terminal velocity was dependent on frother type (Sam et al., 1996).

Bubble surface area flux ($S_b$) is emerging as one of the most useful variables to quantify gas dispersion (Gomez, 2002). It is defined as the amount of bubble surface area delivered per unit time and cell cross-sectional area, and is given by

$$S_b = \frac{6J_g}{D_b}$$  \hspace{1cm} (Eq. 2.13)

where $J_g$ is the superficial gas velocity and $D_b$ is the Sauter mean bubble diameter of the distribution. Gorain in 1997 and Hernandez in 2003 found that the flotation rate constant is directly proportional to $S_b$. Increasing $J_g$, and/or decreasing $D_b$ can increase $S_b$ (Gorain et al., 1997; Hernandez et al., 2003). In practice, an increase in $J_g$ increases bubble size
giving a trade-off. This is one reason why there is an optimum $J_g$ in flotation (Dobby and Finch, 1991).

2.8. BUBBLE FORMATION IN THE FLOTATION SYSTEM

In general, bubble size (or bubble size distribution) in the flotation system is determined by the following three hydrodynamic processes:

- bubble formation in gas generator,
- bubble breakup, and
- bubble coalescence.

The latter two processes are directly governed by the local turbulence. To develop a better understanding of the role of gas bubbles in the flotation system, knowledge of the above-mentioned phenomena is required.

During the process of bubble formation, a fraction of the total energy supplied to the system is directly transformed into the free surface energy of newly created bubbles. A majority of bubbles created in flotation systems are generated in the zone between the impeller and stator blades, which is maximum energy dissipation zone of the cell. Here, air cavities that are initially formed at the low pressure region of the impeller blades are detached from the impellers blades edges and are carried into the high energy dissipation zone where the bubble breakup occurs. Bubble breakup is caused by the dynamic pressure and shear stresses on the bubble surface induced by shear flow and turbulence (Hinze, 1955).
Bubbles created in the high energy dissipation zone are carried along by the radial flow coming from the impeller and are scattered throughout the pulp contained in the body of the flotation cell (van't Riet and Smith, 1973). Thereafter, bubbles climb through the vessel due to their buoyancy but are also randomly pushed around by the existing turbulence in the cell.

In the flotation cell turbulence is the primary mechanism responsible for breakup of the bubbles that are initially created in the high energy dissipation zone (Kolmogorov, 1949). Only eddies that are of the similar length scale as the bubble size can break the bubbles. On the other hand, the large eddies can only transport the bubbles, while very small ones do not affect the bubbles (Olmos et al., 2001). In order to determine the actual size limit of eddies responsible for bubble breakage, Prince and Blanch (Prince and Blanch, 1990) conducted experiments to show that only eddies bigger than $0.2d$ are able to break bubbles of diameter $d$, while eddies bigger than $d$ can only move them. Therefore, based on Wu (Wu et al., 1998), bubble breakup rate depends on the frequency of collisions between bubbles and eddies of a similar size. Moreover, bubbles will break up into smaller bubbles only when the maximum hydrodynamic forces in the liquid (that tend to break up the bubbles) are larger than the surface tension force (that tends to stabilize the bubbles) (Angeli and Hewitt, 2000; Hinze, 1955; Kerdouss et al., 2006). This force balance is typically quantified by the liquid Weber number:

$$We = \frac{\rho u^2 L}{\gamma}$$  \hspace{1cm} (Eq. 2.14)
where $\rho$ is the density of the continuous phase, $L$ is the characteristic length, $\gamma$ is the interfacial tension, and $u_t$ is the velocity of eddies in the inertial sub-range of turbulent eddy spectrum, which can be calculated as (Kerdouss et al., 2006):

$$u_t = 1.4(\varepsilon d)^{1/3} \quad \text{(Eq. 2.15)}$$

where: $\varepsilon$ is the local turbulent dissipation energy and $d$ is the bubble size.

Bubbles are not stable and break up when the Weber number is larger than a critical value for a certain system (Walter and Blanch, 1986). Critical Weber numbers given in literature generally lie in the range from 1 to 5; no general agreement about a constant critical Weber number could be found. The critical Weber number typically used for modeling turbulent multiphase flows is in the range from 1.2 to 1.5 (Kerdouss et al., 2006; Lane et al., 2002; Lane et al., 2004; Rigby et al., 1997). Taking all this into account, Wu (Wu et al., 1998) has developed an expression for bubble break-up rate as follows:

$$N_{br} = C_{br}n \left( \frac{\varepsilon d}{d} \right)^{1/3} \left( 1 - \frac{We_{crit}}{We} \right)^{1/2} \exp \left( -\frac{We_{crit}}{We} \right), \quad We > We_{crit} \quad \text{(Eq. 2.16)}$$

where $C_{br}$ is the dimensionless coefficient.

The process of bubble formation described above defines the initial bubble size in the system, which is then altered due to the coalescence. Coalescence occurs when two or more bubbles collide and remain in contact for sufficient amount of time to allow the liquid in the thin film between them to drain and eventually, when the thin film reaches a critical thickness, to rupture (Kerdouss et al., 2006). Therefore, the rate of bubble
coalescence is related to the frequency of collisions between bubbles that are moving in the turbulent flow field (Prince and Blanch, 1990) with the velocity equal to the turbulent fluctuating velocity of eddies of equivalent sizes (Lane et al., 2002). The general equation for the coalescence rate derived by Wu (Wu et al., 1998), can be expressed as follows:

\[ N_{co} = C_{co} \eta_{co} d^2 u_t n^2 \cdot \frac{1}{\left(1 - \alpha_2^{1/3}\right)} \]  
(Eq. 2.17)

where \( C_{co} \) is the dimensionless coefficient, \( \eta_{co} \) is the coalescence efficiency, \( \alpha_2 \) is the gas volume fraction, and \( n \) is the bubble number density given by:

\[ n = \frac{\alpha_2}{\left(\frac{\pi}{6}\right) d^3} \]  
(Eq. 2.18)

By inclusion of the gas volume fraction in the final collision rate equation, the phenomenon of reducing of the mean free path between bubbles with increasing of the gas volume fraction could be accounted for. The coalescence efficiency for the stirred tank, as defined by Lane et al. (Lane et al., 2002), can be expressed as:

\[ \eta_{co} = e^{-\frac{W_{ef}}{\theta e^{-K_{co}d^3}}} \]  
(Eq. 2.19)

where, \( K_{co} \) is the fitting constant. The first part of the expression, proposed by Chester (Chesters, 1991), quantifies the ratio of the liquid film drainage time and the bubble interaction time. The second part of the expression accounts for the minimum amount of energy required to cause bubble deformation and liquid drainage since, during any bubble collision, the bubble approach velocity has to be sufficient to overcome the pressure rise due to the liquid being forced out between the bubbles (Lane et al., 2002). Therefore,
Coalescence is the phenomenon that significantly reduces the dispersion efficiency of the flotation system (Cho and Laskowski, 2002). Coalescence strongly depends on chemical parameters of the system, such as the liquid surface tension, volumetric fractions of the dispersed phase, and characteristics of the turbulent flow field (Chesters, 1991).

In the flotation systems, coalescence strongly depends on chemical parameters characteristic for the system, such as the liquid surface tension, volumetric fractions of the dispersed phase, and characteristics of the turbulent flow field (Chesters, 1991). For example, for liquids with low surface tension, the sizes of bubbles formed with one bubble generator type are always smaller than bubbles created in clear water. Due to their smaller sizes, the rise velocities are slower, which results in larger residence time, what is typically beneficent for the overall process performance.

After they have been generated, bubbles in the flotation system will be moved into different zones of the cell depending on the balance between the bubble buoyancy force and surrounding drag forces. Eventually, all bubbles leave the pulp by forcing themselves out through the pulp-froth interface and then, by complex processes that occur in the froth phase, a majority of the gas finally leaves the cell directly through the froth surface or is, in less extent, carried from the system by the recovered froth.

Hence, local bubble size and bubble size distribution in the flotation cell strongly depends on various operational, technical, and chemical factors whose effects on bubble size should be taken into account while designing or modeling a flotation process. Some of the factors that have the utmost effect on bubble size in the flotation systems are:

- the total gas intake,
- the total supplied energy,
- the physical and chemical properties of the liquid and solid phases,
- the impeller/stator design,
- the impeller relative location to the bottom of the cell, and
- the size and geometry of the cell.

Therefore, during the flotation process, the gas-liquid hydrodynamics in the system will strongly depend on the type of bubble generator used, so the choice of a proper impeller/stator assembly to satisfy the necessity of optimal gas dispersion is the key for the success and economy of the process. However, the mechanism of bubble generation in the flotation systems has not yet been explored in depth and differences between the bubble generation mechanisms of different flotation cells have not yet been addressed in the literature.

The size of bubbles obtained from the experiments, produced by any flotation device (impeller, sparger etc.), will be significantly different at different locations of the cell or column. Only when the bubbles are measured at different radial distances from the generation zone and device wall can a representative, average bubble size be obtained. Therefore, in order to obtain a representative bubble size, attention should be paid to the fact that the size of bubbles produced may be different in the radial and vertical direction. Larger bubbles are expected near the generator and center of the cell, while smaller bubbles can be found near the highest energy dissipation zone and close to the bottom of the cell (Zhou et al., 1993b).
Only few studies on spatial distribution of bubble sizes in the stirred vessels, including flotation cells, have been published so far ((Alves, 2002; Alves et al., 2002; Barigou, 1987; Barigou and Greaves, 1991; Barigou and Greaves, 1992; Gorain et al., 1995a; Kamiwano et al., 1998; Laakkonen et al., 2007; Laakkonen et al., 2005b; Lu et al., 1993)). On the other hand, a large amount of data is necessary for the successful assessment, support, and validation of the current and future models.

A lot of experimental and empirical methods for determining bubbles sizes have been used so far and include: photographic techniques, electroresistivity measurements (Yasumishi, 1986), dynamic bubble disengagement technique (Standish et al., 1991) and calculations using empirical or semi-empirical approach. The photographic technique is the most common method used so far.

2.9. BUBBLE SIZING METHODS

Techniques capable of measuring bubble size in multiphase flows are usually classified depending upon their operating principles.

2.9.1. Electroresistivity techniques

A commonly used method for bubble size measurement is the two-electrode conductivity probe (Yasumishi, 1986). This probe consists of two needle sensors that are mounted within a small vertical distance. Each of the sensors has a binary output signal that depends on which phase is in contact with the tip. As a bubble passes, the time delay between signals from the two sensors measures the time for the bubble to proceed from one probe tip to the other.
There is a possibility of variations in bubble frequencies recorded with this method. If such variations are statistically significant, this is indication of cross-talk and capacitive effects. This implies that the sensors are too closely spaced. Based on research done in the area of bubble sizing, the optimum spacing between two sensors is a function of the bubble frequency, the range of bubble chord lengths intercepted by the sensors, and the sensor size and geometry.

There are potential problems in applying this method to multiphase flow. Bubbles that are rising in a direction not aligned with the two probes lead to major errors, since it is possible that there is no delay in the signal from the two sensors. This seriously limits use of the probe in turbulent flow fields. To overcome this difficulty, some authors have developed multi-point probes (Raper et al., 1982). However, these probes can be utilized only in flows where the bubble size is at least 6 mm. In order to eliminate the effects of cross-talk between two closely positioned sensors an alternative method of acquiring the mean time delay between the signals from the two tips is to obtain it from the cross-correlation function between the signals (Zun, 1982). In summary, the two-point probe is an acceptable instrument for measuring bubble characteristics only if the bubbles are spherical and not too small.

2.9.2. Ultrasound technique

Ultrasound reflection technique offers a way to determine bubble size distribution in multiphase flows. It is known that bubbles have a resonance frequency that is inversely proportional to the radius of the bubble. This fact has been exploited for detection and estimation of bubble sizes by many authors (Hilgert and Hofmann, 1986; Luebbert et al.,
1987; Broering et al., 1991). Bubbles are excellent sound scatters and have a characteristic resonant frequency dependent on their sizes (Cathignol et al., 1988).

### 2.9.3. Optical techniques

#### 2.9.3.1. Optical fibers

Optical fibers exploit differences in the index of refraction of air/liquid phases and rely on the application of Snell's law at the probe-fluid interface. Depending on which phase is present at the probe tip, the light from the tip is reflected or refracted. The most common optical probe consists of two optical fibers fused and ground to a 45° angle with respect to the probe axis. The other ends of the fibers are free with one of them serving as an emitter and the other as a receiver. Light detection can be achieved with a phototransistor. The principle of detecting bubble size and velocity is identical to that of the two-point conductivity probe described above. Measurement of bubble size using optical probes is reported by Lee (Lee, 1984), and Saberi (Saberi et al., 1995). In general, an optical probe can be used only in transparent systems and at low gas holdup (volume fraction of gas). The success of the probe in discriminating between the phases depends on good contact between the probe tip and the bubble. Thus, if the bubble size was too small the probe would be unable to detect variations. The use of optical probes in a three-phase system is also considered problematic.

#### 2.9.3.2. The isokinetic collection probe

Another optical technique to measure bubble size is the isokinetic sampling probe. The term isokinetic refers to the condition in which bubbles are collected at uniform velocity regardless of their size. In this optical technique, bubbles are sampled and
collected into a capillary tube. The end of the capillary is funnel shaped which provides a uniform acceleration of sampled bubbles through the capillary. In the capillary, bubbles are converted into cylinders filling the capillary cross-section. A narrow collimated beam of light is directed through the capillary glass wall. The variation in intensities of the measured signal due to sequential shifts of gas and liquid slugs are recorded. The time elapsed between the detection of the two ends of a bubble is inferred from the signal. This, along with the known cross-sectional area of the capillary, can be used to estimate the bubble volume. Assuming that the bubble is a sphere, an equivalent spherical bubble diameter can be computed. Employment of this technique is reported for two-phase (gas-liquid and liquid-liquid) dispersions, in three-phase (gas-water-paper) pilot-scale flotation deinking cells, and in flotation columns with gas-water-coal systems.

Capillary size chosen for the method is a function of the smallest bubble needed to be detected. Bubbles smaller than the capillary diameter cannot be transformed into slugs and, consequently, they produce signal pulses of a small amplitude and width that cannot be reliably measured. On the other hand, reducing the capillary size further may cause bubble breakup at the point of bubble entry. Also, risk of the probe blockage increases as the capillary size decreases. Slug velocity is strongly dependent on purity of the capillary, decreasing as the build-up increases. In the case of slurries containing larger particles or slurries with high solids concentration, this limitation can become important.

A variant to the isokinetic sampling probe is introduced by Randall (Randall, 1989), which is shown in Figure 2.9. This method is often referred to as the University of Cape Town (UCT) bubble size analyzer.
Gorain and Franzidis, in one of their most cited research works, used a UCT bubble size analyzer to measure bubble sizes and distributions in the flotation cell under various operating conditions (Gorain et al., 1995a). They used four different impellers that were commonly used in industrial flotation cells, and operated the process at various combinations of air flow rate and impeller speeds around values recommended by the manufacturers. They followed approximately a possible flow pattern (circulation loop) generated by the impellers in the cell where they placed six measuring sites. They concluded that the mean bubble size was largest closest to the shaft above the impeller and smallest at the impeller discharge point as a result of coalescence.
Additionally, they found that the mean bubble size increases with the increase in air flow rate for all generator types, although the increase took place in different patterns as a function of the impeller type and speed.

Figure 2.10 Bubble break-up mechanisms at the entrance of the capillary tube of the UCT bubble size analyzer (after (Grau and Heiskanen, 2002)).

The UCT bubble sizing method has been used in a number of laboratory and industrial-scale studies, e.g., (Aldrich and Feng, 2000; Deglon et al., 2000; Gorain et al., 1995a; Gorain et al., 1997). A deterrent is that it contains a significant number of components which combine to render the system cumbersome for industrial use. It has also been observed that, under inappropriate conditions (e.g., capillary size, suction rate, etc.), bubble breakup may occur (Grau and Heiskanen, 2002). Four different bubble
breakage mechanisms, at the sampling end of the capillary tube, are presented in the Figure 2.10.

Bubble at the moment it touches the capillary, $t_1$, and its fate after the time period $t_2 - t_1$ are given in the figure. Bubble breakage can happen both inside of the capillary and at the entrance point. Large bubbles have typically, more chances to miss the capillary or to be broken by the capillary tip.

Only under very controllable conditions, when bubbles present in the sampled pulp are small and have narrow distribution, this system could be successfully employed. However, studies directed to establish the optimum conditions are scarce in the literature.

2.9.4. Imaging Techniques

The simplest approach for bubble sizing would seem to be the imaging technique. Studies in which images are used to size bubbles proliferate in the literature. In the most common setup, pictures of the dispersion are taken through windows installed in the vessel wall while real sizes are obtained by placing an object of known size (usually a ruler) in the focus plane. In the most commonly used methods bubbles are captured by high-speed CCD cameras due to their great precision (Soler et al., 2003). Automatic sizing through image analysis routines have not been extensively implemented due to the common practical and fundamental problems associated with this method.

One of the fundamental problems of this method, which is not frequently discussed, is the impact of the "inherently variant" distance between the focus plane and the bubble. The oscillating nature of the bubble motion causes the distance between the bubble and the focus plane to vary. Other fundamental problems include the influence of
optical conditions such as, for example, the lighting characteristics, which may vary during the sampling process. Other practical problems are well documented in literature: overlapping, blurring, bubble clustering, poor contrast, etc. Therefore, this is not a straightforward approach in an industrial-scale system and it was usually used in laboratory conditions. A variant of the imaging technique is to direct a sample of the bubbles into a viewing chamber and there expose them for imaging. Some of the most commonly used systems are:

2.9.4.1. Helsinki University of Technology (HUT) bubble size analyzer

This Helsinki University of Technology bubble size analyzer presented in the Figure 2.11 has been designed to work only for two-phase systems and it consists of a viewing chamber made of clear acrylic and small sampling tube. The bubble sampler is filled with water containing the same frother concentration as the liquid in the cell to reduce coalescence problems within the device. The high-magnification ratio used yields really shallow focal depth of fields. This facilitates the identification of overlapping structures.

In 2006, Grau and Laskowski used this method for measuring bubble sizes in the two solutions, with DF-200 and DF-1012 frothers (Grau and Laskowski, 2006b). They tested three locations in the cell prior to the final measurements, and selected a location near the pulp-froth interface (quiescent flow conditions) to be the most suitable location for this measurement method. A sampler tube of the HUT method is not able to identify bubbles that are not traveling upward, as is the case in the turbulent zone. Their assumption was that the collected sample from a single sampling location is
representative of the whole cell, which is probably not always the case under coalescence conditions.

**Figure 2.11** Helsinki University of Technology (HUT) bubble size analyzer and output image example (after (Grau and Heiskanen, 2005)).

2.9.4.2. McGill Bubble size analyzer

Hernandez-Aguilar in 2002 introduced a concept consisting of angled viewing chamber. This simple method along with a proper configuration of diffuse backlighting provided high contrast, while bubble overlapping and bubble blurring were reduced. The quality of the images was such that a simple computational routine could be applied to
analyze the images automatically (Hernandez-Aguilar, 2002a). This method is known as McGill bubble size analyzer and is presented in the Figure 2.12.

![Diagram of McGill bubble size analyzer](image)

**Figure 2.12** McGill bubble size analyzer (MBSA) and output image example (after (Gomez and Finch, 2007; Nesset et al., 2005)).

### 2.9.4.3. LTM-BSizer method

The technique, LTM-BSizer, is a new method developed to measure the bubble size distribution combining microscopy with digital image processing procedures (Rodrigues and Rubio, 2003). It utilizes a sampler to draw bubbles, rising in a column into a special viewing chamber, and exposes them to a digital camera after they have decelerated and stopped (Figure 2.13). This method offers a lot of advantages compared with other methods presently used for bubble sizing, including: more accurate measurement of very fine bubbles, reduced chances of coalescence and enlargement,
improved image quality (bubbles are in the focus), minimization of bubbles overlapping in the image, no requirement for transparent cell wall and no bubble breakage (Rodrigues and Rubio, 2003).

Figure 2.13 Details of the LTM-BSizer system (after (Rodrigues and Rubio, 2003)).

In some cases, liquid-gas mixture is withdrawn from the dispersion. However, this approach is applicable only when the liquid-phase is translucent. An alternative to size in systems with high solids concentration is to allow the bubbles to ascend into the viewing chamber by their natural buoyancy. Jameson and Allum first described this concept in 1984 (Jameson, 1984). Their work appears to be the first attempt to conduct a systematic survey of bubble sizes in industrial flotation cells. Ten years later, in 1994 Tucker, inspired by the approach of Jameson and Allum, developed a new bubble sampling technique also known as UCT bubble sampler (Tucker et al., 1994). The approach has
been used for sizing bubbles in many projects: in coal flotation (Yoon, 1986), for the visualization of bubble-particle (Zhou, 2000) and bubble-bitumen aggregates (Malysa et al., 1999a; Malysa et al., 1999b), in a laboratory-scale flotation cell (Grau and Heiskanen, 2002), in a pilot-scale cell processing a sulfide ore (Chen et al., 2001), and in an industrial-scale cell (Yianatos et al., 2001). However, the same difficulties associated with the imaging method (overlapping, blurring, bubble clustering, contrast, etc.) were present.

2.9.5. Drift flux analysis

The method of drift flux analysis, by which an average bubble size may be calculated from air flow rate and air hold-up determinations, has been applied to column cells and found to correlate well with photographic measurements (Dobby, 1988). Using bubble size values obtained from drift flux analysis, investigations with column cells have confirmed (Diaz-Penafiel and Dobby, 1994) that smaller bubbles deliver a higher rate constant. Gas holdup ($\varepsilon_g$) is the volumetric fraction (or %) of gas in the aerated slurry. Gas holdup, has been calculated by measuring the pressure difference between two tapping points. By changing of the measurement points we can obtain the $\varepsilon_g$ profile throughout the flotation cell height.

\[
P_A = \rho_{sl} \cdot g \cdot L_A \cdot (1 - \varepsilon_{gA}) \quad \text{(Eq. 2.20)}
\]

\[
P_B = \rho_{sl} \cdot g \cdot L_B \cdot (1 - \varepsilon_{gB}) \quad \text{(Eq. 2.21)}
\]

\[
\Delta P = \rho_{sl} \cdot g \cdot \Delta L \cdot (1 - \varepsilon_g) \quad \text{(Eq. 2.22)}
\]
\[ \varepsilon_g = 1 - \frac{\Delta P}{\rho_{sl} \cdot g \cdot \Delta L} \]  

(Eq. 2.23)

Another way to measure gas holdup is with a gas holdup sensor where two flow cells, an open cell to measure the conductivity of aerated slurry \((\kappa_{slg})\) and a siphon cell to measure the conductivity of deaerated slurry \((\kappa_s)\), are used simultaneously. With the two conductivities, Maxwell’s model for conductivity of a dispersion of spherical non-conducting objects (bubbles in this case) in a continuous conducting medium (slurry) is solved.

2.10. REFERENCES


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CHAPTER 3:
A NEW MODULAR PILOT-SCALE SETUP FOR HYDRODYNAMIC AND METALLURGICAL FLOTATION PERFORMANCE EVALUATION

Authors: Miskovic, Sanja; Luttrell, Gerald; Bratton, Robert
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3.1. ABSTRACT

Batch laboratory flotation cells are commonly used to obtain information regarding the maximum achievable mineral recovery and overall process kinetics for a range of chemical conditions. Unfortunately, considerable technical expertise and process experience is required to utilize batch data for the design and scale-up of continuous flotation systems. For this reason, continuous pilot-scale tests are often conducted to further validate and refine performance projections that are initially based on batch data. To facilitate this important type of work, a new continuous pilot-scale flotation circuit has been designed, constructed, and evaluated. This paper describes the flotation test circuit design and gives details about the automation and instrumentation systems used in the new circuit. Hydrodynamic and metallurgical results are also presented from several testing campaigns conducted with the pilot-scale flotation setup using either narrowly-sized hydrophobized glass spheres or rougher circuit copper concentrate from an industrial concentrator as floatable feed materials.
3.2. INTRODUCTION

The design, scale-up, and optimization of flotation systems is challenging and requires considerable technical expertise and process experience due to their multifaceted nature and complexity. As flotation systems are scaled to larger sizes, many of the major parameters controlling the process do not scale in proportion. For example, bubble sizes measured in large industrial flotation cells are often much larger than those measured in small laboratory cells, and the specific power input for industrial cells is up to ten times less than the power used in laboratory units. Moreover, the design of many industrial flotation cells differs significantly from one industrial scale to another due to a number of technical limitations, which additionally hinders direct comparison and simplistic scale-up.

Batch laboratory flotation cells are commonly used to obtain information about maximum achievable mineral recovery and grade, as well as overall process kinetics, for a range of chemical conditions. In many cases, the resultant data obtained from these semi-batch experiments cannot be directly used to design or scale-up industrial cells and continuous flotation systems without using empirical, semi-empirical or theoretical algorithms that account for inherent differences in the operational characteristics of cells of different scales. Nevertheless, information gained from this type of testing is usually necessary since it is neither practical nor feasible to directly conduct experiments on large industrial-scale cells.

The thrust of this work is to provide scientifically sound information that can help bridge a gap in knowledge and help correlate results gained through both industrial and
closely controlled laboratory conditions. Furthermore, this paper provides details about the design of a new continuous pilot-scale flotation circuit and its operational performance over a wide range of operating conditions.

3.3. METHODS

3.3.1. Experimental

For the measurement of major hydrodynamic and metallurgical parameters in pilot-scale mode, a new continuous processing circuit was designed and built (Figure 3.1). The new flotation circuit was specifically designed to reduce limitations associated with laboratory testing and to approach the operating conditions of industrial-scale flotation cells as closely as possible. Further attention was paid to ensure that the system was easy to operate and control and was producing accurate, reliable and reproducible results. The circuit consisted of four major mechanical components: Dorr-Oliver® flotation cell (which was the central component of this pilot-scale setup), air blower, slurry sump, and slurry pump. Additionally, a new state-of-the-art control and data acquisition system was developed and implemented into the new circuit.

3.3.1.1. Flotation Cell

In order to allow testing of different cell geometries and to accommodate various flotation cells in the future, a new modular, stainless-steel, 0.8 m³ flotation tank was fabricated. The modular tank was composed of five interchangeable segments including three body rings with different heights and internal and external launders. Each launder segment was designed with three interchangeable launder rings with heights of 5, 10 and 20 cm, so that effects of different froth depths can be investigated without changing the
effective pulp volume. The cell was also designed to accommodate up to a maximum of eight vertical baffles. Furthermore, new Dorr-Oliver® wetted parts (impeller, stator, and impeller shaft) were designed and fabricated to fit the new tank.

3.3.1.2. Flotation Circuit

A new flotation circuit was designed and built to facilitate continuous flotation tests. For this purpose, a 1 m³ sump was used for slurry circulation where both froth concentrate and tailings streams were collected, mixed together, and then pumped back to the flotation cell through a feed line. It was discovered during initial testing that a large volume of air was introduced into the sump by the concentrate and tailing return lines, and also by the sump propeller mixer, which formed an undesirable mineralized froth atop the mixing sump. Several modifications were made to the slurry sump to reduce froth accumulation and to maximize the material circulation through the circuit. For example, a physical partition was added to the sump to minimize the free surface available for froth accumulation. Furthermore, a custom overhead washing system was installed to facilitate efficient bubble breakup and floatable particle release from the accumulated froth phase. For this application, five small-capacity spiral-jet spray nozzles were installed. By utilizing these nozzles, a fraction of the feed stream was able to be bypassed directly through the nozzles back to the sump. The use of feed slurry for froth breakup avoided the problem of unwanted dilution of the feed that would occur if fresh spray water were added. Finally, all stream lines were configured to enable simultaneous full-stream sampling when necessary.
Figure 3.1 Diagram of the continuous pilot-scale flotation circuit.
As shown in Figure 3.1, the test circuit was equipped with various sensors, including a pressure transmitter, level sensor, conductivity probe, magnetic slurry flow meter, two gas mass flow meters, and a torque meter. To enable data logging, a state-of-the-art control and data acquisition system was developed and installed. The final experimental setup, which also included two bubble sampling systems (ex-situ and in-situ), allowed continuous data collection of multiple flotation parameters such as local bubble size, local gas holdup, local and global superficial gas velocity, froth depth, feed flow rate, and power input.

For the measurement of local bubble sizes, only in-situ bubble sampler developed at Virginia Tech was used. The bubble sampling system was composed of a monochrome GEViCAM GP-21400 CCD high-speed camera and LED light, which were mounted inside two aluminum watertight enclosures that faced each other. The gap width between the light source and the camera was adjustable from 6 to 50 mm. Back illumination of bubbles was achieved using an Advanced Illumination SL2420 LED red spot light, which was pulsed with an Advanced Illumination S4000 strobe controller. A ground glass diffuser was installed in front of the LED light to facilitate more uniform light distribution. The in-situ system made it possible to examine local bubble size distributions at various horizontal and vertical positions within the flotation pulp.

Instrumentation used in this study for the measurement of superficial gas velocity and local gas holdup have been described elsewhere by Nesset and Gomez (Gomez and Finch, 2007; Nesset et al., 2006). In addition, a shaft mounted strain gauge was chosen
for the purpose of power measurements due to its intrinsic capability to directly, reliably, and precisely measure the torque of a rotating shaft. Measurements were performed with an Advanced Telemetrics International’s strain gauge radio telemetry system. A supporting split collar assembly was mounted directly on the shaft, which housed the battery, amplifier, radio frequency transmitter, and all associated electronics. Since there are no mechanical contacting parts to wear out, the system required very little maintenance and was easy to install.

The operation of the pilot-scale flotation circuit required the coordination of multiple functions, which was made possible by an integrated automation system. The main application of the automated system was to provide exact and reproducible execution of the following functions: level control, feed flow, agitation, and aeration rate regulation, and overall data acquisition. The process control system was also designed to provide for real-time monitoring of all input signals from the various sensors. A graphical trend display allowed for constant monitoring of all key process parameters. The visualization, operation, and monitoring of process parameters was managed through the easy-to-use Rockwell Automation FactoryTalk View Studio ME interface, shown in Figure 3.2, while RSLogix™ 500 ladder logic software was used for the control and management of input/output (I/O) tags.

The FactoryTalk Historian software allowed reliable data capture and creation of accurate on-line records of all process parameters. In this way, a single computer, which was also acting as a main human machine interface (HMI), managed multiple operations and allowed the analysis and monitoring of all data from a single location.
Figure 3.2 Snapshot of the main control screen (FactoryTalk View Studio ME).

3.3.2. Materials

Three-phase flotation tests were conducted using two different floatable materials selected for this study: (i) four different size classes (203, 119, 71, and 35 µm) of SPHERIGLASS® soda-lime glass spheres and (ii) copper rougher concentrate from an industrial concentrator. Additionally, MICROSil® CGS ground silica was used as an artificial gangue material during copper flotation tests. Typical particle size distributions of glass spheres and ground silica used in this study are shown in Figure 3.3. Selected physical and chemical properties of these particles are given in Table 3.1. FloMin F 500 (methyl isobutyl carbinol - MIBC) frother was used through all two-phase and three-phase tests with copper concentrate, and Nalco V-20373M glycol-based frother was used during glass sphere flotation testing. Monosized glass beads were hydrophobized with dodecylamine (DDA), while FloMin C 3505 collector was used for copper flotation.
Figure 3.3 Particle size distributions of SPHERIGLASS® solid glass spheres (■ – 35 µm; ▲ – 71 µm; ♦ – 119 µm; ■ – 203 µm) and MICROSIL® CGS ground silica (●).

Table 3.1. Basic physical and chemical properties of selected material.

<table>
<thead>
<tr>
<th>Material Type</th>
<th>Mean Particle Size (µm)</th>
<th>Specific Gravity</th>
<th>Particle Shape</th>
<th>Hardness (Mohs)</th>
<th>Composition (%)</th>
</tr>
</thead>
<tbody>
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<td>SPHERIGLASS 1922</td>
<td>203</td>
<td>2.5</td>
<td>Spherical</td>
<td>6</td>
<td>72.5 13.7 9.8 0.2 3.3</td>
</tr>
<tr>
<td>SPHERIGLASS 2227</td>
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<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>SPHERIGLASS 2530</td>
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<td></td>
<td></td>
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<td>35</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>MICROSIL CGS</td>
<td>131</td>
<td>2.65</td>
<td>Subangular</td>
<td>7</td>
<td>99.7 0.01 0.01 0.023 &lt;0.01</td>
</tr>
</tbody>
</table>

3.4. EXPERIMENTAL PROCEDURE

3.4.1. Experimental conditions

Measurements in two-phase (water/air) and three-phase (water/air/particles) were carried out in order to extensively evaluate both hydrodynamic and metallurgical cell performance over a wide range of operating conditions. The impeller tip speeds chosen
For this study varied in the range from 3 to 8 m/s, while aeration rates of up to 2.2 cm/s were utilized. For testing in continuous conditions, feed flow rates were varied from 90 to 500 l/min, which provided mean residence times of 1 to 8 min. Frother concentrations ranged from 0-14 ppm for MIBC and from 0-7 ppm for Nalco V-20373M frother.

3.4.2. Hydrodynamic measurements

The two-phase tests were performed to investigate both gas dispersion properties and overall power consumption of the flotation cell. During these tests, the cell was operated as a batch reactor. All tests were run in both coalescing and non-coalescing conditions (Grau and Laskowski, 2006a). Once the agitation and aeration rates were set, the cell was run at least three minutes before measurements were performed. For each set of operating conditions, all variables were recorded for approximately 10 minutes and then averaged. To describe the overall variability of each measured variable over a given time period, the standard deviation of each time series was determined and reported with error bars. For derived parameters, such as bubble surface area flux, specific power, and theoretical liquid residence time, the error was calculated by applying the propagation of uncertainty rule (Meyer, 1975).

In-situ bubble sampling methods was used for local bubble sampling. During the two-phase testing, bubbles were acquired from six different positions within the cell to allow assessment of the spatial variation of bubble size distributions. For three-phase tests, bubbles were sampled at a single location positioned in the impeller jet region for this particular study. Bubbles detected at this location are believed to be the most significant since considerable bubble-particle collisions occur in this high energy dissipation region.
The field of view and image capture rate for the in-situ bubble sampling method were 17.5 x 13 mm and 10 frames per second, respectively. Additionally, a gap size between the light and camera of the in-situ bubble sampler was set to 20 mm during two-phase tests and 10 mm during three-phase tests with glass particles. During copper flotation tests, the gap width was reduced to 3 mm and bubble sampling was supported by a peristaltic pump, which enabled isokinetic sampling of the pulp, containing more than 25% w/w of solids.

Approximately two hundred bubble images were taken for each operating condition. Images were randomly chosen from each data set and analyzed with BubbleSEdit image analysis technique (Zabulis et al., 2007). The bootstrap standard deviation (Efron and Tibshirani, 1993) for each analyzed bubble sample was calculated.

Six image examples acquired with the in-situ bubble sampler in both two- and three-phase systems are shown in Figure 3.4. Details of operating conditions and sampling location are given for each presented image. As mentioned before, a detailed description of experimental procedures for the measurement of superficial gas velocity and local gas holdup are given by Nesset and Gomez (Gomez and Finch, 2007; Nesset et al., 2006). Both local gas holdup and local superficial gas velocities were measured in the quiescent zone of the cell, approximately 35 cm below the froth-pulp interface and at the mid-distance between the impeller shaft and the tank wall.

Also, since information about the power input during flotation can give insight into the gas dispersion characteristics of the flotation cell, the torque was measured under both unaerated and aerated conditions. Each measurement was repeated three times and
the average power consumption and standard deviation were determined for each set of test conditions.

**Figure 3.4** Examples of images recorded with a new in-situ bubble sampler: a) 5 m/s agitation rate, 1.37 cm/s aeration rate, 14 ppm MIBC, water/air mixture, 10 mm gap width, without sample pumping; b) 5 m/s agitation rate, 1.31 cm/s aeration rate, 14 ppm MIBC, water/air mixture, 3 mm gap width, with sample pumping; c) 5 m/s agitation rate, 1.30 cm/s aeration rate, 4 ppm V-20373M, water/air/35 µm glass spheres mixture, 10 mm gap width, without sample pumping; d) 5 m/s agitation rate, 1.30 cm/s aeration rate, 4 ppm V-20373M, water/air/203 µm glass spheres mixture, 10 mm gap width, without sample pumping; e) 8 m/s agitation rate, 1.88 cm/s aeration rate, 8 ppm MIBC, water/air/Cu rougher concentrate mixture, 3 mm gap width, with sample pumping; f) 5 m/s agitation rate, 1.25 cm/s aeration rate, 8 ppm MIBC, water/air/Cu rougher concentrate mixture, 3 mm gap width, with sample pumping.

At the end, suspending capabilities of the Dorr-Oliver flotation cell were investigated under unaerated conditions with slurry containing 5% w/w of 203 µm SPHERIGLASS® glass spheres. Slurry samples were taken for three different agitation rates from eleven different locations in the cell. The vertical sampling plane was positioned 38 cm radial distance from the center of the cell. The slurry sample was collected with peristaltic pump in the rate of 10 l/min.
3.4.3. Metallurgical measurements

As described in previous section, continuous flotation tests were performed using four different size classes of glass spheres. Glass spheres were chosen due to their narrow particle size range, consistent chemical and physical characteristics, and reasonable price. By utilizing this material for the metallurgical performance characterization, more controllable experimental conditions were achieved. During initial batch flotation tests, the maximum theoretical recovery for all selected particles was determined for a range of collector and frother dosages. Based on preliminary laboratory testing, standardized collector and frother dosages for the pilot-scale tests were established. It was found that 4 ppm V-20373M frother and 30±2 g/t of DDA in a 5% ethanol solution were optimal dosages for glass particle conditioning. In order to minimize froth effects and bubble loading limits, the froth depth was maintained at 2.5 cm and the weight percentage of solids in the pulp was kept constant at a relatively low value of 0.5% w/w.

The basic test procedure used in all glass flotation tests was as follows. The collector was added to the water/particles mixture and circulated through the flotation circuit. After about ten minutes of conditioning time, the frother was added directly to the feed sump. The feed flow rate was then adjusted to provide the desired residence time of slurry within the cell. Once the aeration and agitation rates were adjusted, a minimum of three residence times were allowed for the system to reach the steady-state, after which all parameters were measured. Also, the timed cuts of all three streams were collected simultaneously at the sampling points indicated in Figure 1. Samples of tailing, concentrate, and feed were weighed, filtered, and dried, and then dry weighted. Mass balancing was performed to determine the mass and water recoveries.
The second group of continuous pilot-scale flotation tests was conducted using a rougher concentrate sample from an industrial concentrator. The original slurry sample was decanted and the remaining wet material, with approximately 5% w/w of water, was used in tests. In this procedure, MICROSIL® CGS ground silica was used as an artificial gangue material. To reproduce conditions comparable to industrial, both the rougher concentrate and ground silica were added in proportions to achieve approximately 25% w/w of total suspended solids, of which approximately 4% w/w was floatable material. The generated slurry was treated with 25±1 g/t of FloMin C 3505 collector and additionally 8±0.5 ppm of MIBC frother was added to the system. Furthermore, in order to evaluate the effect of froth on flotation performance, three froth depths were tested (i.e., 2.5, 5 and 7.5 cm).

The basic test procedure used in all copper flotation tests was as follows:

- Ground silica and rougher copper concentrate were added in the slurry sump and circulated through the system.
- FloMin C 3505 collector was then added and the system was run for approximately 15 minutes before frother addition.
- The froth depth was set to the desired height and the blower was activated to introduce air.
- After the target aeration and agitation rates were set, the system was run for at least three residence times.
- Tailings, concentrate, and feed samples were collected.
The samples were weighted, filtered, and dried, and the dry sample was weighted and sent for assay analysis.

3.5. RESULTS AND DISCUSSION

3.5.1. Hydrodynamic investigation in two-phase

To better understand the physical processes that govern flotation, accurate measurements of all main process parameters under controllable conditions are required. Generally, there are three hydrodynamic parameters used to characterize gas dispersion properties within a flotation cell: superficial gas velocity \( J_g \), Sauter mean bubble diameter \( D_{32} \), and gas holdup \( \varepsilon_g \). In the case of aerated stirred reactors, the Froude number, Aeration number, and the ratio of unaerated to aerated power consumption can be used to describe gas dispersion and mixing efficiency (Harris, 1974; Harris, 1976; Harris and Mensah-Biney, 1977).

Among the gas dispersion parameters, bubble size is perhaps the most important since it defines the free surface area over which solid particles and bubbles interact, which directly controls the system hydrodynamics and overall flotation performance. In order to have a better insight into the overall gas dispersion efficiency, bubbles should be screened at multiple locations within the cell. The spatial variation of bubble sizes in a two-phase system (water-gas mixture) was determined using the in-situ bubble sampling method. In the current study, bubble images were collected at six different locations within the flotation pulp. Figure 3.5 shows bubble size distributions and mean bubble sizes obtained under non-coalescing conditions (14 ppm MIBC). During this batch test, the impeller tip speed and aeration rate were set at 5 m/s and 1.37 cm/s, respectively. For
each location, one representative image was chosen and shown in Figure 3.5. Number mean \((D_{10})\) and Sauter mean \((D_{32})\) bubble diameters with corresponding bootstrap standard deviations and number frequency distributions are reported for each location. The fraction of the total gas volume contained in a certain bubble size class is also given.

As shown in the figure, bubble populations varied significantly at different vertical and radial distances from the impeller/stator assembly. At this particular operating condition, a proportionally larger fraction of larger bubbles within a screened population was recorded in the region above the impeller/stator gap. On the other hand, bubble populations containing the least number of larger bubbles were detected in the impeller jet region (zone with the highest energy dissipation) and close to the bottom of the cell. The number of bubble-particle collisions occurring in the high energy dissipation zone of the cell is strongly affected by the size of the bubbles in this zone. For that reason, bubble sizes obtained from the impeller jet region were reported and used for the Aeration number calculations.

Sauter mean and number mean bubble diameters (with corresponding bootstrap standard deviations) and local gas holdup values are presented in Figure 3.6 as a function of different aeration and agitation rates. As can be seen from the plot, the Sauter mean bubble diameter increased proportionally with the local gas holdup under different aeration conditions and for constant impeller tip speed. As a result of this relationship, and the fact that it can be measured directly in real-time, gas holdup can be used as a direct indication of local gas dispersion properties in the cell.
Figure 3.5 Sampling locations with representative bubble images and corresponding mean bubble diameters and bubble size distributions at 5 m/s impeller tip speed and 1.37 cm/s superficial gas velocity. ■ – Fraction of the total gas volume; ▲ – Number frequency distribution.
For a given constant gas flow rate, more intense agitation resulted in an increase in gas holdup, while bubble size decreased. This trend can be explained by the fact that smaller bubbles rise at lower velocities and have longer gas residence times, which is directly reflected through the increased gas holdup. Figure 3.6 also shows a linear increase of gas holdup with respect to superficial gas velocity measured for various combinations of agitation and aeration rates under non-coalescing conditions.

![Graph showing Sauter and number mean bubble diameters and gas holdup as a function of superficial gas velocity.](image)

**Figure 3.6** Sauter and number mean bubble diameters and gas holdup as a function of superficial gas velocity.

To provide a better insight into mixing, gas dispersion, and energy dissipation properties within the cell, measurements of power consumption were performed over a wide range of operating conditions. In the case of aerated systems, the total power consumption is always lower than that in unaerated systems due to great influence of
aeration on power transfer from the impeller to the fluid (Nienow, 1977). Van’t Riet and Smith (van't Riet and Smith, 1973) explained that the formation of gas cavities behind the impeller blades and difference in the fluid density under aerated and unaerated conditions is the reason for the power reduction. Depending on the type of impeller and aeration rates used, the ratio of the power consumption in aerated \((P_a)\) and unaerated \((P_0)\) conditions is usually in the range from 30% to 100%. This ratio determines the actual power input of the flotation operation and gives insight into the gas dispersion characteristics of the impeller.

Aeration number \((N_Q)\) is used to describe the nature of the gas–liquid flow within the cell and can be calculated using following expression: \(N_Q = Q_g/ND^3\), where \(Q_g\) represents the total volumetric gas flow rate, \(N\) is the agitation rate, and \(D\) is the impeller diameter. Figure 7 shows the \(P_a/P_0\) ratio as a function of the Aeration number at different agitation rates (i.e., 3 to 8 m/s impeller tip speed). The drop in aerated power consumption over the unaerated power consumption with an increase of Aeration number can be observed for all agitation rates. The \(P_a/P_0\) ratio of the agitation mechanism used in this study ranged from 0.5 to 1.0, depending on the agitation and aeration rates used.

As suggested by Nienow (Nienow, 1977), knowledge of the minimum agitation rate for complete dispersion of introduced gas, \(N_{CD}\), is necessary for better understanding of various flow patterns that occur in the cell. In Figure 3.7, this value is represented by the agitation rate, where the \(P_a/P_0\) ratio is minimal for a certain aeration rate. This operating condition is characterized by a fully dispersed, bubbly fluid flow pattern, without impeller flooding (Pangarkar et al., 2002). However, recent research (Hall et al., 2005) has suggested that gas holdup at this condition cannot support sufficient gross
recirculation of the bubbles throughout the cell, and that the optimal aeration rate for a certain agitation rate is shifted towards the left of the $P_a/P_0$ minimum.

The jump in the $P_a/P_0$ values observed for the 3 and 4 m/s impeller tip speeds represents a transition of the overall flow pattern from “loaded” to “flooded” conditions. A similar trend cannot be observed for higher agitation rates due to constraints associated with the maximum aeration number that could be achieved in the available setup. One of the explanations of the power consumption increase during the loaded-flooded transition is the upward distortion of the gas cavities formed behind the impeller blades, which result in an increase of the total blade surface area in contact with the liquid. Below the

Figure 3.7 Ratio of the power consumption in aerated ($P_a$) and unaerated ($P_0$) conditions as a function of the Aeration number at different impeller tip speeds: ♦ - 8 m/s; ■ - 7 m/s; ▲ - 6 m/s; ● - 5 m/s; × - 4 m/s; and + - 3 m/s impeller tip speed.
loaded-flooded transition, the curve has the familiar “S” shape as a result of gas cavity formation behind the impeller blades (Harnby et al., 1997).

Figure 3.8 Experimental results of the loaded-flooded transition at four different conditions: ○ – water-gas mixture; □ – water-gas and 7 ppm MIBC; ■ – water-gas and 14 ppm MIBC; ▲ – water-gas-solids, 25% w/w, and 8 ppm MIBC.

To further investigate gas dispersive patterns under extreme operating conditions, a critical aeration rate, when flooding conditions at the top of the tank surface could be observed, was recorded for a constant agitation rate. The flooding conditions were characterized by a wavy, boiling like motion of the free liquid surface in the central tank region, close to the impeller shaft. Since this type of measurement represents an attempt to assign a numerical value to a visually observed phenomenon, which is strongly
subjective, three or more measurements were performed randomly during both falling and increasing aeration rate.

Average critical aeration rates were used for the calculation of the Aeration number. Figure 3.8 summarizes the results obtained from the series of tests conducted with the water-gas only and water-gas-frother (7 and 14 ppm MIBC) mixtures. The same measurements were performed during the three-phase operation (25% w/w of solids in slurry) and are also reported in this plot. The aeration and agitation rates at the loaded-flooded transition are expressed in terms of the Aeration number and Froude number ($Fr$). The Froude number is given as $Fr = N^2D/g$, where $g$ represents the acceleration due to gravity. The standard deviation of three measurements, representing the uncertainty in the visual observations, is also reported through error bars in the plot. The uncertainty generally increased as the aeration and agitation rates increased due to more turbulent free surface at the top of the cell, therefore creating an indistinct loaded-flooded transition.

An effect of agitation rate on particle suspension performance of the flotation cell was investigated under unaerated conditions. The solids content of slurry samples, taken from eleven different locations, was determined for three different impeller tip speeds (4.5, 5.5, and 6.5 m/s). The particles used for this investigation were monodisperse 203 μm glass spheres, and their concentration in the slurry was 5% w/w. Each test was repeated three times and both average solids concentration and corresponding error bands are determined (Figure 3.9). As can be seen from the Figure 3.9, three distinct zones, with significantly different particle concentration profiles, are formed within the cell and this trend is observed for all three operating conditions. As expected, the zone with the
highest particle concentrations (> 5% w/w) was observed at the bottom of the cell (app. 0 to 10 cm from the cell’s bottom). In the middle zone, which is considered to be a highly-turbulent zone of the cell, the particle concentration profile was constant. Lastly, in the top, quiescent zone of the cell, a gradual decrease in particle concentration was observed, with concentrations as low as 0.4% w/w detected close to the launder lip (110 cm). The increase in agitation rate increases particle of-the-bottom suspension and increases particle concentration in the cell, which ultimately creates more favorable conditions for bubble-particle encounter in the high-turbulent zone of the cell.

Figure 3.9 Particle concentration profile along the vertical cross-section of the cell at three different impeller tip speeds. Dotted line - the concentration profile of ideally suspended mixture; Full line – average particle concentration profile; Shaded region – error band.
3.5.2. Hydrodynamic and metallurgical investigation in three-phase - glass spheres

A series of flotation tests were run independently on all four selected glass sphere samples to determine the effect of particle size, liquid residence time, aeration rate, and agitation rate on overall material recovery. Nine different combinations of aeration and agitation rates were tested for each particle size with one condition being repeated three times randomly in a testing sequence.

In this way, the reliability of the testing procedure was examined by evaluating the standard deviation of repeated measurements. Absolute errors of all measured mass recoveries fell in the narrow range from 0.5% to 1.8%, which confirmed the test reproducibility.

Figure 3.10 shows the total mass recovery as a function of particle size, obtained through pilot-scale testing, under the same chemical conditions (4 ppm V-20373M; 30±2 g/t DDA) and 1.5±0.4 min residence time. For reference, the maximum achievable recoveries for each particle size obtained through preliminary laboratory testing, under the same chemical conditions as in pilot-scale experiments, are also plotted.

The drop off in recovery for fine particles may be attributed to decreased efficiency of bubble-particle encounter associated with smaller particle sizes. On the other hand, the relatively poorer recovery of coarse particles can be explained by an increased probability of bubble-particle detachment and decreased stability of the bubble-particle aggregates due to the absence of fine particles.
Figure 3.10 Material recovery as a function of particle sizes for different operating conditions. ■ – 6 m/s; ▲ – 5 m/s; and ● – 4 m/s impeller tip speeds; full line – 1.7 cm/s; large dashed line – 1.3 cm/s; small dashed line – 0.8 cm/s; and dotted line – 0.6 cm/s aeration rate.

Figure 3.11 shows the material recoveries for different particle sizes (35, 71, and 119 µm), as a function of theoretical liquid residence time. For fine and intermediate particle sizes, recovery increased as residence time increased over all operating conditions. This result was expected since, for the identical operating conditions, the probability of bubble-particle attachment increased with residence time increase. In contrast, the recovery of coarser particles was not as significantly affected by the increased residence time, which is reflected by the flattening of the recovery-residence time curve for coarser particles. This trend suggests that coarse particle recovery can be challenging even when operating at relatively long residence times.
Figure 3.11 Material recovery as a function of theoretical liquid residence times and particle size at three different operating conditions. ■ – 35 µm; ▲ – 71 µm; and ♦ – 119 µm; full line – 6 m/s, 1.7 cm/s; dashed line – 5 m/s, 1.3 cm/s; and dotted line – 4 m/s, 0.6 cm/s agitation and aeration rates.

The data summarized in Figure 3.11 also shows a relationship between flotation recovery and parameters such as aeration and agitation rate. For all particle sizes, higher aeration and agitation rates resulted in higher material recoveries. It is well documented that the gas dispersion conditions in the flotation cell have a direct influence on flotation performance (Schwarz and Alexander, 2006). This is understandable since the amount of created gas-liquid interfacial area affects particle collection kinetics. Pursuing this notion, researchers have found that the flotation rate constant was not readily related to bubble size, gas holdup, or gas rate individually, but was related to bubble surface area flux (Gorain et al., 1997; Gorain et al., 1998b). Bubble surface area flux ($S_b$), calculated as $S_b$
\( = 6J_g/D_{32} \), is therefore one of the most useful parameters for quantifying gas dispersion effectiveness in a flotation cell (Finch et al., 2000).

Figure 3.12 shows the relationship between flotation rate constant \((k)\) and bubble surface area flux for constant froth depth and identical chemical conditions. The error bars in the plot correspond to calculated uncertainties of bubble surface area flux based on measurement errors of the superficial gas velocity and bootstrap standard deviations of the Sauter mean bubble diameter for each tested condition.

**Figure 3.12** Flotation rate constant as a function of bubble surface area flux for constant froth depth (2.5 cm) and four different particle sizes. ■ – 35 µm; ▲ – 71 µm; ♦ – 119 µm; and ● – 203 µm.

As shown, a strong correlation between the flotation rate constant and bubble surface area flux was observed for all particle sizes tested. Over the range of operating
conditions and a single particle size class, the correlation was found to be roughly linear, which is in agreement with previous publications (Deglon et al., 1999; Finch et al., 2000; Gorain et al., 1999; Gorain et al., 1998b; Heiskanen, 2000; Vallebuona et al., 2005).

Figure 3.13 illustrates the effect of specific power input ($P^*$), defined as a power-to-volume ratio, on the flotation performance. Specific power was estimated from the direct torque readings and the effective liquid volume, while taking into account the overall gas holdup in the cell. Three sets of operating conditions were chosen from the test matrix and corresponding results are presented in the plot (4 m/s and 0.82±0.05 cm/s; 5 m/s and 1.34±0.05 cm/s; and 6 m/s and 1.74±0.05 cm/s impeller tip speed and global superficial gas velocity).

![Figure 3.13](image)

**Figure 3.13** Flotation rate constant as a function of specific power input for four different particle sizes. ■ – 35 µm; ▲ – 71 µm; ♦ – 119 µm; and ● – 203 µm.
Based on the results shown in Figure 3.13, the flotation rate constant increased with an increase of specific power input for each of the four particle sizes tested. However, specific power increase mostly affects intermediate and fine particles, which is reflected through notably stronger $k-P^*$ relationship. These results are in good agreement with the findings of previous studies (Deglon, 2005; Ralston et al., 2007; Ralston et al., 2010; Schubert, 2008).

Sauter and number mean bubble diameters and local gas holdup, measured in the impeller discharge stream, are shown as a function of specific power input in Figure 3.14. The selected operating conditions are the same as those presented in the Figure 3.13.

**Figure 3.14** Sauter ($D_{32}$) and number mean ($D_{10}$) diameters and gas holdup as a function of specific power for three selected operating conditions (4 m/s and 0.82±0.05 cm/s; 5 m/s and 1.34±0.05 cm/s; and 6 m/s and 1.74±0.05 cm/s impeller tip speed and global superficial gas velocity) and 35 µm particle size. ▲ – $D_{32}$; ♦ – $D_{10}$; □ – gas holdup.
As can be seen from the figure, the increase in the specific power input has negligible effect on the bubble sizes, which is well reflected through insignificant changes in both Sauter and number mean diameters. On the other hand, the gas holdup increases significantly with increase in power input, suggesting that more bubbles with similar diameters have been generated. It can be concluded that the power input has a positive effect on the number of bubbles created in the cell, which consequently increases the probability of bubble-particle attachment, increases overall carrying capacity, and therefore increases the flotation rate constant (Figure 3.13).

3.5.3. Hydrodynamic investigation in three-phase - copper concentrate

To further explore the capabilities of a new pilot-scale flotation circuit, an additional series of tests were performed with the copper concentrate and ground silica, totaling 25% w/w of solids, in order to achieve more realistic industrial-like conditions. Hydrodynamic measurements of all major gas dispersion parameters and torque measurements were performed for a series of operating conditions and under identical chemical conditions.

Figure 3.15 shows the relationship between local gas holdup values and local superficial gas velocities measured in the quiescent zone of the cell. Error bars in the plot reflect the standard deviation of a time series for gas holdup and the standard deviation of multiple readings for superficial gas velocity. The test data shows the gas holdup increased from 8% to 13% as the superficial gas velocity increased from 0.8 to 2 cm/s.

A gradual broadening of confidence intervals around estimated values can be observed for both measured parameters as the aeration rate increased. This is largely due
to more dynamic response of all measurement devices as a result of more heterogeneous gas distribution and wider bubble size distributions generated within the cell at the higher gas rates.

**Figure 3.15** Local gas holdup as a function of local superficial gas velocity for the following global superficial gas velocities. ♦ – 0.88±0.02 cm/s; ■ – 1.20±0.02 cm/s; ▲ – 1.54±0.02 cm/s; ● – 1.88±0.03 cm/s; and X – 1.97±0.05 cm/s.

Figures 3.16 and 3.17 illustrate the effect of both aeration rate and power input on Sauter mean bubble diameter measured in the high energy dissipation zone. For the four selected impeller tip speeds (i.e., 5, 6, 7, and 8 m/s) and the range of superficial gas velocities (0.8 to 2 cm/s), the Sauter mean bubble diameter ranged from 0.65 to 2.55 mm.
Figure 3.16 Sauter mean bubble diameter as a function of local superficial gas velocity at four different impeller tip speeds. ♦ – 5 m/s; ■ – 6 m/s; ▲ – 7 m/s; and ● – 8 m/s.

This range of mean bubble sizes, which is wider than bubble size distributions previously reported in the literature (Fuerstenau, 2007; Laskowski et al., 2003; Sawyerr et al., 1998), can be attributed to increased precision achieved with the in-situ bubble sizing method by which up to 98% of all recorded bubbles in an image were detected and included in the analysis. In addition, the results of the automated image analysis were revised and manually corrected to ensure that all bubble clusters and non-spherical bubbles, composed of big bubbles and/or gas slugs, were counted and included in the bubble sizing analysis. In the current study, this approach significantly increased the calculated $D_{32}$ values compared to previous studies. However, it is believed that the Sauter mean bubble diameters determined in this way present more accurate estimation of
real gas dispersion conditions in the cell, characteristic for both homogeneous and heterogeneous regimes (low and high superficial gas velocities), typically occurring in mechanical flotation cells.

![Figure 3.17 Sauter mean bubble diameter as a function of specific power input for the following global superficial gas velocities. ♦ – 0.88±0.02 cm/s; ▭ – 1.20±0.02 cm/s; ▲ – 1.54±0.02 cm/s; ● – 1.88±0.03 cm/s; and x – 1.97±0.05 cm/s.](image)

Figures 3.16 and 3.17 also indicated that the Sauter mean bubble diameter was strongly affected by both aeration rate and specific power input. The decrease in mean bubble diameters, or shifting of the bubble size distribution toward smaller bubble sizes, was observed for reduced superficial gas velocity or increased specific power input. From the data given in Figure 3.16, it can be seen that mean bubble sizes gradually increased over the range of superficial gas velocities evaluated in this study. On the other hand,
flattening of the $D_{32}-P^*$ trend can be observed for all aeration rates tested, as shown in Figure 3.17. These results suggest that there is a minimum energy input needed to achieve an optimal bubble size distribution for a given constant aeration rate.

The bubble size decrease, which is caused by the increase in power input at the constant aeration rate, results in a creation of larger bubble surface area, as shown in the Figure 3.18.

![Graph of bubble surface area flux as a function of specific power input for following superficial gas velocities.](image)

**Figure 3.18** Bubble surface area flux as a function of specific power input for the following superficial gas velocities. ♦ – 0.88±0.02 cm/s; ■ – 1.20±0.02 cm/s; ▲ – 1.54±0.02 cm/s; and ● – 1.88±0.03 cm/s.

On the other hand, the increase in bubble surface area flux is not directly reflected in better flotation performance, as would be expected (Figure 3.19).
Figure 3.19 Flotation rate constant as a function of bubble surface area flux for the four superficial gas velocities. ♦ – 0.88±0.02 cm/s; ■ – 1.20±0.02 cm/s; ▲ – 1.54±0.02 cm/s; and ● – 1.88±0.03 cm/s.

One of the explanations for this observed phenomenon is the detrimental effect of high agitation rates on flotation kinetics in the pilot-scale cell. In other words, even though the selected range of impeller tip speeds (from 5 to 8 m/s) is analogous to the agitation rates found in industrial conditions, which is believed to be the best approach for the flotation scale-up, other negative effects observed at higher agitation rates (7 and 8 m/s) decrease the flotation performance in pilot-scale. For example, it is observed that at 8 m/s impeller tip speed impeller generates a strong discharge stream, which, due to the short radial distance from the impeller to the wall, collides intensively to the cell wall. After the impact, this high intensity stream has sufficient energy to rise directly to the cell top and causes “geysering” effect at the froth surface. For these reasons, both higher
detachment probability, due to extremely turbulent conditions in the cell, and lower froth recovery, due to significantly disturbed dynamics of the froth zone, lead to decrease in flotation performance. Therefore, care must be taken during the scale-up process since not all parameters relevant to flotation scale in proportion. It is important to note here that different scales of flotation cells, used through the scale-up procedure, have different range of optimal operating conditions and their own characteristic limitations.

Figure 3.20 shows a positive relationship between the overall water recovery and gas holdup for all operating conditions. In this case the froth depth was kept constant and shallow (2.5 cm).

![Figure 3.20](image) Copper grade in the concentrate stream as a function of water recovery and gas holdup in the cell.
As seen in the Figure 3.15, high aeration and agitation rates are responsible for generation of larger number of bubbles in the pulp, which is reflected through the higher gas holdup values. Therefore, increase in the number of bubbles entering the froth zone from the pulp directly increases the overall water recovery. Additionally, observed froth surface instability due to both “geysering” effect and high gas influx may contribute to the increase in the water recovery. Finally, the increase in water recovery results in decrease in the concentrate copper grade, as shown in the Figure 3.20.

Based on determined Sauter mean bubble diameters, the bubble surface area flux was computed using both locally measured and global superficial gas velocities and obtained results are summarized in Figure 3.21.

![Figure 3.21 Local versus global bubble surface area flux.](image)
The error bars in this plot represent the uncertainty of $S_b$ estimation, which accounts for measurement errors of local superficial gas velocity, total gas flow rate, and bootstrap standard deviation of $D_{32}$.

For most of the operating conditions, the bubble surface area flux values estimated from global superficial gas velocities appear to overestimate the bubble surface area flux. Small relative errors from 0.2% to 2% were observed for low aeration rates (0.8 to 1.3 cm/s), while relative errors form 1% up to 14% were found for higher aeration rates (1.5 to 2 cm/s). One explanation of this finding is that, at higher aeration rates, a part of the introduced gas contained in large bubbles rises directly from the impeller to the froth zone. As a result of this heterogeneous gas distribution within the cell, global superficial gas velocities, calculated by dividing the overall gas flow rate with the cell cross sectional area, will give slightly larger estimations than superficial gas velocities obtained by local measurements at a half radial distance from the impeller shaft.

3.6. **CONCLUSION**

A fully-instrumented 0.8 m³ pilot-scale flotation circuit was developed for the purpose of providing performance data that can be more readily utilized for the engineering design, scale-up, and optimization of industrial flotation circuits. This highly flexible system enabled measurement, monitoring, and control of a number of hydrodynamic and metallurgical parameters over a wide range of operating conditions comparable to those found in industrial scales. Preliminary tests indicate that the continuous circuit can be successfully operated with slurry containing more than 30% w/w of solids.
Performance of the system was evaluated in both two- and three-phase tests, utilizing either glass spheres or copper mineral concentrate. Several important conclusions obtained through hydrodynamic and metallurgical testing in pilot-scale are listed below:

- Bubble populations vary significantly at different vertical and radial distances from the impeller/stator assembly under the same operating condition, and the nature of this variation strongly depends on the operating condition.
- Measurement of power consumption in flotation cells allows better insight into the gas dispersion properties of the cell.
- Ratio of the aerated to unaerated power plotted as a function of the aeration number gives important information about the minimum agitation rate necessary for complete dispersion of the introduced gas.
- The first sign of the transition of the overall flow pattern from “loaded” to “flooded” conditions can be easily observed through the increase of the aerated to unaerated power ratio as a result of aeration rate increase under constant agitation rate.
- During the operation of mechanical flotation cells, three distinct zones, with significantly different particle concentration profiles, are formed within the cell.
- The increase in agitation rate increases coarse particle of-the-bottom suspension and increases their concentration in the cell, which ultimately creates more favorable conditions for bubble-particle encounter in the high-turbulent zone of the cell.
For all particle sizes, higher aeration and agitation rates resulted in higher material recoveries and for fine and intermediate particle sizes recovery increased as residence time increased over all operating conditions.

A correlation between the flotation rate constant and bubble surface area flux is observed for all glass particles tested, while the nature of this correlation strongly depends on the size of the particles.

The power input has a positive effect on the number of bubbles created in the cell, which increases the probability of bubble-particle attachment, increases overall carrying capacity, and therefore increases the flotation rate constant.

The Sauter mean bubble diameters measured in this study ranged from 0.65 to 2.55 mm, which is wider than bubble size distributions previously reported in the literature.

Larger mean bubble diameters are obtained due to increased precision achieved with the in-situ bubble sizing method by which up to 98% of all recorded bubbles in an image were detected and included in the analysis.

The decrease in mean bubble diameters can be achieved by the decrease in superficial gas velocity and increase in the specific power input.

Flattening of the $D_{32}$-$P^*$ trend can be observed for all aeration rates tested, which suggests that there is a minimum energy input needed to achieve an optimal bubble size distribution in the cell for a given constant aeration rate.

As a result of this heterogeneous gas distribution within the cell, the bubble surface area flux values estimated from global superficial gas velocities overestimate the bubble surface area flux.
• Care must be taken during the flotation scale-up process since not all parameters relevant to flotation scale in proportion.

• Different scales of flotation cells, used through the scale-up procedure, have different range of optimal operating conditions and their own characteristic limitations.

Data obtained using the pilot-scale system can be used as a baseline for advanced modeling, control, and optimization of flotation processes. With its functional versatility, the system can be easily adapted to almost any process condition and, in that way, provide valuable process-related knowledge necessary for the development of successful scale-up strategies and more efficient flotation cells.

3.7. ACKNOWLEDGEMENT

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3.8. REFERENCES


CHAPTER 4: COMPARISON OF TWO BUBBLE SIZING METHODS FOR PERFORMANCE EVALUATION OF MECHANICAL FLOTATION CELLS

Authors: Miskovic, Sanja; Luttrell, Gerald
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4.1. ABSTRACT

A new in-situ optical bubble sampling method capable of collecting representative samples through different regions of a flotation cell has been developed. The new system was compared to the standard McGill ex-situ bubble sampling method. All experiments were carried out using a fully automated, pilot-scale 0.8 m³ Dorr-Oliver® flotation cell. Bubble images were taken from multiple locations within the cell. The cell was operated as a batch reactor under various operating conditions by altering impeller tip speed, gas flow rate, and frother concentration. Two methods of image analysis were also evaluated, i.e., a new software package called BubbleSEdit and the standard McGill/ Northern Eclipse software. BubbleSEdit is a template matching technique that analyzes overlapped bubbles and bubble clusters, which results in detection of more than 90% of all bubbles in an image. Bubbles observed with the in-situ sampling method appeared to be larger than bubbles recorded with the McGill ex-situ method. Furthermore, it was found that the mean bubble size determined by the McGill/Northern Eclipse bubble sizing method was smaller than the BubbleSEdit values.
4.2. INTRODUCTION

The dispersion of gas into bubbles and their rise due to buoyancy are very important fundamental phenomena that contribute significantly to the hydrodynamics of the flotation process. Therefore, accurate gas dispersion data are required to better understand the physical processes governing flotation. The importance of bubble size on flotation efficiency has first been recognized by Nevett in 1920. He concluded that air must be completely atomized in the pulp in order to reach optimal separation conditions (Nevett, 1920). Similarly, using high-speed cinematography in his experiments, Bennett (1958) concluded that, for a constant air supply rate, flotation rate increases by producing a larger number of smaller bubbles. This relationship is understandable since the amount of created gas-liquid interfacial area directly affects particle collection kinetics. In other words, bubble size governs the surface area over which solid particles and bubbles interact. Furthermore, this free surface area contributes significantly to system hydrodynamics and overall flotation performance. Research efforts to better understand gas dispersion in flotation cells have intensified because of the desire to design larger and more efficient flotation cells that are necessary to treat the lower grade and more finely disseminated ores currently being mined (Sawyerr, 1998).

The bubble size distribution (BSD) in flotation systems generally depends on operational variables such as aeration rate, pulp surface tension, impeller rotation speed, as well as design variables such as impeller and stator configurations and cell geometry. Unfortunately, current flotation models developed from first principles cannot predict bubble size distributions from these basic input variables. On the other hand, by combining computational fluid dynamics (CFD) and population balance models (PBM),
the calculation of local BSDs becomes feasible. In order to make them reliable tools for flotation machine design and optimization, predicted BSDs need to be validated experimentally by performing local bubble size measurements (Ranade, 2002). For that reason, a reliable and accurate method for local bubble size measurement, in both turbulent and quiescent regions of the flotation machine, needs to be developed for laboratory-, pilot- and industrial-scale conditions.

4.2.1. Bubble size measurements

One of the first laboratory measurements of bubble size was performed by Rodger (1956). At that time, bubble sizing was labor intensive and fully manual process. Over the last six decades, the bubble sizing methodology has been improved through the implementation of various new techniques that, typically, brought a greater level of automation to the process. While automation has increased the speed of the analysis, it produced undesirable side effects, such as sensitivity to image noise and ignoring of large bubbles. For this reason, a considerable amount of manual work is still required to assure correctness of obtained results.

Techniques capable of measuring bubble size in multiphase flows are typically classified depending on their operating principles. A number of experimental and empirical methods for determining bubbles sizes have been used so far and include photographic (Grau, 2002; Tucker, 1994; Yianatos, 1988), laser diffraction (Couto, 2009), interferometric laser imaging (Kawaguchi, 2002), electro-resistivity (Yasumishi, 1986), acoustic (Pandit, 1992), dynamic bubble disengagement (Standish, 1991), ultrasonic pulse transmission (Stravs, 1986), and many other techniques. Calculations using empirical or semi-empirical approaches have also been employed (Dobby, 1988).
Of these, the photographic technique is the most commonly used method for bubble sizing today. The availability of high speed digital cameras and advanced and automatic image processing techniques has made photographic measurements much easier. Most commonly, images of bubbles are taken by high-speed CCD cameras through windows installed in the vessel wall (Soler, 2003). One of the fundamental problems of this technique is the impact of the inherently variant distance between the focal plane and bubbles in motion. Another problem with this technique is that a small change in optical conditions, such as lighting, significantly affects image quality. The minimum detectable bubble size depends on factors such as camera resolution, CCD cell sensitivity, and optics type. Also, to eliminate blur caused by bubble motion, image exposure time has to be kept very short.

Another important aspect of photographic bubble sizing is image analysis. The image analysis of bubbles captured in multiphase flows, with high concentrations of bubbles and/or suspended particles, is challenging due to bubble overlap and bubble clustering (Junker, 2007). Also, bubbles present in the optical path between the camera and the focal plane, and bubbles behind the field of focus, significantly degrade image quality. For that reason, image analysis methods available today often fail to clearly distinguish individual bubbles from bubble swarms, and are limited to transparent, two phase (gas/liquid) mixtures with low bubble concentrations (<10% by volume). Moreover, photographic bubble sizing is challenging in industrial-scale systems and is more often used in simplified laboratory evaluations. The sampling technique must be rugged to withstand harsh environmental conditions. This is especially important when the equipment has to be deployed in industrial plants, where erosive/corrosive conditions
are present and flotation cells may be wet, dirty, vibrating and difficult to access. Additionally, the bubble measuring system should be compact and light in order to provide the least interference with the natural fluid flow in the cell.

The size of bubbles obtained in flotation experiments also vary significantly at different locations within a flotation cell. A good measure of overall bubble population should be obtained by multiple measurements at different points within the cell. Bubble sizes obtained this way should be weighed based on their relative position to the high energy dissipation zone (impeller/stator zone) and averaged to get an overall mean bubble size. Bubble sizes differ in radial and vertical directions throughout the flotation cell, and larger bubbles are expected near the impeller top and center of the cell, while smaller bubbles can be found near the highest energy dissipation zone and close to the bottom of the cell (Zhou, 1993). Therefore, care must be taken to ensure that the method used to obtain samples of bubbles within the pulp is not biased.

An *ex-situ* bubble sampling method is the most commonly used photographic method for evaluation of flotation cells. In this approach, bubbles are directed from the pulp zone of the flotation cell into an external viewing chamber where they are exposed for imaging. Some of the most commonly used *ex-situ* bubble sizing systems in flotation cells are include the University of Cape Town (UCT) bubble size analyzer (Tucker, 1994), Helsinki University of Technology (HUT) bubble size sampler (Grau, 2002), McGill bubble size analyzer (MBSA) (Hernandez-Aguilar, 2002), USM bubble size sampler (Yianatos, 2001), and LTM-BSizer method (Rodrigues, 2003). Unfortunately, the bubble sizes measured in various flotation systems, which include different scales and cell types, show a notable spread. The reason for this deviation has not yet been
adequately addressed in detail in the technical literature. It is not clear whether the problem is caused by chemical variations among the systems, cell configuration (cell size, type, and geometry), or problems with the experimental sampling and imaging techniques.

This article describes a new photographic in-situ sampling method for determining BSDs within a flotation cell. The in-situ method has been tested using a 0.8 m³ Dorr-Oliver® pilot-scale flotation cell operating as a batch two-phase (air-water) system. The data obtained from the in-situ method have been compared with measurements obtained using standard MBSA ex-situ method. Both methods have been used to measure local BSDs at different locations within the flotation cell. This article highlights the advantages and limitations of each method and discusses major sources of discrepancies between results obtained with these two methods.

4.3. EXPERIMENTAL

4.3.1. Experimental setup

The bubble sizing work was conducted using a 0.8 m³ Dorr-Oliver® flotation cell that was designed and installed in the Mining and Minerals Engineering Laboratory at Virginia Tech. The modular stainless-steel tank was composed of five interchangeable segments, including three body rings with different heights and two top launders (internal and external). For the bubble sizing tests, the tank was configured with an internal launder with the overflow lip positioned 1.12 m from the tank floor. New Dorr-Oliver® wetted parts (impeller, stator and impeller shaft) were designed and fabricated to fit the new tank. The tank had a diameter (T) of 1.02 m and impeller diameter (D) of 20.3 cm, which gave an overall T/D ratio of 5.
As shown in Figure 4.1, the test cell was equipped with various sensors, including a pressure gauge, level sensor, conductivity probe, mass flow meter, and torque meter. To facilitate data logging, a state-of-the-art control and data acquisition system was developed and installed. The final experimental setup, which also included two compact high-speed CCD cameras, was designed to continuously log data regarding bubble size, gas holdup, superficial gas flow rate, and energy input. The cell was operated as a batch two-phase system by forced introduction of gas into a water-frother mixture. The frother selected for the experiments was methyl isobutyl carbinol (MIBC). Two sampling systems were utilized for bubble sample collection, i.e., a standard ex-situ MBSA bubble viewer and a new in-situ bubble sampler.

**Figure 4.1** Experimental setup used for the bubble sizing experiments.
4.4. IMAGE ACQUISITION

A compact GEViCAM GP-21400 1.5 mega pixel Giga-bit Ethernet high-speed CCD camera (34x34x68 mm) was used to collect bubble images. The required magnification was achieved using a FUJINON HF9HA-1B 9 mm f/1.4 fixed focal lens. The monochrome camera was operated with the maximum resolution of 1392x1040 pixels. For measurements with the MBSA method, bubble images were captured as they passed through the external viewing chamber using a camera frame rate of 23 fps and 1/250 s shutter speed. The depth of field (DOF) was estimated to be 3.8 mm for the selected optical conditions (observed bubble distance from CCD detector of 7 cm, selected aperture of f/3.6, and circle of confusion of 10 µm). This DOF was also measured and confirmed directly by using the Edmund Optics DOF 5-15 depth of field target.

Details of the MBSA method and corresponding bubble sampling procedures have been presented previously by Hernandez-Aguilar (Hernandez-Aguilar, 2002). Based on findings from an initial series of tests with the MBSA method, a couple of new features were added to this standard system to improve image quality and make it more suitable for the tests that had to be performed. In order to allow isokinetic bubble sampling, a peristaltic pump was used to continuously draw a bubble sample to the MBSA viewing chamber at a rate of 1 and 2 l/min for the quiescent and turbulent zones of the flotation cell, respectively. To further improve sampling conditions, the suction tube was aligned with the flow direction by adding either 45° or 90° elbows to the end of the sampling tube. For example, a 90° elbow was chosen when sampling the primary horizontal jet from the impeller. The halogen light source provided with the original setup
was replaced with a more uniform 5500 K daylight white panel LED light (3000 lumens) to provide brighter backlighting. Finally, the standard glass on the viewing chamber was replaced with frosted glass to provide uniform light diffusion. In this way, image quality and bubble detection accuracy was improved significantly.

Alongside the standard MBSA ex-situ bubble sampler, a new in-situ photographic method was utilized. A second GEViCAM GP-21400 CCD camera equipped with a FUJINON HF9HA-1B 9 mm lens was used with the in-situ system. Back illumination of bubbles was achieved using a compact, bright field, Advanced Illumination SL2420 LED spot light, which was pulsed with an Advanced Illumination S4000 strobe controller. A red LED light (660 nm) was chosen since the long-wavelength light travels longer in scattering media and was preferable for observation of contaminated liquids and multiphase flows. Also, the bubble images obtained this way had improved border definition. A ground glass diffuser was also installed in front of the LED light to provide uniform light distribution. The system was set up to provide 10-100 µs long LED light pulses, which defined the total image exposure time. Using this exposure time, bubbles moving in a fluid flow of up to 15 m/s could be clearly captured. The camera provided 1392x1040 pixel monochrome images at a frame rate of 10 fps. The DOF was estimated to be 3 mm for the selected optical conditions (distance from focused bubbles to CCD detector was 6 cm, lens aperture was fixed at f/5, circle of confusion was 10 µm and a 1 mm extension ring was added). An Edmund Optics 5-15 depth of field target was also used to confirm the expected DOF value.

For the in-situ system, both the high-speed camera and LED spot light were mounted inside aluminum watertight enclosures that faced each other (bottom left, Figure
1). The gap width between the light source and the camera was manually adjustable from 6 to 50 mm. As the gap size decreased, more light passed through the moving bubble swarms. A smaller gap also decreased the number of bubbles viewed per single image, which provided less overlap and fewer counting errors for dense bubble populations. A gap size of 20 mm was selected and used through all experiments, which provided about two times larger depth of the view volume than the largest expected bubble diameter. This gap width was found to be large enough to allow entrance of larger bubbles into the measurement field and small enough to produce good bubble images. In this way, the influence of the sampling method on measured bubble sizes was reduced to a minimum.

Due to its compact and cylindrical design, the in-situ system had a relatively small effect on fluid flow within the cell. The field of view (FOV) inside the flotation cell obtained with in-situ method was 17.5 x 13 mm and the system was capable of measuring bubble sizes in the range from 50 µm to 10 mm. In comparison, the FOV for the ex-situ MBSA method was 31 x 23 mm, which gave a maximal detectible range of bubble sizes of 90 µm to 20 mm.

4.4.1. Experimental procedure

Measurements of gas holdup and superficial gas rate were conducted for all experiments performed using the pilot-scale Dorr-Oliver® flotation cell. A detailed description of the procedures and instrumentation used in these experiments has been given elsewhere (Gomez, 2007; Gomez, 2002; Hernandez-Aguilar, 2002). Torque measurements were also performed for all experiments. The cell operating conditions were modified by changing the impeller speed and gas flow rate. Experiments were carried out at two impeller tip speeds (4 and 5 m/s) and three gas flow rates (20, 40, and
60 m³/h) giving overall superficial gas flow rates of 0.69, 1.37, and 2.06 cm/s, respectively. In order to run tests under non-coalescing conditions, 14 ppm of MIBC frother was used in all experiments. This concentration (14 ppm) was higher than the critical coalescence concentration (CCC) for the MIBC frother (Nesset, 2007).

Two bubble sampling methods, the standard MBBA ex-situ method and new in-situ method, were used simultaneously for local bubble sampling. Since BSDs varied depending on the sampling location, bubbles were acquired from three different positions in the cell for both methods (locations 1, 3, and 4) and from three additional locations using the in-situ method (locations 2, 5, and 6). All sampling locations are presented in the Figure 2. The locations were selected to allow bubble sampling from the major fluid streams in the cell and were distributed evenly to allow determination of the spatial variation of BSDs within the flotation machine.

All sampling positions, except location 6, were arranged vertically at a radial distance of 15 cm from the stator ring. Location 6 was positioned directly above the impeller-stator gap, 3 cm above the stator ring. Location 1 was located in the quiescent zone of the cell, while sampling points 3 and 4 were located in the high energy dissipation zone. Sampling point 3 was positioned 2.5 cm above the stator ring, sampling point 4 was 1 cm below the stator ring, and sampling point 5 was positioned 6 cm below the stator ring. Bubbles detected at location 4 were indicative of bubbles carried by the main high dissipation energy jet leaving the impeller. Based on existing turbulent flotation models, location 4 was considered most significant since considerable bubble-particle collisions occur in this region.
During testing, the flotation cell was run for approximately 10 min to ensure that the operation had reached steady-state. After steady state was reached, bubble sampling was initiated and bubble images were acquired from different sampling locations for the next 20 min. In order to obtain a representative sample, approximately 200 bubble images were taken by each technique at each location and for each operating condition. Bubble images were randomly chosen from each data set and processed with two available image analysis techniques. The total number of images analyzed was a function of the bubble size and the total number of bubbles per each image. Figure 4.3 gives schematic illustration of complete bubble sizing procedure.
Figure 4.3 Bubble sizing procedure.
4.4.2. Image processing

After sampling was completed, the captured images were typically pre-processed and then analyzed using image analysis software. Even though this procedure provided an efficient alternative to manual counting, miscounting or misclassifications occurred frequently. The main reason for this problem was the presence of overlapped bubbles and bubble clusters in the image. Most software packages used for image analysis of multiphase flows fail to discern individual bubbles from bubble clusters. This problem became more significant when a wide range of bubble sizes was present in the bubble sample. In particular, the presence of large bubbles in the image greatly biased the BSD since they covered a significant portion of the image, overlapped smaller bubbles, created bubble clusters, had an irregular shape, and touched image edges. For this reason, a more sophisticated processing technique had to be used to reduce these image processing errors.

For the analysis of all image sets acquired by MBSA method, a customized Empix Imaging Northern Eclipse 6.0 software package interfaced with an Excel/Visual Basic user interface was used. This software offers several advantages including full automation, fast analysis of a large image sets (2 s/image), and direct export to Excel spreadsheets where statistical analysis is automatically performed. On the other hand, this technique also has a number of inherent shortcomings including high sensitivity to the selected threshold value, lack of visual confirmation of analysis accuracy, incorrect reporting of noise as bubbles, and the exclusion of overlapped objects (Figure 4.4). In most cases, only 5 to 40% of all objects in one image were recognized as bubbles using this commercial package.
Figure 4.4 Output from the Northern Eclipse image analysis software (shaded regions represent bubble clusters excluded from the analysis).

To improve the image analysis procedure, a new software package known as BubbleSEdit (BSE) was used in the current study. BSE employs a template matching technique where manually selected bubbles are used as prototypes for automatic detection of remaining bubbles. Prototypes are exhaustively compared for visual similarity with all image neighborhoods. To detect bubbles independently of their size, the selected prototype is iteratively resized in order to evaluate occurrences of the prototype in a wide range of scales. The normalization of cross correlation facilitates the detection of prototypes despite global image intensity changes (Zabulis, 2007). In its automatic analysis mode, the software can automatically detect circular bubbles in the images and use them as prototypes for the prototype based bubble detection. Depending
on the image characteristics, the overall detection accuracy of the automated sizing procedure typically exceeds 80% (Figure 4.5).

Figure 4.5 BSE image analysis output (white circles represent bubbles detected in the first software run after 10 s).

For images obtained by the ex-situ MBSA method that contain mostly circular bubbles, the detection accuracy can be as high as 95% using BSE. On the other hand, bubble images collected by the in-situ sampling method usually contain a number of irregular shaped bubbles, which makes image analysis a challenging task. This irregularity can be explained by the fact that under high Reynolds numbers in the turbulent zone of the cell, bubbles smaller than 1 mm are generally spherical, while bubbles above 2 mm are non-spherical in shape (Clift, 1978). With the BSE technique, the estimated time for automatic analysis of one image ranges from 15 to 45 s. All analysed
images are automatically stored and the user can easily revisit them and make manual corrections if necessary. Typically, in the final step of BSE analysis, the user is required to manually add ellipsoidal and irregular bubbles using free-hand drawing tools.

4.4.3. Results reporting

Mean bubble size is the most commonly used parameter to describe a bubble size distribution. This term can be calculated using:

\[ D_{pq} = \left( \frac{\sum_{i=1}^{n} n_i D_i^p}{\sum_{i=1}^{n} n_i D_i^q} \right)^{\frac{1}{p-q}} \]  

(Eq. 4.1)

where \( D_{pq} \) is bubble mean diameter, \( n_i \) is the number of bubbles of diameter \( d_i \), and \( p \) and \( q \) are dimensionless indices defining the moment of the mean. The most commonly used mean diameters include the number or arithmetic mean diameter (\( D_{10} \)), surface mean diameter (\( D_{20} \)), volume mean diameter (\( D_{30} \)), Sauter mean diameter (\( D_{32} \)) and mean mass diameter (\( D_{43} \)).

In order to obtain number frequency and number cumulative BSDs, as well as volume frequency and volume cumulative BSDs, measured bubbles are typically classified into a number of finite size categories. A typical BSD measured in a flotation cell usually contains a large number of small bubbles and several large non-spherical bubbles. For that reason, a dense discretization was used for small bubbles, and size categories were linearly shifted toward larger bubble sizes. To obtain statistically significant BSDs for wide bubble populations, a greater number of images and a large number of bubbles must be analyzed. The range of expected bubble sizes reported in literature varied from 40 µm to 20 mm, with most studies reporting the size range from 40 µm to 2 mm. In most cases, authors working on multiphase flow characterization
claimed to reach statistically significant results by analyzing only from 500 to 1000 bubbles (Junker, 2006). While this might be true for monodispersed, or nearly monodispersed bubble populations, widely dispersed bubble populations require much larger samples. In order to reach statistically significant mean bubble diameter, a population of 1000 bubbles was generally found to be sufficient for $D_{10}$ analysis. On the other hand, a sample of 4000 or more bubbles was found to be necessary for $D_{32}$ and $D_{43}$ analyses. In order to reach this population size, a great number of images had to be analyzed. Generally, the required number of images that had to be analyzed decreased as the mean bubble size decreased. In light of these requirements, bubble images captured with modified MBSA method at locations 1, 3, and 4 were analyzed with both the standard NE and new BSE image analysis techniques. In order to reach a minimum of 4000 bubbles necessary to generate statistically significant results, at least 150 images were processed with the fully automated NE technique and about 15 random images were analyzed with the semi-automated BSE technique.

4.5. RESULTS AND DISCUSSION

4.5.1. Effect of image analysis technique on measured bubble size

Figure 4.6 shows the volume and number frequency distributions obtained with both image analysis techniques for identical image sets. For any experimental condition and any sampling location, the NE technique failed to count bubbles larger than 2 mm, which is well illustrated in the volume frequency distribution plots. The number frequency BSDs revealed insignificantly small numbers of bubbles larger than 2 mm for most tests. On the other hand, large bubbles represent a significant portion of the total gas volume and, therefore, need to be detected accurately.
Figure 4.6 Number and volume frequency bubble size distributions obtained with BubbleSEdit (solid line) and Northern Eclipse (dashed line) image analysis techniques.
Figure 4.7 plots the volumetric fraction of bubbles larger than 1.5 mm for each experiment against the measured Sauter mean bubble diameter. For the same experimental condition, the volumetric fraction of bubbles larger than 1.5 mm is almost two times greater in the case of images analyzed with BSE technique. The data suggest that the NE image analysis technique biases the mean bubble size toward smaller values. To avoid this problem, the BSE software was used for all future image analysis.

![Image of Figure 4.7](image)

**Figure 4.7** Sauter mean bubble diameter versus volume fraction of gas contained in bubbles larger than 1.5 mm using two image analysis technique.

### 4.5.2. Effect of sampling method

Two bubble sampling methods, *ex-situ* and *in-situ*, were used simultaneously to capture bubble images at three different sample locations in the flotation cell. Figure 4.8 shows the local volumetric frequency and cumulative BSDs for both sampling methods
under various operating conditions. In order to produce clear presentation of the results and provide easier comparison between sampling techniques, both BSDs are presented in a logarithmic bubble diameter scale. Also, presenting the data as cumulative distributions helped to smooth out variations found in the frequency distributions. The comparison of BSDs obtained with both methods suggests that the majority of the gas volume is carried by large bubbles, which makes them extremely important for flotation performance evaluations. The presence of relatively large bubbles or bubble slugs significantly increases the Sauter mean diameter, decreases the superficial gas flow rate, and consequently, decreases the bubble surface area flux, which is known to be a factor directly related to flotation kinetics.

Intrinsic differences in the optical characteristics of each sampling system did not greatly influence the size limit of the smallest bubble detected, and this size was nearly identical for both sampling methods. On the other hand, the size of the largest detected bubble differed significantly between the methods. Under the same conditions, the in-situ method managed to capture larger bubbles in the gas-liquid mixture. The data obtained with the ex-situ method indicate that the majority of bubbles detected are within a narrow size range (<2 mm), which are contained in the left peak of the bimodal volumetric BSD curve. It is also interesting to note that all of the cumulative BSDs obtained with the ex-situ method do not have the standard characteristic S-shape. This irregularity was found to become more severe as the sampling tip was moved from the top to the bottom of the cell. This trend may be an indication of the increased likelihood that bubbles interact and coalesce as they travel along the sampling tube, which would be an inherent problem for the ex-situ method at greater depths.
Figure 4.8 Volume frequency and cumulative bubble size distributions obtained with ex-situ (solid line) and in-situ (dashed line) sampling techniques under different operating conditions.
4.5.3. **Effect of sampling location**

Figure 4.9 shows the effect of the sampling location on BSDs for a 5 m/s impeller tip speed and 1.37 cm/s superficial gas flow rate. In this figure, a representative image from the image set obtained with the *in-situ* method was chosen for each of six different sampling locations identified in Figure 4.3.

![Image showing representative bubble images obtained at different cell locations for 5 m/s impeller tip speed and 1.37 cm/s superficial gas flow rate.](image)

**Figure 4.9** Representative bubble images obtained at different cell locations for 5 m/s impeller tip speed and 1.37 cm/s superficial gas flow rate.
Under the same operating condition, bubbles observed at locations 1, 2, 3 and 6 appeared to be larger than bubbles observed at locations 4 and 5, which are positioned bellow the stator ring. Moreover, regardless of the experimental conditions, bubbles or slugs of gas that exceeded 6 mm diameter size were observed at most positions except location 5.

This observation indicates that larger bubbles, with higher rise velocities than local fluid flow velocities, do not follow the main fluid flow pattern. Only a fraction of generated bubbles from the impeller/stator zone appear to follow the main streamlines and return to the high energy dissipation zone multiple times.

This viewpoint is in good agreement with visual observations of flow patterns in gassed stirred vessels that indicate that very large bubbles often pass directly through the stirred tank, concentrating near the center of the vessel, while small bubbles recirculate in the cell (Nienow, 1977).

Occasionally, very large gas slugs are observed at sampling location 6, directly above the impeller/stator gap, indicating the inability of the impeller to fully disperse all introduced air. This data also appears to verify that the local fluid flow direction and velocity, and local bubble velocity, have a direct effect on the local BSD, which is in agreement with an earlier study made by Schäfer (2000).

To quantitatively assess the influence of sampling location on the measured bubble size, Sauter mean diameters and volume fractions of bubbles larger than 1.5 mm are plotted versus sampling locations in Figure 4.10. For a constant impeller tip speed of 5 m/s and various aeration rates, the largest Sauter mean bubble diameters were measured at location 3. In contrast, the smallest Sauter mean diameters and lowest volume fractions
of bubbles larger than 1.5 mm were measured in the impeller discharge jet (location 4), which is in agreement with earlier studies (Alves, 2002; Laakkonen, 2005).

**Figure 4.10** Sauter mean bubble diameter and volume fraction of gas contained in bubbles larger than 1.5 mm at different sampling locations and different superficial gas flow rates.

### 4.5.4. Effect of aeration rate and impeller speed

Figure 4.11 shows the effect of aeration rate on bubble size distributions. As expected, an increase in aeration rate shifted the distribution towards larger bubble sizes. The same trend was observed at all sampling locations. By looking at the volume frequency plot, it can be concluded that a number of large bubbles increased significantly as the aeration rate increased. On the other hand, when looking at the number frequency plot, it seems that a majority of bubbles remained small as the aeration rate increased, with most bubbles contained in a relatively narrow range of bubble diameters.
Figure 4.11 Effect of aeration rate on BSDs at sampling location 4 and at 5 m/s impeller tip speed (in-situ bubble sampling).

Tip speed was found to be another variable that influenced the bubble size distribution throughout the cell. As shown in Figure 4.12, higher impeller speeds shifted the BSD towards smaller bubble diameters. It is extremely important to note here that by using different bubble sampling methods and different image analysis techniques, which in most cases fail to detect large bubbles, the effect of impeller tip speed on bubble size can be easily overlooked.

Figure 4.12 Effect of impeller tip speed on BSDs at sampling location 1 and at 1.37 cm/s superficial gas flow rate (in-situ bubble sampling).
4.6. CONCLUSIONS

An existing ex-situ and a new in-situ bubble sampling methods were used simultaneously to investigate local bubble size distributions in a pilot-scale (0.8 m³) Dorr-Oliver® flotation cell. In addition, two types of image analysis software, the fully automated Northern Eclipse package and semi-automated BubbleSEdit package, were used to analyze captured images and to obtain bubble size distributions for each image set. All experiments were conducted in two phase system and the flotation cell was run as a batch reactor.

The experimental data showed that significant variations in bubble sizes occurred throughout the cell and under different operational conditions. Due to this variability, care must be taken when performing bubble measurements in mechanical flotation cells. Bubbles found below the froth-pulp interface contribute significantly to the processes occurring in the froth zone, while bubbles sampled in the impeller discharge stream contribute to the overall bubble-particle interaction dynamics occurring in the turbulent zone of the cell.

Measured bubble size distributions and Sauter mean bubble diameters revealed significant differences between sampling methods and image analysis techniques. In general, the commonly used ex-situ bubble sampling methods and image analysis techniques fail to detect larger bubbles. This oversight can result in misleading conclusions since larger bubbles carry a significant fraction of the total gas volume, which under some operating conditions can exceed 70% of the total introduced gas volume that are carried by the bubbles larger than 1.5 mm.
Nevertheless, the general simplicity and ease of use makes the *ex-situ* method useful whenever a large number of tests have to be performed in a short timeframe, which is typical for experiments in industrial settings. However, information gained by the ex-situ sampling method give local mean diameters of bubbles entering the froth phase. Therefore, in order to achieve a better insight into the spatial gas distribution profile in the cell a radial screening of bubble sizes has to be performed.

The new *in-situ* sampling method and BubbleSEdit image analysis technique are more demanding, but provide a more accurate estimation of the true bubble size distribution at all locations within a mechanical flotation cell.

### 4.7. ACKNOWLEDGEMENTS

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### 4.8. REFERENCES


CHAPTER 5:
COMPARISON OF GAS DISPERSION MECHANISMS
FOR FORCED-AIR AND SELF-AERATED
MECHANICAL FLOTATION CELLS

Authors: Miskovic, Sanja; Luttrell, Gerald, Ragab, Saad; Elhady Fayed, Hassan
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5.1. ABSTRACT

A comprehensive pilot- and industrial-scale experimental investigation was conducted to examine the gas dispersion performance of two commercial mechanical flotation cells (forced-air Dorr-Oliver® and self-aerated WEMCO®). Measurements were performed over a wide range of operating conditions in both two- and three-phases. Experimental bubble sizing data obtained utilizing a standard ex-situ bubble sampling method revealed differences in bubble size distributions between the two cells. Subsequently, a detailed pilot-scale study utilizing an in-situ bubble sampling method was conducted to compare local bubble size distributions and local mean bubble diameters for the two different machine designs. Differences in the spatial distribution of bubble sizes between forced-air and self-aerated agitating mechanisms were observed. Based on this information, the mechanisms of gas dispersion for each cell type are discussed.
5.2. INTRODUCTION

Flotation is the most widely used separation process in the mineral processing industry today. This process is used for separation of almost all sulfide and many non-sulfide metallic minerals, industrial minerals and energy minerals such as coal and bitumen. There are four main types of flotation cells used in the mineral processing industry: mechanical flotation cells, pneumatic flotation cells, froth separators, and flotation columns. From the beginning, mechanical flotation cells have been the most widely used flotation cells in the mineral industry. Mechanical flotation cells consist of a tank, typically cylindrical shaped, fitted with an impeller drive assembly and a stator. The main function of the stator, which is positioned around the impeller, is to transform tangential flow of the pulp in the cell in the radial direction. The impeller, on the other hand, provides the energy necessary for successful flotation operation and is therefore considered to be the heart of the flotation cell. Bubbles are generated and dispersed by forced introduction of the air through a deeply submerged rotating impeller in forced-air mechanical cells and by self-aeration of shallow rotating impellers in self-aerated mechanical cells.

The main goal in the process of designing of any flotation cell is to maximize gas-liquid interface in the pulp (the particle-water-air mixture) and hence to increase the probability of collisions between air bubbles and hydrophobic particles. Furthermore, every flotation cell should be designed to achieve the following performance functions:

- to generate adequate turbulent conditions for successful bubble-particle attachment in the contact zone,
• to allow sufficient solids suspension,
• to perform efficient gas dispersion,
• to allow unhindered particle transfer from the pulp to the froth zone,
• to allow proper froth removal, and
• to provide optimal retention time for all three phases (gas, liquid, and solid) necessary for achieving maximal material recovery. (Lelinski, 2005).

The total energy introduced by mechanical agitation is, therefore, spent to perform three main functions: (i) particle suspension and transport, (ii) gas break-up into bubbles and dispersion throughout the flotation cell and (iii) generation of micro-turbulences necessary to facilitate bubble-particle collisions (Figure 5.1).

![Figure 5.1 Distribution of the total power consumed during the flotation process.](image)

A fraction of the energy supplied to the system is transported throughout the cell by means of kinetic energy contained in the generated turbulent eddies of different scales.
Most of this energy is dissipated through micro-interactions of all three phases, which include:

- bubble-particle collisions, attachment, and aggregation (Luttrell and Yoon, 1992a),
- liquid-particle viscous friction and lubrication (Eskin et al., 2005), and
- particle-particle partially inelastic collisions.

In general, bubble size (or bubble size distribution) in the flotation system is determined by the following three hydrodynamic processes:

- bubble formation in gas generator,
- bubble breakup, and
- bubble coalescence.

The latter two processes are directly governed by the local turbulence. To develop a better understanding of the role of gas bubbles in the flotation system, knowledge of the above-mentioned phenomena is required.

During the process of bubble formation, a fraction of the total energy supplied to the system is directly transformed into the free surface energy of newly created bubbles. A majority of bubbles created in flotation systems are generated in the zone between the impeller and stator blades, which is maximum energy dissipation zone of the cell. In this zone, air cavities that are initially formed at the low pressure region of the impeller blades are detached from the impellers blades edges and are carried into the high energy dissipation zone where the bubble breakup occurs. Bubble breakup is caused by the
dynamic pressure and shear stresses on the bubble surface induced by shear flow and turbulence (Hinze, 1955).

Bubbles created in the high energy dissipation zone are carried along by the radial flow coming from the impeller and are scattered throughout the pulp contained in the body of the flotation cell (van't Riet and Smith, 1973). Thereafter, bubbles climb through the vessel due to their buoyancy but are also randomly pushed around by the existing turbulence in the cell.

In the flotation cell, turbulence is the primary mechanism responsible for breakup of the bubbles that are initially created in the high energy dissipation zone (Kolmogorov, 1949). Only eddies that are of the similar length scale as the bubble size can break the bubbles. On the other hand, the large eddies can only transport the bubbles, while very small ones do not affect the bubbles (Olmos et al., 2001). In order to determine the actual size limit of eddies responsible for bubble breakage, Prince and Blanch (Prince and Blanch, 1990) conducted experiments to show that only eddies bigger than $0.2d$ are able to break bubbles of diameter $d$, while eddies bigger than $d$ can only move them. Therefore, based on Wu et al. (Wu et al., 1998), bubble breakup rate depends on the frequency of collisions between bubbles and eddies of a similar size. Moreover, bubbles will break up into smaller bubbles only when the maximum hydrodynamic forces in the liquid (that tend to break up the bubbles) are larger than the surface tension force (that tends to stabilize the bubbles) (Angeli and Hewitt, 2000; Hinze, 1955; Kerdouss et al., 2006). This force balance is typically quantified by the liquid Weber number:
\[ We = \frac{\rho u_t^2 L}{\gamma} \]  
(Eq. 5.1)

where \( \rho \) is the density of the continuous phase, \( L \) is the characteristic length, \( \gamma \) is the interfacial tension, and \( u_t \) is the velocity of eddies in the inertial sub-range of turbulent eddy spectrum, which can be calculated as (Kerdouss et al., 2006):

\[ u_t = 1.4(\varepsilon d)^{1/3} \]  
(Eq. 5.2)

where \( \varepsilon \) is the local turbulent dissipation energy and \( d \) is the bubble size.

Bubbles are not stable and break up when the Weber number is larger than a critical value for a certain system (Walter and Blanch, 1986). Critical Weber numbers given in literature generally lie in the range from 1 to 5, although no general agreement about a constant critical Weber number could be found. The critical Weber number typically used for modeling turbulent multiphase flows is in the range from 1.2 to 1.5 (Kerdouss et al., 2006; Lane et al., 2002; Lane et al., 2004; Rigby et al., 1997). Taking all this into account, Wu et al. (Wu et al., 1998) has developed an expression for bubble break-up rate as follows:

\[ N_{br} = C_{br} n \frac{(\varepsilon d)^{1/3}}{d} \left(1 - \frac{We_{\text{crit}}}{We}\right)^{1/2} e^{-\frac{We_{\text{crit}}}{We}}, \quad We > We_{\text{crit}} \]  
(Eq. 5.3)

where \( C_{br} \) is the dimensionless coefficient.

The process of bubble formation described above defines the initial bubble size in the system, which is then altered due to the coalescence. Coalescence occurs when two or more bubbles collide and remain in contact for sufficient amount of time to allow the
liquid in the thin film between them to drain and eventually, when the thin film reaches a critical thickness, to rupture (Kerdouss et al., 2006). Therefore, the rate of bubble coalescence is related to the frequency of collisions between bubbles that are moving in the turbulent flow field (Prince and Blanch, 1990) with the velocity equal to the turbulent fluctuating velocity of eddies of equivalent sizes (Lane et al., 2002). The general equation for the coalescence rate derived by Wu (Wu et al., 1998), can be expressed as follows:

\[ N_{co} = C_{co} \eta_{co} q^2 u_t n^2 \cdot \frac{1}{\left(1 - \alpha_2^{1/3}\right)} \]  

(Eq. 5.4)

where \( C_{co} \) is the dimensionless coefficient, \( \eta_{co} \) is the coalescence efficiency, \( \alpha_2 \) is the gas volume fraction and \( n \) is the bubble number density given by:

\[ n = \frac{\alpha_2}{\left(\frac{\pi}{6}\right) d^3} \]  

(Eq. 5.5)

By inclusion of the gas volume fraction in the final collision rate equation, the phenomenon of reducing of the mean free path between bubbles with increasing of the gas volume fraction could be accounted for. The coalescence efficiency for the stirred tank, as defined by Lane et al. (Lane et al., 2002), can be expressed as:

\[ \eta_{co} = e^{-\sqrt{\frac{We}{\pi}} e^{-K_{co} d^3}} \]  

(Eq. 5.6)

where \( K_{co} \) is the fitting constant. The first part of the expression, proposed by Chester (Chesters, 1991), quantifies the ratio of the liquid film drainage time and the bubble interaction time. The second part of the expression accounts for the minimum amount of energy required to cause bubble deformation and liquid drainage since, during any bubble
collision, the bubble approach velocity has to be sufficient to overcome the pressure rise due to the liquid being forced out between the bubbles (Lane et al., 2002). Therefore, coalescence is the phenomenon that significantly reduces the dispersion efficiency of the flotation system (Cho and Laskowski, 2002). Coalescence strongly depends on chemical parameters of the system, such as the liquid surface tension, volumetric fractions of the dispersed phase, and characteristics of the turbulent flow field (Chesters, 1991).

After they have been generated, bubbles in the flotation system are moved into different zones of the cell depending on the balance between the bubble buoyancy force and surrounding drag forces. Eventually, all bubbles leave the pulp by forcing themselves out through the pulp-froth interface and then, by complex processes that occur in the froth phase, a majority of the gas finally leaves the cell directly through the froth surface or is, in less extent, carried from the system by the recovered froth.

Therefore, local bubble size and bubble size distribution in the flotation cell strongly depends on various operational, technical and chemical factors whose effects on bubble size should be taken into account while designing or modeling a flotation process. Some of the factors that have the utmost effect on bubble size in the flotation systems are:

- the total gas intake,
- the total supplied energy,
- the physical and chemical properties of the liquid and solid phases,
- the impeller/stator design,
- the impeller relative location to the bottom of the cell, and
- the size and geometry of the cell.
Therefore, during the flotation process, the gas-liquid hydrodynamics in the system strongly depend on the type of bubble generator used, so the choice of a proper impeller/stator assembly to satisfy the necessity of optimal gas dispersion is the key for the success and economy of the process. However, the mechanism of bubble generation in the flotation systems has not yet been explored in depth and differences between the bubble generation mechanisms of different flotation cells have not yet been addressed in the literature.

This work presents results obtained through an initial industrial-scale experimental investigation on gas dispersion performance of two commercial FLSmidth mechanical flotation cells: forced-aerated Dorr-Oliver® and self-aerated WEMCO®. The preliminary results revealed significant differences in bubble size distributions between the two cells. To better understand this phenomenon, a comprehensive pilot-scale study under non-coalescing conditions was conducted. Non-coalescing conditions were chosen to allow separate investigation of bubble break-up mechanisms for each cell type by eliminating coalescence effects on the final bubble size. Furthermore, the region under consideration during this investigation was the turbulent zone surrounding the impeller/stator assembly, where the bubble break-up process occurs and determines the bubble mean size (Laakkonen et al., 2005a; Parthasarathy and Ahmed, 1994; Parthasarathy, 1994; Parthasarathy, 1991).

This study aims to determine the effect of different generator designs on the bubble generation process and on the gas dispersion pattern within the flotation cell. Further, knowledge of gas dispersion mechanisms provides valuable information on the
bubble size and bubble interfacial area in the flotation system, which are key parameters governing the flotation process.

5.3. EXPERIMENTAL

5.3.1. Industrial-scale flotation cells

Two industrial-scale mechanical flotation cells, the 300m$^3$ WEMCO® SuperCell and 330m$^3$ Dorr-Oliver® SuperCell, were tested in order to investigate their gas dispersion characteristics during a plant survey of a commercial copper concentrator. In order to collect local bubble samples from the flotation cells a visual/photographic technique was used. The photographic method was chosen since it is able to detect the broadest bubble size distribution in comparison with other methods used for bubble sizing.

For the purpose of this investigation, bubble sizing experiments were performed on both cells using the McGill ex-situ bubble sampling method (Gomez and Finch, 2002; Gomez and Finch, 2007). By using this method, bubbles are observed, as they rise from the sampling tube and enter the viewing chamber, through an inclined glass window of the viewing chamber illuminated from the back and placed outside of the flotation tank (Figure 5.2). The suction point of the bubble sizing apparatus was positioned 130 cm below the launder lip and at the mid-radial distance from the froth crowder to the launder. The sampling locations for each cell tested are shown in Figure 5.3.
Figure 5.2 Ex-situ bubble sampling apparatus in operation.

Figure 5.3 Sampling locations for 330 m$^3$ Dorr-Oliver and 300 m$^3$ WEMCO mechanical flotation cells (after FLSmidth).
Due to the presence of weak hydrodynamic forces in the pulp surrounding the bubble sampling tube and relatively long length of the sampling tube, the sampling apparatus had to be anchored at four points to allow unimpeded operation. Bubble images were collected for several different operating conditions. Operating conditions were changed by varying the impeller agitation rate (from 5 to 9 m/s impeller tip speed) and aeration rate (from 0.7 to 1.88 cm/s global superficial gas velocity) for the Dorr-Oliver cell and by changing the aeration rate (from 0.8 to 1.5 cm/s superficial gas velocity) and the froth depth (from 18 to 68 cm) for the WEMCO cell. Bubbles were recorded at the end of each test period, which was at least fifteen minutes from the moment operating conditions were changed, to allow the system to reach a steady-state. Approximately 400 bubble images were collected for each operating condition and then subsequently analyzed with BubbleSEdit image analysis software (Zabulis et al., 2007). BubbleSEdit utilizes a cross-correlation technique, which ensures detection of all bubbles in an image and, in that way, allows their inclusion into the final analysis of bubble size distributions.

5.3.2. Pilot-scale flotation cells

In order to extensively investigate a bubble generation process within the flotation cells, a series of bubble size measurements were performed on two 0.8 m³ pilot-scale, Dorr-Oliver and WEMCO, flotation cells. The cells were operated as batch reactors, with water and air only (a two-phase system). The frother selected for this investigation was methyl isobutyl carbinol (MIBC), which was added in concentrations over its critical coalescence concentration. This created a non-coalescing environment in the cell allowing investigation of gas dispersion characteristics of each cell type pertinent to bubble breakage mechanisms.
Although an ex-situ bubble sampling method represents a well-designed solution for the bubble sampling in the upper quiescent zone of industrial flotation cells, it cannot be used successfully for bubble sampling from the lower regions of the cell. Bubbles present in the turbulent zone generally follow major fluid streamlines and therefore fail to enter the sampling tube. Since bubble size distribution varies within a flotation cell and is a direct function of the relative location from the impeller, a new bubble sampling method, capable of collecting representative samples throughout all regions of the flotation cell, was developed and used in this pilot-scale study.

A new *in-situ* bubble sizing apparatus was designed to perform accurate measurements of bubble sizes through different regions of the flotation cell and to minimize potential flow interference within the cell. The system can be submerged into any location of the flotation cell, where it directly captures images of bubbles from the bubble stream. Moreover, it is capable of measuring bubble sizes in the range of 50 μm to 10 mm as defined by the resolution and magnification of the optical system employed, which was GP-21400 GEViCAM GigE high-speed CCD camera with 1392×1040 pixels resolution and a compact red LED backlight. Both elements were mounted inside of watertight, pressurized enclosure (Figure 5.4).

Gap width between the light source and the camera was set to be 10 mm. This gap width was found to produce images with the least number of bubble overlaps, which created better conditions for subsequent image analysis. Image analysis was performed utilizing BubbleSEdit software. Results of automatic image analysis were manually corrected for each image analyzed, which ensured that more than 90% of all bubbles in an image were included in the analysis.
Bubble images are captured for a series of operating conditions by changing impeller rotation speed from 4 to 7 m/s impeller tip speed for both cells and by changing the aeration rate, from 0.8 to 1.9 cm/s superficial gas velocity, for the Dorr-Oliver cell. Before each test the WEMCO cell was completely filled with water and then put into operation. In this way, for each agitation speed tested, the cell was allowed to displace all excess water before it reached the steady-state operation. This allowed us to test the cell while operating with its natural water level for each monitored agitation speed.
Since the bubble size distribution is affected not only by operating conditions but is also a function of the relative location in the cell, a bubble sample was taken from the four selected locations in the cell, representing the major flow patterns in the turbulent zone of the cell. The locations chosen for each cell type, with the expected main fluid streamlines, are presented in Figure 5.5.

**Figure 5.5** Locations used for bubble sampling in Dorr-Oliver (left) and WEMCO (right) pilot-scale flotation cells. Also given are the main liquid phase (full arrows) and gas phase (dotted arrows) pathways in the cell.

As can be seen from the figure, bubble measurements were conducted in a single vertical plane, 15 cm away from the gas disperser, for the WEMCO cell. On the other
hand, the Dorr-Oliver cell was screened at three different locations in a vertical plane, 15 cm from the stator ring, and from one other location positioned 3 cm above the impeller stator gap. In both cases, the location 2 was selected to allow bubble sampling from the discharge stream coming from the impeller. Also, positions 1 and 3 represent, respectively, fluid streams sampled 8 cm above and 8 cm below sampling point 2.

5.4. RESULTS

Figure 5.6 shows bubble size frequency (bars graph) and bubble size cumulative (dashed line) distributions, and fractions of the total gas contained in each size class (area graph) for the 330 m³ Dorr-Oliver flotation cell at different operating conditions. The illustration also shows corresponding Sauter mean and number mean diameters, as well as a fraction of the total gas volume contained in bubbles larger than 1.5 mm. The top row in the illustration shows the effect of increasing aeration rate on the bubble size distribution, under the same agitation rate. It can be clearly seen that an increase in aeration rate shifts the bubble size distribution toward larger bubble sizes, which is also reflected through an increase in mean bubble diameters from 1.5 to 1.97 mm for the Sauter mean diameter, and from 0.92 to 1.18 mm for the number mean diameter.

The bottom row in the illustration shows the effect of an increase in agitation rate on bubble size distribution in the cell, under the same aeration rate. In this case, measured bubble sizes and corresponding distributions do not follow expected trends. The bubble population shifts toward larger sizes when the impeller speed increases from 5.44 to 7.2 m/s and then they move back toward smaller sizes when the impeller speed changes from 7.2 to 8.95 m/s. The reason for this irregularity may lie in the method of bubble sampling.
Throughout this investigation, bubble sampling was performed in the quiescent zone of the cell, at half-radial distance from the launder lip to the froth crowder.

\[
\begin{align*}
TS &= 7.2 \text{ m/s} \\
J_g &= 0.75 \text{ cm/s} \\
TS &= 7.2 \text{ m/s} \\
J_g &= 1.23 \text{ cm/s} \\
TS &= 7.2 \text{ m/s} \\
J_g &= 1.88 \text{ cm/s}
\end{align*}
\]

**Figure 5.6** Bubble size frequency and bubble size cumulative distributions with the corresponding Sauter and number mean diameters for the Dorr-Oliver SuperCell under different operating conditions. Also given are the fractions of the total gas volume contained in a certain size class. Top row represents the effect of the aeration rate increase; bottom row represent the effect of the agitation rate increase.

Under the assumption that the gas dispersion limit for this cell was exceeded at impeller tip speed of 5.44 m/s and aeration rate of 1.43 cm/s, the cell was operated under so-called
‘boiling conditions’. This assumption leads to a conclusion that a fraction of the introduced gas was not carried by the discharge stream coming from the impeller, but was by-passing the system through the central zone of the cell. In this case, escaping gas could not be detected at the chosen sampling location, which could have led to the bias in final results. Therefore, using a single location for bubble screening in large industrial systems might significantly affect overall detection efficiency, which leads to undercoverage error.

Figure 5.7 shows bubble size results obtained for the 300 m³ WEMCO cell. For the same impeller tip speed, operating conditions were changed by altering the froth depth, which resulted in change of the aeration rate. Generally, an increase in froth depth increases the aeration rate. The illustration shows the increase in bubble size with the increase in aeration rate (froth depth). This increase became more significant as the froth depth was increased from 43 to 68 cm, which corresponds to 1.18 and 1.42 cm/s superficial gas velocity, respectively. At the deepest froth depth, the fraction of the total gas contained in bubbles larger than 1.5 mm reached 78%, and the number mean bubble diameter became 1.29 mm. This indicates that lowering the pulp level below a certain (natural) level for the self-aerated cells results in a decrease of the impeller pumping capacity, which proportionally decreases the impeller gas dispersing capabilities.
Figure 5.7 Bubble size frequency and bubble size cumulative distributions, fractions of the total gas volume contained in a certain size class, and corresponding Sauter and number mean diameters for the WEMCO SuperCell for different operating conditions.

It is important to note here that the direct comparison of these two flotation cells is probably not completely justified. One reason for this is their significantly different design and principles of operation. Moreover, and more importantly, they were tested on two different days, which might have caused variations of some operational and chemical factors that are strongly affecting bubble size, due to expected variations during the plant operation. Nevertheless, bubble populations sampled from each cell, under similar operating conditions, were still compared.

Operating conditions selected for this comparison were 7.2 m/s impeller tip speed and 1.23 cm/s superficial gas velocity for the Dorr-Oliver cell and 7 m/s impeller tip speed and 1.18 cm/s superficial gas velocity for the WEMCO cell. Under these conditions, the WEMCO cell generated smaller bubbles, which is well reflected through both number mean (0.89 mm - WEMCO vs. 0.97 mm - Dorr-Oliver) and Sauter mean bubble diameter (1.39 mm - WEMCO vs. 1.69 mm - Dorr-Oliver). In addition, a much narrower bubble size frequency distribution was noted for the WEMCO cell.
In order to explore this phenomenon in more depth and to allow fair comparison of these two flotation cells, a detailed pilot-scale study was initiated under controlled experimental conditions. Figure 5.8 shows representative bubble images for both pilot-scale flotation cells taken at four selected sampling locations. Both cells were operated at 7 m/s impeller tip speed, while the superficial gas velocity was fixed at 1.44 cm/s for the Dorr-Oliver cell and was measured to be 1.41 cm/s for the WEMCO cell.

It can be readily noticed that, under analogous operating conditions, the WEMCO cell tends to generate a larger number of smaller bubbles that can be detected at all sampling locations. It can also be noted from the presented images that bubble populations sampled below the discharge flow coming from the impeller contain an insignificant number of big bubbles in comparison with locations 1 and 2. This indicates that the process of bubble segregation, based on bubble size, occurs in the discharge stream. Figure 5.8 also shows the assumed gas phase pathways within the cell.

Figure 5.9 and Figure 5.10 show bubble size distribution results obtained for the 0.8 m³ Dorr-Oliver flotation cell, at four different operating conditions.
Figure 5.8 Representative bubble images obtained for each sampling location for both flotation cells when the impeller tip speed was 7 m/s and aeration rates were 1.44 cm/s for the Dorr-Oliver cell and 1.41 cm/s for the WEMCO cell.
Figure 5.9 Bubble size frequency and cumulative distributions, fractions of the total gas volume contained in a certain size class, and Sauter and number mean diameters for the pilot-scale Dorr-Oliver cell at different locations and for low aeration rate.
Figure 5.10 Bubble size frequency and cumulative distributions, fractions of the total gas volume contained in a certain size class, and Sauter and number mean diameters for the pilot-scale Dorr-Oliver cell at different locations and for high aeration rate.
Bubble diameters measured at different locations in the cell when it was operated under low aeration rate (0.86 cm/s superficial gas velocity) and medium (5 m/s) and high (7 m/s) agitation rates are presented in Figure 5.9. As can be seen from the figure, the increase in agitation rate, under a low aeration rate, does not affect the characteristics of the majority of bubbles present in the bubble population, which is well reflected through the relatively constant number mean diameter (goes from 0.45 to 0.49 mm) obtained for both conditions at locations 1, 2 and 3. On the other hand, Sauter mean bubble diameter decreases from approximately 1.1 to 0.7 mm, indicating that the number of big bubbles in the population drops significantly with the increase in agitation rate. This trend can be clearly observed by looking at the area graph in the figure, indicating that the fraction of total gas contained in large bubbles drops significantly when the agitation rate increases. Results obtained for location 4 show an insignificant number of big bubbles carried into the zone above the impeller, indicating that the machine dispersed gas effectively for both operating conditions.

The effect of agitation rate increase on bubble size distribution under high aeration rate (1.88 cm/s) is presented in Figure 5.10. In this case, a large fraction of gas was detected at location 4 when the cell was operated under a 5 m/s impeller tip speed, suggesting that the gas dispersion limit was exceeded at that operating condition and that the cell was running under boiling conditions. On the other hand, a significantly large fraction of gas was also noted at location 4 when the impeller tip speed was increased to 7 m/s. Even though this might not be a direct indication of boiling conditions in the cell, it suggests that, under high aerating conditions in the forced aerated cell, the total volume of gas introduced into the system cannot be completely dispersed into small bubbles in
the generator zone, which could potentially result in the escape of larger bubbles through the impeller/stator gap. In addition, comparison of the number and Sauter mean bubble diameters leads to the conclusion that, again, the majority of bubbles remain in the same size scale (around 0.7 mm) but, on the other hand, the number of big bubbles in the population decreases significantly when the impeller tip speed increases from 5 to 7 m/s. This is also well reflected through the area graph presented in the Figure 5.10. Finally, when comparing bubble sizes obtained at the same agitation rate and at two different aeration rates, a significant increase of both number mean and Sauter mean diameters can be noticed. This clearly indicates the strong impact of aeration rate on bubble generation process, and therefore on bubble size, in forced-aerated flotation cells.

Figures 5.11 and 5.12 show bubble size distributions obtained during pilot-scale, two-phase, gas dispersion testing of the WEMCO flotation cell. The operating conditions were altered by increasing of the impeller tip speed, which resulted in the change in aeration rate, which was accordingly measured. As impeller speed increased from 4 to 7 m/s, the aeration rate gradually increased from 0.5 to 1.4 cm/s. From the figures, the measured number mean bubble diameter remains practically the same (0.45±0.04 mm) under all operating conditions and at all sampling locations. This suggests that the basic mechanism of bubble creation remains the same, since the majority of bubbles in the population have the same characteristics for all tested conditions, which can easily be noticed from the number frequency distribution graphs.
Figure 5.11 Bubble size frequency and cumulative distributions, fractions of the total gas volume contained in a certain size class, and Sauter and number mean diameters for the pilot-scale WEMCO cell at different locations and for 4 and 5 m/s impeller tip speeds.
Figure 5.12 Bubble size frequency and cumulative distributions, fractions of the total gas volume contained in a certain size class, and Sauter and number mean diameters for the pilot-scale WEMCO cell at different locations and for 6 and 7 m/s impeller tip speeds.
On the other hand, the Sauter mean bubble diameter and area graphs presented show significant increase of a number of big bubbles in the cell with the increase in agitation rate. At the location 2, the Sauter mean diameter increased from 1.1 mm at lowest agitation rate to 1.6 mm at the highest aeration rate. Over the whole range of agitation rates tested, bubble populations sampled at the location 1 contained significantly larger number of big bubbles, which is seen in the strong peak over the large size fractions in the area graphs.

For the range of operating conditions tested, a self-aerated flotation cell generates a greater number of small bubbles (bubbles smaller than 0.5 mm) than a forced-aerated cell. This can be clearly seen from the number frequency and cumulative frequency distribution graphs (Figures 5.9 to 5.12) for both tested cells. This indicates that there are some fundamental differences in the way bubbles are generated for these two flotation cells.

5.5. DISCUSSION

The comparison between the gas dispersion results obtained for two flotation cells reveals a relatively wide variation of bubble size distributions measured at different locations in the turbulent zone of each cell. The data show that the characteristics of a bubble population strongly depend on aeration rate for the forced–aerated cell and on agitation rate for the self-aerated cell. Figures 5.13 and 5.14 provide a simplified graphic representation of the gas phase pathways for each cell type under three extreme operating conditions. The thickness of the arrows in the drawings is roughly proportional to the fraction of the total gas volume that is transported in that direction.
Figure 5.13 Gas phase pathways within the forced-aerated, Dorr-Oliver flotation cell for three different aeration rates (low, medium, and high) and constant, high agitation rate. The thickness of the arrow is proportional to the fraction of the gas being transported in that direction.

Figure 5.14 Gas phase pathways within the self-aerated, WEMCO flotation cell for three different aeration rates (low, medium, and high) and constant, high agitation rate. The thickness of the arrow is proportional to the fraction of the gas being transported in that direction.
Figure 5.13 shows gas phase pathways inside of the forced-aerated Dorr-Oliver flotation cell at a constant high agitation rate and at three different aeration rates (low, medium and high). The drawings suggest that the majority of the gas introduced into the cell is carried by the discharge stream dispersed radially from the impeller. From this stream, a fraction of the gas contained in small bubbles has a chance to be carried to the bottom of the tank, where part is re-entrained into the generator zone. On the other hand, based on the findings from this study, a larger fraction of the gas carried by the discharge stream is moved toward the upper zone of the cell and eventually leaves the cell from the top.

Depending on the operating conditions, one fraction of the gas contained in large bubbles that are carried by the discharge stream coming from the impeller does not follow the main fluid stream lines, but leaves the cell directly. This phenomenon is very important for understanding the overall hydrodynamics inside of the flotation cell. Generally, there are two main forces contributing to the hydrodynamics and, consequently, to overall gas distribution pattern in the flotation cell: turbulent dispersion forces, reflected directly through the drag force of the continuous phase (liquid) (Simonnet et al., 2007), and buoyancy of the dispersed phase (gas) (Sokolichin et al., 2004). For each location in the cell, the balance of these two forces defines the bubble path and its terminal velocity. At the same agitation rate, an increase in aeration rate increases both overall gas holdup in the system and volume of the recirculated gas, which dampens the turbulent kinetic energy of the discharge stream. On the other hand, an increase in aeration rate results in the increased production of large bubbles in the generator zone. Both of these effects hinder the discharge stream capacity to transport
large bubbles and increase the probability of the of escape large bubbles from the radial jet to the upper flow field. In this case, the probability of detecting large bubbles in the discharge stream decreases as the bubble size increases. Moreover, the probability of detecting large bubbles above the discharge stream, in this case, will be greater than in the discharge stream.

At first, this mechanistic description might appear contradictory since the discharge stream is the region of high energy dissipation rates and it is expected that the large bubbles are broken-up here before they have a chance to escape, which should result in the small bubbles in both regions. Even though, in general, this might be the case for the majority of big bubbles, and it is well reflected through the similar shape of the size frequency distribution curves (Figures 5.9 to 5.10), some bubbles will still have a chance to escape. It is very important to take large bubbles into consideration since a small number of large bubbles carry a significant fraction of total gas from the system and in that way decrease process efficiency.

These results point to an important fact for modeling of the two phase flows in flotation cells. Care must be taken when modeling break-up in the impeller discharge stream since it is not the only phenomenon that defines final bubble size. Recirculation of the primary bubbles and bubble buoyancy should also be incorporated into the modeling codes for simulation of flotation systems.

As a result of the observed phenomenon represented in Figure 5.14, it can be seen that the overall gas distribution in the cell goes from a uniform to non-uniform pattern as the aeration rate is increased from low to high. This is understandable in a view of the
fact that significantly narrower bubble size distributions, which are shifted toward smaller bubble sizes, are generated when the aeration rate is low, while broad bubble size distributions, shifted toward larger bubble sizes, are generated during operation with the high aeration rates and at constant agitation rate (right columns of Figures 5.9 and 5.10). Generally, small bubbles tend to follow all major fluid streamlines in the turbulent zone of the cell while larger bubbles have more chances to overcome local drag forces due to their higher buoyancy, and are moving directly toward the cell surface.

This non-uniformity can be best observed by looking at the radial distribution of the gas phase right below the pulp-froth interface. Here, uniformly distributed gas entering the froth zone, while running at low aeration rates, becomes more concentrated toward the cell wall as the aeration rate is increased. During operation with high aeration rates, the gas distribution profile takes a saddle shape, with higher gas fractions leaving the cell in the zone closer to the cell wall and around the impeller shaft. After this point, if the aeration rate were increased, the gas dispersion capacity of the impeller would be exceeded, which means the cell would run under boiling conditions, and the gas distribution profile at the top of the pulp would be strongly skewed toward the center of the cell.

Gas distributions for the three selected operating conditions for the self-aerated WEMCO flotation cell are described in the Figure 5.14. The three limit operating conditions reflect low, medium and high agitation rates. As can be observed in the figure, the expected gas distribution pattern, in the turbulent zone of the self-aerated cell resembles the pattern in the turbulent zone of the forced-aerated cell. Due to the shorter radial distance from the disperser hood to the cell wall and shorter vertical distance from
the discharge stream to the cell surface, the gas distribution profile, close to the pulp-froth interface, is somewhat different from that of the forced-aerated cell. For all selected operating conditions, the fraction of the gas reporting to the froth will be slightly higher close to the cell wall and gradually decreases toward the disperser hood. It is also found that, for the self-aerated cell, the gas preferentially concentrates in the upper zone of the cell with practically no bubbles present in the bottom half of the cell.

Schematic representations of bubble generation processes occurring in the high-energy intensive (impeller) zone, for both mechanical flotation cells, are presented in Figures 5.15 and 5.16. In both cases, gas cavities are formed at the low-pressure, trailing edge of the impeller blades, which is the first stage of the process of bubble creation. Thereafter, bubbles are shed from the tail of the cavity by the turbulent eddies. Small bubbles are formed as the high circulating velocity in eddies dissipates through the radial flow of the fluid. Therefore, energy dissipation occurring when turbulent, high-intensity eddies disintegrate is one of the main factors important for the creation of small bubbles (Stephenson et al., 1998). In the Dorr-Oliver cell (Figure 5.15), the gas is introduced directly to the impeller through the six openings at the bottom of the impeller disc. The introduced gas accumulates in the gas cavities where it is radially distributed toward the cell body. Since cavities are, at the same time, receiving and releasing the gas, they are also known as ventilated cavities.
Figure 5.15 Cavity formation, bubble generation mechanism, and main fluid pathways for the Dorr-Oliver impeller. Dashed lines represent different cavity profiles for different aeration conditions. 1 - boiling conditions; 2 – optimal conditions; 3, 4 – medium aeration rates; 5 – high aeration rate.

Figure 5.16 Cavity formation, bubble generation mechanism, and main fluid pathways for the WEMCO impeller. Dashed lines represent different cavity profiles for different horizontal locations of the impeller. 1 – top end of the impeller; 5 – bottom side of the impeller.
The cavity profile, represented through the dashed lines in the figure, strongly depends on the volumetric gas rate, the flow characteristics of the up-coming fluid stream (gray bordered arrows in the figure), and the characteristics of the turbulent eddies created at the impeller blade edge. Five different cavity profiles, as a function of different aeration rates under a constant, high agitation rate, are shown in the figure. Numbers 1 to 5 reflect operating conditions with high to low aeration rates.

Profile 2 represents the operation just below the maximal gas dispersion limit, which is typically well reflected through the minimal value of the gassed power to ungassed power ratio for a certain agitation rate. This operating condition represents the optimal operation point which typically results in the maximal performance. The upward shift of the cavity profile (line 1 in the figure) represents process operation when the gas dispersion limit is exceeded (boiling conditions). In this case, part of the gas introduced into the impeller by-passes the generator zone and escapes directly through the gap between the impeller and stator.

For the WEMCO cell, aeration rate is a function of the total fluid flow pumped through the impeller. The tangential flow that is generated in the disperser region is important for the creation of the forced vortex and surface aeration of bubbles, which result in gas induction (Mundale and Joshi, 1995). The surface aeration of bubbles is the main bubble generation mechanism in the self-aerated cells. During the operation, the liquid level in the generator region and cell is directly affected by the impeller rotational speed. Generally, as impeller speed increases the liquid level decreases. It is followed by an increase in the volume of the introduced gas up to a certain point when it starts decreasing. This reduction in the induced gas flow rate is a result of two phenomena
occurring in the generator cavity as suggested by Patil and Joshi (Patil and Joshi, 1999a; Patil and Joshi, 1999b):

1. impeller increased exposure to the gas phase due to the very low liquid level, which results in a sudden drop of the impeller pumping capacity, and
2. impeller drowning as a result of the flow reversal in the upper part of the generator cavity, which reduces the impeller capability to induce the gas.

Figure 5.16 shows the estimated cavity profile formed at the low pressure side of the WEMCO impeller when operated at optimal agitation rate. Profile lines presented in the figure (left image) depict the gas cavity profile at different horizontal levels of the impeller (right image). As can be seen from the figure, one side of the blade is almost completely covered with the gas phase, while the other side of the blade is covered with the liquid phase. Created long cavities and much longer impeller blade lengths, which are capable of dispersing the gas by breaking the created free surface, support the creation of the large numbers of small bubbles.

5.6. **CONCLUSIONS**

Operating conditions and created flow conditions in a mechanical flotation cell have a considerable impact on the distribution of the gas phase throughout the cell. Typically, large bubbles generated in the high-energy impeller-stator zone can be found in the impeller discharge stream, suggesting that the large scale vortices present in the discharge stream have the capability to capture and transport larger bubbles. On the other hand, with the increase in aeration rate for forced-aerated cells or with an increase in agitation rate of self-aerated cells, a significant number of large bubbles leaving the
generator zone can escape from the discharge stream. This phenomenon is strongly affected by the balance of local drag force coming from the continuous phase and the bubble buoyant force. However, when flotation cells are operated under optimal conditions, the largest fraction of the total gas entering the cell is contained in small bubbles, which are carried by the discharge stream to the tank wall. From there, one fraction is transported to the bottom of the cell and reintroduced into the high-intensity zone.

Findings from this study suggest that the bubble diameter in a flotation system is not determined by a single phenomenon. Typically, bubble break-up due to the high energy dissipation rates is the determining factor, but there are several other mechanisms that have to be taken into account. Bubble buoyancy, recirculation of the primary bubbles and trailing vortices generated behind the large bubbles and bubble swarms are some mechanisms that should also be considered.

In summary, local bubble size distributions for two different types of mechanical flotation cells and for different operating conditions have been reported. These results are now available for the further development and refinement of existing flotation models and for the validation of existing numerical simulations.

5.7. REFERENCES


CHAPTER 6:
SUMMARY AND CONCLUSIONS

A fully-instrumented 0.8 m$^3$ pilot-scale flotation circuit was developed for the purpose of providing performance data that can be more readily utilized for the engineering design, scale-up and optimization of industrial flotation circuits. This highly flexible system enabled measurement, monitoring and control of a number of hydrodynamic and metallurgical parameters over a wide range of operating conditions. Additionally, a new, robust, in-situ bubble sampling apparatus was developed and validated in both two- and three-phase conditions. The new apparatus allowed bubble monitoring at different locations within the pilot-scale mechanical flotation cell, which could not be performed in the past by utilizing some of the commercially available bubble sampling systems. These newly developed state-of-the-art systems allowed for the detailed investigation of flotation performance of different pilot-scale flotation machines over a wide range of operating conditions. The flotation cell was operated as either batch or continuous reactors in both two-phase (liquid/gas) and three-phase (liquid/gas/solid) modes.

The main conclusions obtained from the comprehensive hydrodynamic and metallurgical investigation are listed below:

- The continuous flotation circuit was successfully and easily operated with slurry containing more than 30% w/w of solids and allowed continuous and simultaneous monitoring of multiple flotation parameters over a wide range of operating conditions.
• Two different bubble sampling methods (i.e., a new in-situ technique and a commercially available ex-situ technique) were used simultaneously to investigate local bubble size distributions in a pilot-scale machine. The measured bubble size distributions and Sauter mean bubble diameters revealed significant differences between the two sampling methods. Most importantly, the commonly used ex-situ bubble sampling method failed to detect larger bubbles present in the flotation pulp.

• Two types of image analysis software, the fully automated Northern Eclipse package (edge detection image analysis technique) and semi-automated BubbleSEdit package (cross-correlation image analysis technique), were used to analyze captured images and to obtain bubble size distributions for each image set. The commonly used edge detection image analysis technique generally failed to detect larger bubbles due to their non-spherical shape and greater chance for bubble overlap.

• The miscounting of large bubbles from the image sets can result in misleading conclusions about the gas dispersion properties inside the flotation cell since large bubbles represent a significant fraction of the total gas volume. Experimental data collected in the current study indicated that the fraction of the total introduced gas carried by larger bubbles (>1.5 mm) could exceed 80% of the total gas volume when the cell was operated under high aeration rates and low agitation rates.

• General simplicity and ease of use makes the ex-situ method useful whenever a large number of tests have to be performed in a short timeframe. However, information gained by the ex-situ sampling method gave local mean diameters of bubbles entering the froth phase, which produced misleading results when the bubbles were sampled from only one location.
• Although the new in-situ sampling method and BubbleSEdit image analysis technique was more demanding, this method provided a more realistic estimation of the true bubble size distribution at all locations within a mechanical flotation cell.

• Bubble populations were found to vary significantly at different vertical and radial distances from the impeller/stator assembly, and the degree of the variation strongly depended on the operating condition. Due to this variability, care must be taken when performing bubble measurements in mechanical flotation cells. Bubbles found below the froth-pulp interface contribute significantly to the processes occurring in the froth zone, while bubbles sampled in the impeller discharge stream contribute to the overall bubble-particle interaction dynamics occurring in the turbulent zone of the cell. Therefore, in order to achieve better insight into the spatial gas distribution profile in mechanical flotation cells, radial screening of bubble sizes needs to be performed.

• The Sauter mean bubble diameters obtained in this study ranged from 0.5 to 3 mm, which is wider than bubble size distributions previously reported in the literature. The larger mean bubble diameters were obtained due to increased precision achieved with the in-situ bubble sizing method by which up to 98% of all recorded bubbles in an image were detected and included in the analysis.

• The measurement of power consumption provided considerable insight into the gas dispersion capabilities of the rotor/stator mechanism. Specifically, the ratio of aerated-to-unaerated power plotted as a function of the aeration number provided important information about the minimum agitation rate necessary for complete dispersion of the gas introduced into a flotation machine. The first sign of the
transition of the overall flow pattern from “loaded” to “flooded” conditions can be easily observed through power input monitoring.

- The power input had a positive effect on the number of bubbles created in the cell, which increased the probability of bubble-particle attachment, increased overall carrying capacity, and therefore increases the flotation rate constant.
- A decrease in mean bubble diameter was achieved by decreasing superficial gas velocity and increasing specific power input.
- Flattening of the $D_{32}-P^*$ trend was observed for all aeration rates tested, which suggests that there is a minimum energy input needed to achieve an optimal bubble size distribution in the cell for a given constant aeration rate.
- For all particle sizes, high aeration and agitation rates resulted in higher material recoveries. For fine and intermediate particle sizes, recovery increased as residence time increased over all operating conditions.
- For coarser particles, an increase in agitation rate increased off-the-bottom suspension and increased particle concentration suspended in the pulp. This condition ultimately created more favorable conditions for bubble-particle encounter in the high-turbulent zone of the cell.
- A correlation between the flotation rate constant and bubble surface area flux was observed for all glass particles tested, while the nature of this correlation strongly depended on the size of the particles.
- As a result of the heterogeneous gas distribution within the cell, the bubble surface area flux values estimated from global superficial gas velocities overestimated the bubble surface area flux.
• A significant number of large bubbles leaving the generator zone were found to escape from the discharge stream with an increase in the aeration rate for forced-aerated cells or with an increase in agitation rate for self-aerated cells. This phenomenon is believed to be strongly affected by the balance of local drag force coming from the continuous phase and bubble buoyant force.

• When flotation cells are operated under optimal conditions, the largest fraction of the total gas entering the cell was found to be contained in the small bubbles. These smaller bubbles were carried by the rotor discharge stream to the tank wall. From there, a fraction of these small bubbles was transported to the bottom of the cell and reintroduced into the high-intensity zone created by the rotor.

• The bubble diameter in a flotation system was not determined by a single phenomenon. Typically, bubble break-up due to the high energy dissipation rates was the determining factor, but there were several other mechanisms that must be taken into account. Bubble buoyancy, recirculation of the primary bubbles, and trailing vortices generated behind the large bubbles and bubble swarms are mechanisms that should also be considered.

• Local bubble size distributions were carefully measured for two different types of mechanical flotation cells over a range of different operating conditions. These experimental results are now available for the further development and refinement of existing flotation models and for the validation of existing numerical simulations.

In summary, data obtained using the pilot-scale system can be used as a baseline for advanced modeling, control and optimization of flotation processes. With its functional versatility, the system can be easily adapted to almost any process condition and, in that
way, provide valuable process-related knowledge necessary for the development of successful scale-up strategies and more efficient flotation cells. However, care must be taken during the flotation scale-up process since not all parameters relevant to flotation scale in proportion. Different sizes of flotation cells, used through the scale-up procedure, have different ranges of optimal operating conditions and their own characteristic limitations.