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MIXING CHARACTERISTICS OF DRAFT TUBE AIRLIFT BIOREACTOR USING THE ELECTRICAL RESISTANCE TOMOGRAPHY

by

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B.Sc., Kwame Nkrumah University of Science and Technology, Kumasi, Ghana, 2005

A Thesis

Presented to Ryerson University

in Partial Fulfillment of the Requirements for the Degree of

Master of Applied Science

in the Program of Chemical Engineering

Toronto, Ontario, Canada, 2010

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ABSTRACT

Farouzatu Yakubu-Gumery, Mixing characteristics of draft tube airlift bioreactor using the electrical resistance tomography, MASc Chemical Engineering, Ryerson University Toronto, 2010.

In this work, mixing characteristics in terms of mixing time, hydrodynamics (liquid circulation velocity and gas hold up) and shear rate were performed in the downcomer of a draft tube airlift bioreactor with different geometries (i.e., A_d/A_r between 0.38 - 2.31 and bottom clearances between 0.003-0.009 m). Newtonian (water and 34.5% coalescing sugar solution) and non Newtonian (0.2% and 0.5% xanthan gum solutions) with different viscosities were used as the liquid phase. Compressed air was used as the gas phase which was introduced through cross and circular shaped sparger configurations at superficial velocities $U_{gr} = 0.00165$ -0.00807 m/s. The combined effects of geometric parameters (A_d/A_r , bottom clearances), sparger configuration, and liquid viscosity on mixing characteristics have been presented.

Results showed that the increase in superficial gas velocity (U_{gr}) corresponds to an increase in energy generated, and thus decreases in mixing time. However, the increase in U_{gr} corresponds to the increase in liquid circulation velocity, gas holdup and shear rate values. Moreover, bottom clearances and draft tube diameters show effects on flow resistance and frictional losses which affect results of mixing parameters investigated. The influence of sparger configurations on mixing time and liquid circulation velocity is significant due to their effect on gas distribution. Mixing time decreased to about 40% in air-water media using the cross shaped sparger. Results obtained with cross shaped sparger showed even and uniform distribution of gas, which provided better mixing as compared to the circular shaped sparger configuration. However, the sparger configuration effect on shear rate is not as significant (about 20% reduction in shear rate values using the cross shaped sparger). The effect of fluid viscosity had a significant influence on both mixing times and circulation velocity, especially in the coalescing media of sugar and xanthan gum solutions.

Results from this work will help to develop a clear pattern for operation and mixing that can help improving several industrial processes, especially the ones related to emerging fields of technology such as the biotechnology industry.

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CHAPTER 1

INTRODUCTION

Mixing is an important unit operation in biotechnology, petroleum refinery, wastewater purification, readymade food production and other allied chemical industries. With requirements that vary from fluidization, stripping of gas to leaching, mixing enhances homogeneity in order to achieve a desired product quality. This process can be accomplished in stirred tanks, bubble columns or airlift bioreactors for multiphase mixing in industry. Mixing begins from complex interactions between molecules and thus affects chemistry of products formed. It is therefore, necessary to understand the mixing parameters such as the hydrodynamics (gas holdup and liquid circulation velocity), transport properties (heat and mass transfer coefficients) and mixing time for a successful design and scale-up. In scaling-up the mixing process to industrial level, one should be able to predict the performance comparable to that of laboratory scale. When scale-up fails, it implies inefficient mixing, which compromises product quality. The cost of correcting this anomaly is high as it was estimated that the US chemical industry alone spent upto \$10 billion extra due to poor mixing (Smith, 1990). Based on this figure, the economic potential for efficient mixing is highly marked in addition to improvement in product.

We are in the biotechnology era, with mixing in bioreactors being the core of any bioprocess. The commercial manufacture of pharmaceutical (penicillin), chemicals (citric acid), foods (cheese, beer, vinegar) are largely dependent on biotransformation of organisms in a bioreactor (Chisiti, 1989). Mixing of fermentation broth in most bioreactors is very important because it affects productivity (Guillard and Tragardh, 1999). Considerable studies to evaluate the mixing performance of a bioreactor have been documented and summarized in review papers by Chisti

and Moo-Young (1988), Chisti (1998), and Petersen and Margaritis (2001). Conventional measuring techniques to assess mixing include the magnetic flow follower (Merchuk and Stein, 1981; Weiland and Onken, 1981), the use of conductivity probes (Weiland and Onken, 1981) and hot film anemometer (Young *et al.*, 1991). These techniques are intrusive therefore disrupting the process flow and are not suitable for opaque reactors largely in use for industrial work.

With the development of imaging techniques for diagnostics purposes in the medical field using the computerized axial tomography (CAT), scientists and engineers have adapted and improved the tomography system to visualize the internals of process vessels. For industrial purposes, the tomography measurements are based on electromagnetic (X-ray, γ -ray etc.), electrical (ERT, ECT, and EIT) and acoustic (ultrasonic) with most popular one being the ERT (Williams and Beck, 1995). Quite recently, the ERT system has been used for evaluating the performance of different types of vessels (Fransolet *et al.*, 2001, Haibo *et al.*, 2006, Razzak *et al.*, 2007, Pakzad *et al.*, 2008). The ERT technique is cheaper as compared to the ones used for medical diagnostics. It is also non intrusive to process flow and also very suitable for opaque reactors.

The objectives of this research therefore would be to exploit the use of ERT to evaluate the mixing performance in a draft tube airlift bioreactor.

CHAPTER 2

LITERATURE REVIEW

2.1 Types of Bioreactors

Bioreactors used for mixing are stirred tanks, bubble columns, airlift and fluidized beds (Table 2.1, Fig.2.1). Agitation in a stirrer tank reactor is accomplished by mechanical stirring (using impellers). They provide excellent mixing and good values of mass and heat transfer coefficients. Drawbacks of the stirred vessels are mechanical issues related to the moving parts (e.g. shafts and impeller) which can introduce some level of contamination while operating cost is also higher due to high energy consumption and cost of maintenance (Joshi *et al.*, 1990).

The airlift reactor was first patented for use as a bioreactor by Lefrancois in 1955 (Gavrilescu and Roman, 2004). Due to the fluid dynamic characteristics, it continues to gain popularity for its success in the production of microorganisms (Siegel *et al.*, 1988; Kim *et al.*, 1997). Airlift reactors are also used in petroleum (Mehrnia *et al.*, 2004; Shariati *et al.*, 2007), and wastewater treatments (Jin *et al.*, 2002), where satisfaction of high oxygen levels for higher mass transfer rates is necessary for gas-liquid contact.

The bubble column and airlift bioreactors are pneumatically agitated and often employed in bioprocesses where gas-liquid contact is important. The role of the gas is to provide contact with the liquid for mass transfer processes such as absorption or desorption and to provide energy through gas expansion or bubble buoyancy for liquid mixing. In these two pneumatically agitated reactors, gas is sparged usually through the bottom and the buoyancy of the ascending gas bubbles causes mixing. The main difference between these two pneumatically agitated reactors is in their fluid flow characteristics. The flow in the airlift bioreactor is ordered and in a cyclic pattern like in a loop beginning from top through to bottom.

The airlift bioreactor differs from the bubble column by the introduction of inner draft tubes which improves circulation, whereas the bubble column is a simple tower. In the airlift, liquid recirculation occurs due to the four distinct sections: the riser, downcomer, gas separator and bottom or base. The bubble column is a simple vessel without any sectioning making the flow rather a complex one.

Types of Bioreactors	Advantages	Disadvantages
Airlift Reactors	 Similar to bubble column. Draft tube to improve liquid circulation. Better mixing with high mass transfer properties. Provides low shearing due to the absence of mechanical mixers. Low cost due to simple construction of vessel with lower power requirement. 	 Bubble coalescence can sometimes reduce mass transfer properties. Not suitable for very viscous solutions.
Bubble Column Reactors	 Very simple. Uses sparged air for mixing (minimal energy consumption). The absence of moving parts avoids mechanical breakages and contamination. 	• Inhomogeneous shear might be due to that air is sparged at a focal point.
Fluidized Bed Reactors	 Fluid is used to fluidize catalysts. Enhanced heat and mass transfer properties Insoluble and high viscosity solutions can be used. 	• High energy input is required to fluidize the solid particles.
Stirred Tank Bioreactor	• Uses mechanical agitation (impellers) for efficient distribution of heat and mass transfer properties.	 Possibility of contamination and high shear environment. High energy input is needed to power the impellers.

Table 2.1 Bioreactors used for mixing.



Fig.2.1. Types of Bioreactors for Mixing

Some attractive features of the airlift bioreactor are the low power consumption, simplicity in construction with no moving parts, high mass and heat transfer rates and uniform distribution of shear (Chisti and Moo Young, 1994; Merchuk, 1990). The advantage of its low power consumption is of particular importance in effluent (e.g. wastewater) treatment where the product value is comparatively low. Therefore, operational cost (efficient use of energy) is greatly considered since this work is usually on a large scale. Homogenous shear is particularly important for biological processes that are shear sensitive. In the conventional stirred tank, shear is greatest at the stirrer and decreases away from it to the walls of the vessel. This creates a gradient of shearing which can have adverse effect on the morphology or sometimes can damage cells (e.g. animal and plant cells). The simple construction of the airlift without shafts makes it not only aesthetically pleasing to look at but also eliminates contamination associated with the conventional stirred tank which is a major drawback in the production of microorganism. A

sterile environment is crucial for growing organisms especially in the bioprocesses since contamination reduces product quality, generates wastes, also more time and money are spent to restore the whole process.

2.2 Airlift Bioreactors and Their Configurations

There are two main types of airlift bioreactors; (1) the external loop vessel where the liquid circulation in the riser and downcomer takes place in two separate compartments. The external loop usually reaches nearly total gas disengagement at the top section giving rise to a higher difference in density or hydrostatic pressure. This results in lower gas recirculation (thus lower mass transfer) with higher liquid circulation velocity to enhance fluid mixing; (2) the structure of the internal loop vessel has various modifications with regards to the placement of the draft tubes (Fig.2.2), examples of which are the split-cylinder, draft-tube, concentric tube etc. It is the difference in configuration that allows variation of operation of the reactors in terms of liquid circulation and gas disengagement.

Irrespective of the configurations of the airlift reactors, it has four distinct zones or sections with variable local hydrodynamics (gas holdup, liquid velocity and mass transfer rates) thus, accurate description and prediction of its performance are challenging. Fig.2.3 illustrates the various sections in an airlift bioreactor using the internal loop as an example.



Fig.2.2. Types of airlift bioreactors: (a) split-cylinder-internal loop; (b) draft- tube internal loop and (c) external loop.

The four sections are; (a) the riser, this is the section where gas is normally sparged and the flow is upwards. This section usually has a higher gas holdup and it is here that most of the gas-liquid mass transfer takes place; (b) the downcomer is parallel to the riser and combines the top and bottom parts. The flow in here is downwards and the liquid recirculates as a result of pressure or density difference due to partial or total disengagement of gas at the top of the draft tube; (c) gas separator, which is at the top, connecting the riser and downcomer and allows gas disengagement and liquid recirculation. This section affect gas holdup and liquid velocity depending on its geometry and (d) bottom section connects the riser and downcomer and the geometry of this section has an impact on gas holdup, liquid velocity and solid flow (for three phase flow).



Fig.2.3. Individual sections of a draft tube airlift (internal loop vessel).

Although various flow regimes in gas-liquid flow have been reported and studied (Heijnen *et al.*, 1997; Vial *et al.*, 2001; Gavrilescu and Tudose, 1998), the homogenous bubble and heterogeneous turbulent regimes are of great interest. The homogenous regime occurs at low gas velocities where the flow is characterized by a uniform distribution of relatively small bubbles. This offers a larger surface area for mass transfer. In this case, the velocity of the gas phase is equal to that of liquid phase, resulting in less or no turbulence. The heterogeneous turbulent regime consisting bubbles of varying sizes can be observed with an increase in the gas flow rate or when the diameter of the reactor increases (Merchuk, 1990). In this case, the gas holdup is low, since, too many large bubbles tend to occupy the entire volume of the reactor and would encourage bubble coalescence subsequently limiting mass transfer rates. Within these two flow regimes, three circulation pattern regimes also exist (Fig.2.4).



Fig.2.4. Circulation regime in an internal loop airlift bioreactor.

In the first regime, the gas input is the lowest. Therefore the liquid velocity is insufficient to entrain gases into the downcomer. In the second regime (normally the transition regime) there is stratification of bubbles. As gas flow rate increases, induced liquid circulation velocity is adequate to entrain bubbles to the downcomer. The entrainment of bubbles into the upper part of the downcomer is visible. There is complete gas (bubble) recirculation in the third regime due to sufficient energy from liquid velocity in the downcomer. This is attributed to an increase in gas flow which promote the gradual descent and eventually the gas recirculation in the whole vessel.

2.3 Hydrodynamic Characteristic of Airlift Bioreactors

The fluid hydrodynamics is simply its characteristics when in motion. Fluid mixing is influenced by the mixing time and gas holdup which defines the fluid circulation and mass transfer properties. The fluid recirculation causes the difference in hydrostatic pressure and density due to partial or total gas disengagement at the top clearance (TC). Studies have been documented during the last two decades with various correlations applicable for hydrodynamic parameters (Chisti, 1998; Joshi *et al.*, 1990; Petersen and Margaritis, 2001). This implies that for a successful design, fundamental understanding of mixing parameters is important for industrial scale-up.

It is difficult to generalize the performance of the bioreactor according to the process for which the airlift will be employed. For example, in aerobic fermentation, oxygen is important for mass transfer and therefore, it is imperative to consider a design where there will be less disengagement of gas resulting in higher gas holdup for a higher mass transfer rate. In this case, the liquid circulation velocity is low because less gas is disengaged at the top resulting in a lower differential pressure and density. Furthermore, other processes require good mixing other than a high mass transfer rate. However, provision can be made by increasing the gas disengagement at the top to improve the liquid recirculation as in the case for anaerobic fermentation. Therefore, it is safe to conclude as have been confirmed (Merchuk *et al.*, 1994; Gavirilescu and Tudose, 1998; Chisti, 1998) that, the geometry parameters such as the top clearance (TC), ratio of cross sectional area of the downcomer to the riser (A_d/A_r), bottom clearance (BC), the cross sectional areas of riser (A_r) and that of the downcomer (A_d), draft tube internal diameter (D_d), and height of the column (H) and superficial gas velocity in the riser (U_{gr}) have an influence on fluid hydrodynamics. Table 2.2 lists some of the operation parameters investigated during mixing in an airlift bioreactor.

There is extensive information on the measurement of fluid hydrodynamics published with a handful of equations. However, most of them cannot be correlated due to the different medium (Newtonian versus non-Newtonian) used and various assumptions made. The different measuring techniques often used are not suitable to opaque reactors whiles others disturb process flow.

2.3.1 Gas Holdup

Gas holdup is defined as the gas fraction within the total bioreactor volume, equation 2.1;

$$\varepsilon_g = \frac{V_G}{V_G + V_L + V_S} \tag{2.1}$$

Where V_G , V_L and V_S are volumes of gas, liquid and solid phases, respectively.

The gas holdup ε_{g} , offers two main advantages (Merchuk and Gluz, 1999), gas in the liquid volume determines the residence time of the gas and liquid in combination with the bubble size; thereby affecting the gas-liquid interfacial area available for mass transfer. As gas holdup increases, the greater the area for mass transfer rate. This however, depends on the amount of oxygen present as continuous recirculation of gas depletes the bubbles of O₂. The difference in gas holdup creates the driving force for liquid circulation in the riser and downcomer regions.

Reference	$D_{c}(\mathbf{m})$	D_d (m)	A_d/A_r	Gas sparger	U_{gr} (m/s)
Bendjaballah et al.,	0.10	0.057	0.33	62 holes, d _o =1mm	0.01-0.17
1999					
Blazej et al., 2004	0.108-	0.070-0.200	1.23-1.01	Multiple orifice sparger, 25-90 holes,	0.01-0.03
	0.294			d _o = 0.5-1mm	
Mehrina et al.,	0.14	0.11-0.09	0.707-1.306	Perforated ladder, 30 holes, d _o =1mm	0.01-0.08
2004					
Miron et al., 2004	0.193	0.144	0.8	Perforated pipe sparger, 13-17 holes,	0.001-0.05
				d _o =1 mm	
Vial et al., 2002	0.06	0.1	0.36	Multiple orifice 62 orifices d _o =1 mm	0.01-0.24

Table 2.2 Operating parameters investigated in an airlift bioreactor.

2.3.1.1 Techniques to Evaluate Gas Holdup

Gas holdup is usually measured with the aid of manometers in different parts of the bioreactor "locally" or as an "overall" in which ε_g is calculated from the difference between volumes of aerated and unaerated liquid phase as in equation 2.2 (Merchuk *et al.*, 1994);

$$\varepsilon_g = \frac{h_D - h_L}{h_D} \tag{2.2}$$

where h_L is gas free liquid height and h_D is dispersion height after the introduction of air. The use of manometers to measure the pressure difference (Merchuk and Stein, 1981; Al Marsy, 2001) may cause contaminations. Under sterile fermentation conditions special sensors such as pressure transducers (Luo *et al.*, 1997) may be used. Their mode of operation is such that there is minimal interaction with the slurry under observation.

Hot film anemometer probe (Karamanev *et al.*, 1996) was also used to measure the difference of heat conductivity between liquid and gas phase. A downside is that this technique cannot be used in fermentation media because it is an invasive technique and does interfere with process flow.

In non-invasive ultrasonic technique (with introduction of acoustic wave), liquids have a better absorption than in gases. This technique uses the variation in acoustic velocity of sound wave travelling through a medium. It is therefore, suitable for opaque systems and has a faster response (Chang *et al.*, 1984; Stolojanu and Prakash, 1997). Although this technique can be used in fermentation media, Soong *et al.* (1995) described the weakness of this technique due to low variation of ultrasonic velocity in fermentation media.

In the recent past, the electrical resistance tomography (ERT) has been employed for gas holdup measurement in bubble column (Haibo *et al.*, 2006). This technique measures the conductivity of the media and the gas holdup is calculated by applying the Maxwell equation. Gas holdup measurement with the ERT was in agreement with measurement taken with the pressure transmitter. The most important feature of this technique is that it is non invasive and non intrusive to process flow and can be used for opaque medium as well.

2.3.1.2 Correlations and Effect of Geometry on Gas Holdup

Generally, gas holdup behavior shows a dependence on gas flow rate, geometric parameter of the vessel (i.e., A_d/A_r BC and TC; see Table 2.3), type of sparger and the physical properties of slurry (with the introduction of solids). As gas flow rate increases, ε_g will also increase. Research (Gavrilescu and Tudose, 1996 and Bello *et al.*, 1989) have also shown that generally a decrease in the ratio of A_d/A_r increases ε_g as this dictates the liquid and gas residence time in the vessel.

Models presented in Table 2.3, employed the pressure measuring technique using manometers or piezometers. The model proposed by Hwang and Cheng (1997), overestimates the gas holdup values and this they attributed to bubble coalescence due to high viscosity and gas flow rate. However, correlations of Kemblowski *et al.* (1993) does provide better insight into the hydrodynamics process taking place in the reactor as it takes into account, the friction and pressure losses, and density of the liquid including the geometric parameter of the reactor for all the other models (Kemblowski *et al.*, 1993).

Reference	Media	Conditions	Design	Equation
Hwang and Cheng, 1997	Air Water CMC	$A_r/A_d = 0.69-3.22$ H = 2.5 m $D_d = 0.9-0.14 \text{m}$	Internal Loop	$\varepsilon_{gd} = 0.001 U_g^{0.733} \left(\frac{A_d}{A_r}\right)^{-0.378} \cup_h H_d^{0.125} (1+\varepsilon_s)^{-2.060}$
Kemblowski et al., 1993	Air Water Glycol Syrup CMC	$A_d/A_r = 1-1.33$ $U_{gr} = 0.001-$ 0.15m/s.	External Loop	$\varepsilon_{gr} = 0.203 \frac{F_r^{0.31}}{M_o^{0.012}} \left(\frac{U_{gr}A_r}{U_{lr}A_d}\right)^{0.74}$
Popovic et al., 2004	Air CMC	$A_d/A_r = 0.11-0.44$ H = 1.88m $D_d = 0.15$ m	Internal Loop	$\varepsilon_{gr} = 0.93 U_{gr}^{0.65} (1 - \varepsilon_s)^{0.74} \left(1 + \frac{A_d}{A_r}\right)^{-1.05}$
Renzo, 2005	Air Water	N/A	External Loop	$\varepsilon_g = \frac{U_g/A_r}{0.25 + 1.1(U_g/A_r + U_l/A_r)}$

Table2.3. Models for gas holdup in airlift bioreactors

Gavrilescu and Tudose (1996) evaluated the gas holdup and liquid circulation velocity using alcohol, NaCl and glucose in two external–loop airlift reactors (ELAR). On the laboratory scale, they varied A_{d}/A_{r} (0.111-1.0) and superficial gas velocity (0.016-0.178 m/s), while for the pilot scale, they varied the same as 0.040-0.1225 and 0.010- 0.120 m/s respectively. Later, they reported that generally, the value of ε_{g} in both reactors increased with increasing U_{gr} , however, greatest in reactor with the lowest A_{d}/A_{r} ratio, with the laboratory ELAR recording the highest ε_{g} . This finding was in agreement with Bello *et al.* (1985). The pilot ELAR recorded lower ε_{g} because it had a higher A_{d}/A_{r} which influenced liquid velocity as the U_{g} increased, thereby reducing the residence time of the gas holdup in the riser. The liquid properties (surface tension and viscosity) had little influence on ε_{g} although water had higher gas holdup rates as compared to the other solutions since, water is a non coalescing fluid and the gas holdup effect in the laboratory ELAR was more pronounced.

The importance of geometry on hydrodynamics cannot be over emphasized. Gavrilescu and Tudose (1998) reported the effect of geometry in three concentric tube airlift (0.07, 2.5 and 5.2 m³), using air-water medium with A_d/A_r between 0.1-0.9 and U_{gr} of 0.005-0.1 m/s. Their extensive study explained how these clearances affect the gas holdup, liquid circulation velocity and thus the pressure drop. In general, ε_{gr} and ε_{gd} decreased when TC increased, resulting in an increase in the driving force and hence increase in liquid velocity. The total gas holdup becomes smaller, for an increasing ratio of A_d/A_r , the flow resistance at downcomer entrance is reduced. For an equal TC and BC of 0.07m, ε_{gr} at a given U_{gr} of 0.07 m/s was found to be 65% higher as compared to bottom clearance BC of 0.25 m. The effect of BC was more pronounced as compared to the TC because of the resistance to flow in the downcomer. When the total gas

holdup diminishes as the A_d/A_r increases, the flow resistance at downcomer entrance is also reduced. They proposed correlations for gas holdup in the riser ε_{gr} (equation 2.3) and gas holdup in downcomer ε_{gd} which were validated with previous data (Merchuk *et al.*, 1994). In the correlations, bottom clearance (BC) was found to have an effect on gas holdup in all regions whereas, the U_{gr} effect was more pronounced for ε_{gr} and at the separator:

$$\varepsilon_{gr} = 0.0008 \times G_a^{0.21} \times F_r^{0.82} \times B^{-0.19} \times Y^{-0.43} \times T^{-0.10} \times R^{-0.17}$$
(2.3)

where, $B = \frac{BC}{D_d}$ is the bottom spatial ratio, $F_r = \frac{U_{gr}}{\sqrt{gD_d}}$ is the Froude number, $G_a = \frac{g\rho_{lD_d^3}}{\mu_l}$ is galilei number, *T* is top spatial ratio, $Y = \frac{TC+D_d}{D_s}$ is the gas separator ratio and *R* is the ratio of cross sectional area of downcomer to riser.

Performance of rectangular airlift bioreactor measuring the effect of geometry with gas sparged in the annulus was studied (Klinozo *et al.*, 2007). Experiments were performed in air-water, NaCl and Carboxyl Methyl Cellulose (CMC) with viscosity of 0.02-0.5 Pa.s and surface tension of 0.065-0.085 N/m. It was shown that the bottom clearance (BC) had an effect even in the viscous media. In both systems a decrease in BC resulted in an increase in ε_{gr} . That is, as BC is decreased, velocity of the liquid phase was affected by pressure drop resulting in lower velocity of both bubbles and liquid. A large amount of bubbles were entrained with a large residence time. A lower top clearance (TC) resulted in shorter residence time in the gas-liquid separator.

In three phase systems using calcium alginate as the solids in water, aqueous salt and glycerol solutions in an internal loop bioreactor, the gas holdup decreased with increasing solid concentration (Koide *et al.*, 1992). As the solid loadings increased, viscosity of the system

increased with a decrease in liquid velocity. Furthermore, bubbles coalesced and reduced the gas holdup, although gas holdup values were similar for water and salt solutions. In a similar three phase system using CMC, Wen *et al.* (2005) confirmed that when U_g increased, ε_g in both regions also increased as suggested previously (Hwang *et al.*, 1997). They further analyzed the effect of solid loadings on the hydrodynamics of a three phase system and found that generally, the liquid velocities in the riser and downcomer decreased with an increase in the solid loadings. This could be a result of frequent bubble coalescence which concurs with the idea of Wei *et al.* (2000). Also, the ε_g declined with an increase in solid loadings attributing to a decrease in the flow area. This phenomenon further decreased the mass transfer rate (K_La) since the bubbles coalescence provides less room for interfacial flow area (Wen *et al.*, 2005).

In an air-water-silica sands system using annulus airlift bioreactor, a model was developed to predict liquid recirculation and gas holdup (Sun *et al.*, 2005). They showed that liquid velocity and gas holdup in the riser increased with increasing superficial gas velocity with a lower solid concentration. This observation was similar to that of internal loop reactors (Lu *et al.*, 1994). At a higher concentration of solids, viscosity of the medium increased, while decreasing the gas holdup. Their data on the liquid recirculation velocity and the average gas holdup in the riser varied within \pm 10% and \pm 20% respectively. According to experimental data from water-in-kerosene microemulsion medium using draft tube airlift bioreactor with a diameter of 0.14 m and draft tube heights of 1.1, 1.48 and 1.897 m (Mehrnia *et al.*, 2004) the TC, had very little effect on overall ε_g and K_La which was closely agreed with Koide *et al.* (1985). However these results were not in agreement with Merchuk *et al.* (1994) who reported otherwise which could be

attributed to the medium they used. They observed an increase in ε_g and $K_L a$ with increasing H/D with an increase in liquid velocity.

2.3.1.3 Effect of Liquid Property

Properties of liquids such as the viscosity, surface tension and density also play a crucial role in the hydrodynamic parameters of an airlift reactor. Viscosity, causing resistance to flow can greatly affect the gas holdup hence can also affect the K_La . As viscosity increases, the resistance to flow increases and coalesce more bubbles. As the bubbles coalesce it lowers the interfacial surface area resulting in a decrease in K_La . Erikson and Deshpande (1981) evaluated the effect of viscosity on gas holdup in a split cylinder airlift reactor using CMC. They also reported that the gas holdup gradually decreased with increasing viscosity, since an increase in viscosity shortened the residence time of gas bubbles. This observation concurs with results obtained using CMC for 3 phase flow in an internal loop with draft tube (Wen *et al.*, 2005).

Surfactants are substances that form a monolayer at interfaces and have the ability to inhibit coalescence. They reduce the liquid surface tension hindering bubble coalescence with increases in interfacial area and mass transfer rate. This seemed to have been in the case of Erikson and Despande (1981) when they used sodium lauryl sulphate as a surfactant. By investigating mixing using Kenon 10 and Kenon 4 as surfactants in a draft tube internal loop employing water in diesel microemulsion, it was found that an increase in the surfactant resulted in a decline in the values of \mathcal{E}_g and $K_L a$, which was attributed to the increase in the viscosity of the microemulsions (Shariati *et al.*, 2007). In this case, the viscosity of the fluid clearly dominated the influence of surface tension. The observed trend was clearly in disagreement with Koide *et al.* (1984b) and

Koide (1985) who reported otherwise, as the addition of surfactant increased ε_g and $K_L a$, which also contributed to a reduction of surface tension of the slurry.

Snape *et al.* (1992) compared the gas holdup in NaCl, KCl, Na₂SO₄ and CaCl₂ solutions with 0.5-8% (w/v) sucrose and water and found that, salts (electrolytes) increased the gas holdup. That was due to reduction in surface tension allowing smaller bubbles to form and preventing coalescence. However, in the sugar solution the gas holdup was lowest at a higher sugar concentration. This effect could be explained by the opposing effects of viscosity, density and surface tension. With an increase in sucrose, the density and viscosity have increased while surface tension decreased. The gas holdup was expected would be higher and this showed that the effect of surface tension alone is not sufficient to characterize the liquid property.

The effect of methanol, propanol and butanol at 0.01-0.1% (v/v) on the gas holdup, liquid velocity and mass transfer co-efficient in a split rectangular airlift bioreactor had been investigated (El Azhera *et al.*, 2005). The gas holdup increased with the increasing alcohol level due to the inhibition of bubble coalescence. They also reported a decrease in K_La which contradicted the theory as it was assumed that the larger the interfacial area the larger the transfer rates. They further explained that this effect could have been due to the extended carbon chain, thus the reduction of surface tension was probably minimal as a result of oxygen depletion (as the bubbles recirculated over and over in the bioreactor).

2.3.1.4 Effect of Sparger and Bubble Distribution

Several gas distributors are used for flow distribution in an airlift reactor. Basic designs are either tube-like or circular with several orifices (holes) of different diameters; annulus sparger, draft tube sparger, multiple and single orifice (Fig.2.5). Contreras et al. (1999) found that in a bubbly flow regime, smaller bubbles produced higher interfacial area and hence, ε_g and $K_L a$ were increased. However, in the transition and heterogeneous regimes, the pore size had little effect. During the transition to turbulent churn flow, various factors affected the size of the bubbles by changing the degree of coalescence (Maruyama et al., 1981; Lockett et al., 1975). Furthermore, they reported that the heterogeneous regime had a greater influence on hydrodynamics. Although, the bubble size will be determined by the porosity of the sparger, the effect of the sparger pores on the degree of coalesces cannot be completely ruled out as previously reported (Snape et al., 1992). A higher ε_g and $K_L a$ were observed with increasing number (but smaller diameter) of holes, although at a higher liquid height and in the heterogonous flow, the bubble size effect was determined by coalescing effect (Zhao *et al.*, 1994). The position of the sparger is also important in gas distribution and bubble coalescence. Gas spargers or distributors placed just inside the riser, enhanced gas distribution since the downcomer flow joins riser under the sparger as opposed to spargers placed at the entrance of the riser. In this scenario, the downcomer stream maldistributes the gas bubbles to the walls of the vessel encouraging coalescence (Chisti, 1989). A second sparger (at the top of the downcomer) has been proposed in aerobic fermentation to replenish oxygen (Siegel *et al.*, 1986.)



Fig.2.5. Sparger Configuration; (a) Perforated pipe ladder type gas sparger, (b) multiple orifice sparger, (c) single orifice, (d) annulus ring sparger.

2.3.2 Liquid Circulation and Mixing Time

Liquid circulation velocity U_l is an important parameter in airlift bioreactors which affects the residence time of gas, mass transfer $K_L a$ and mixing time t_m . Studies showed that liquid circulation velocity was affected by the gas flow rate and geometric parameters of the vessel (Gouveia *et al.*, 2003; Luo 2008; Mehrnia *et. al.*, 2004). Liquid circulation occurs due to the difference in hydrostatic pressure or density between the riser and downcomer. When gas flow rate increases, the liquid velocity also increases, thereby entraining most of the bubbles from the riser in to the downcomer. This will reduce the difference in hydrostatic pressure (compromising the liquid velocity). In general, a higher liquid velocity reduces the residence time of the bubbles in the riser and downcomer, as it encourages the recirculation of gas through the downcomer and back to the riser.
2.3.2.1 Techniques for Measuring Liquid Circulation and Mixing Time

Liquid velocity has been measured using the tracer response techniques which measure the conductivity and pH over time. In these methods, the probes that are connected to sensors and microprocessor are inserted into the riser and downcomer of the bioreactor to measure the conductivity or pH upon addition of a tracer. To calculate the velocity, the time taken for the tracer to travel from one probe to the other is measured from recorded peaks or the time taken to record consecutive peaks (Onken and Weiland, 1980). In measuring the conductivity, a conductivity probe is inserted in the bioreactor which measures the response of an electrolyte trace, whereas in measuring the pH an acid or base (eg. NaOH) acts as the tracer. This is a very popular technique due to its relative simplicity. They however disrupt process flow since they come into contact with the slurry under investigation.

The flow follower is another technique which employs a neutrally buoyant solid particle (flow follower) to measure liquid circulation and mixing time. The importance of this technique is that the choices of material selected for the follower must be detected inside a bioreactor and possess fluid like properties of the medium under investigation. Bryant (1976) housed a radio transmitter (radio pill) in a solid particle and thereafter, Bonakdarpour *et al.* (1994) developed a flow follower using polystyrene beads covered with aluminum foil coated with silicone rubber to prevent the absorption of water. Results of this follower were compared with that of the chemical reaction method Bonakdarpour *et al.* (1994) in which iodine-sodium thiosulphate was used and were under 15% deviation. The iodine sodium thiosulphate method recorded the process of decolorization as a measurement of mixing time as used by Carreau *et al.* (1976).

With the thermo anemometry, a thin wire film probe made of platinum is used with heat generated by electricity. The principle of this is that as the liquid flows, the heat is transferred to the liquid and the film is cooled. This technique is ideal for fluctuation velocities where the time interval is short. Results are usually excellent if the probe is frequently calibrated and the sensor is positioned properly to the maximum flow direction during measurements (Brunn, 1996).

The dye or coloring method (Serrano and Galindo, 1997) requires the addition of iodine or methylene blue where the dispersion of the dye is followed visually or by spectrophotometer and therefore, usually reserved for the detection of stagnant regions. This technique however is not applicable to fermentation broths as dyes may be absorbed and used up in fermentation process. Apart from that, most industrial reactors as well as the fermentation media are turbid making quantification quiet tedious.

Ultrasonic doppler velocimetry technique requires the addition of tracer particles to the medium to produce echo for the estimation of doppler frequency. As tracer particles are loaded and start to move around, they transmit and scatter ultrasound beams emitted by the probe generating backscattering at all angles. The frequency generated is used to estimate the doppler frequency that can be used in the calculation of the particle velocity. A disadvantage of this technique is that care must be taken to determine the accurate doppler angle. Wang *et al.* (2003) proposed a correlation to determine the doppler angle which extended this technique to multiphase systems.

2.3.2.2 Correlations and Effect of Geometry on Liquid Circulation and Mixing Time

Although various correlations exist for prediction of liquid circulation velocity, the most popular one was proposed by Chisti and Moo Young (1988) based on an energy balance. The correlation (equation 2.4) was developed for low viscous media and has been widely used both in internal and external loop reactors and adequately validated (Wachi *et al.*, 1991; Cai and Nieuwstad, 1999; Fraser and Hill, 1993; Kemblowski *et al.*, 1993)

$$U_{lr} = \left[\frac{2gh_D(\varepsilon_{gr} - \varepsilon_{gd})}{\frac{K_T}{\left(1 - \varepsilon_{gr}\right)^2} + K_B \left(\frac{A_r}{A_d}\right)^2 \frac{1}{\left(1 - \varepsilon_{gr}\right)^2}}\right]^{0.5}$$
(2.4)

where; U_{hr} is the superficial liquid velocity, Ar is the riser cross-sectional area, A_d is the downcomer cross-sectional area, h_D is the liquid dispersion height, ε_{gr} is the riser gas holdup, and ε_{gd} is the downcomer gas holdup. K_B and K_T are the hydraulic pressure loss coefficients in the bottom and at the top. For an internal loop reactor, K_T is negligible compared to K_B due to open channel in the top section. In the external loop reactors the authors assume an equal value for both K_B and K_T . The frictional co-efficient of K_B can be calculated from the following empirical correlation (equation 2.5) Chisti *et al.* (1988) for internal loop.

$$K_B = 11.40 \left(\frac{A_d}{A_b}\right)^{0.79}$$
(2.5)

where, A_d is the downcomer cross-sectional area and A_b is the cross section at the bottom of the airlift reactor. A weakness; however here is that the local gas holdup in the riser and downcomer

are employed in the equation instead of principal gas flow rate which determines the gas holdups.

Another model predicts liquid circulation velocity and gas holdup in both two phase and three phase flow (Heijnen et al., 1997). For a turbulent regime they assumed a constant ratio of gas holdup between the riser and downcomer with respect to superficial gas velocity (characteristic typical to a homogenous bubble) to solve the momentum balance. Their model consists basically of the gas holdup and liquid velocity which is a representation of the driving force for liquid circulation in the airlift reactor. In the derivation of this model the difference in gas holdup (equation 2.6) was estimated as an overall function of superficial gas velocity for the following expression for estimating the circulating liquid velocity (equation 2.7);

$$\varepsilon_{gr} - \varepsilon_{gd} = 2 \frac{U_g - \varepsilon_g U_{sb}}{U_l} \tag{2.6}$$

where, $\varepsilon_{gr} - \varepsilon_{gd}$ is the difference in gas holdup between riser and downcomer; U_{sb} is bubble swarm velocity, U_l is circulation velocity; U_g is gas superficial velocity.

$$U_{l} = \left(\frac{4gh_{t}}{\kappa_{f}}\right)^{1/3} \left[\left[U_{g} - \varepsilon_{g} U s_{b} - \varepsilon_{s} U_{sp} \left(\frac{\rho_{s}}{\rho_{L}} - 1\right) \right] \right]^{1/3}$$
(2.7)

where, K_f is friction coefficient and h_t is the draft tube height.

This expression does not account for an interaction between gas and solids and also the friction coefficient (K_f) for the top section is not explicit. Although the model is simplified, it has successfully predicated a pilot reactor of 400 L and a 2840 hecto-L fully scale bioreactor for wastewater treatment. Table 2.4 lists a few correlations for liquid circulation and superficial liquid velocity. It can be seen that liquid circulation velocity is greatly influenced not only by the

superficial gas velocity but also by the geometry parameter of the vessel as well as the liquid property.

The predictions of these correlations involve the application of theoretical equations combined with mass and momentum balance over a circulation loop and various empirical correlations for gas holdup and pressure drop. The predictions were based on steady state balance between hydrostatic pressure differences, gas holdup differences between riser and downcomer. Also the pressure drop resulting from resisting forces in circulation paths due to introduction of internals (draft tubes), and flow rates (Merchuk and Stein, 1981).

Mixing is crucial to industrial scale from the laboratory analysis because it is compromised at the industrial level. Since the environmental conditions in the bioreactor fluctuates, mixing time in the bioprocess is important to predict the optimum operating conditions such as pH, temperature, and substrate concentration required for maximum productivity. The estimation of mixing time provides an indication of a time interval during mixing. In such processes knowing the mixing time hinders the product formation out of cells, which reduces product formed and sometimes cell damage.

Mixing time is usually defined as the time required for reaching a certain level of homogeneity (I) (usually 95%) after the introduction of a tracer. The level of I can be expressed as;

$$I = \frac{C - C_{\infty}}{C_{\infty}} \tag{2.8}$$

or

$$t_m = \frac{c - c_\infty}{c_\infty} \tag{2.9}$$

Reference	Media	Conditions	Design	Equation	
Bello <i>et</i> <i>al.</i> , 1984	Air Water NaCl.	U_{gr} =0.0137- 0.086m/s A_d/A_r =0.11-0.69 H= 1.8m D_d = 0.152m	Internal and External Loop	$U_{lr} = 1.55 \left(\frac{A_d}{A_r}\right)^{0.74} U_g^{0.33}$ $U_{lr} = 0.66 = \left(\frac{A_d}{A_r}\right)^{0.78} U_g^{0.33}$	
Chisti et al., 1988	Air, Water NaCl	$U_{gr}=0.01-$ 2.0m/s $A_r/A_d=0.5-9.1$ H=3.21m $D_d=0.142$ m	Internal and External Loop	$U_{lr} = \left[\frac{2gh_{D(\varepsilon_{r}-\varepsilon_{d})}}{K_{B\left(\frac{1}{(1-\varepsilon_{r})^{2}} + \left(\frac{A_{d}}{A_{r}}\right)^{2}\frac{1}{(1-\varepsilon_{d})^{2}}\right)}\right]^{0.5}$	
Gouveia <i>et</i> <i>al.</i> , 2003	Air Water NaCl.	U_{gr} =0.0126- 0.044m/s A_d/A_r =0.63	Internal Loop	$U_{lr} = 0.178 \left(\frac{U_{gr}}{\sqrt{gD_d}}\right)^{0.297} \left(\frac{t_c}{D_d}\right)^{0.004} \left(\frac{t_c}{D_d} + 1\right)^{0.095}$	
Kemblows ki <i>et al.</i> , 1993	Air Water Glycol Sugar Syrup CMC	$U_{gr}=0.001-$ 0.15m/s $A_{d}/A_{r}=1-1.33$	External Loop	$U_{lr} = \sqrt{\frac{2gh_D\varepsilon_r}{\frac{K_T}{(1-\varepsilon_r)^2} + K_B\left(\frac{A_r}{A_d}\right)^2 + 4h_D\left[\frac{f_r}{d_r} + \frac{f_d}{d_d} + \left(\frac{A_r}{A_d}\right)\right]}$	
Schlotelbur g <i>et al.</i> 1999	Air Water CMC	$U_{gr}=0.01-$ 0.06m/s $A_d/A_r=1.56$ $D_d=0.20$ m	Internal Loop	$U_{lr} = 0.25 U_{gr}^{0.33} \left(\frac{A_d}{A_r}\right)^{0.78} \cup_{app}^{-0.29} (1 - \varepsilon_s)^{0.74}$	

Table2.4. Mathematical models for liquid circulations inside airlift bioreactors

where, c and c_{∞} are the tracer initial and mean concentration respectively. The residence time distribution (RTD) is another method used to evaluate how well a solution is mixed and is usually based on the liquid recirculation in airlift bioreactor, however mixing time remains very popular and easier. Gavrilescu and Tudose (1998), Lu and Hwang (1994) and analyzed mixing using the axial dispersion model (ADM) which can be expressed in equation 2.10 as;

$$\frac{\partial C_r}{\partial \tau} = \frac{1}{B_o} \frac{\partial^2 C_r}{\partial Z^2} - \frac{\partial C_r}{\partial Z}$$
(2.10)

where, B_o is the Bodenstein number (also known as the Peclet number, P_e) VL/E_z , where E_z is the axial dispersion co-efficient (a higher E_z implies a lower B_o), C_r is dimensionless concentration C/C_{∞} and τ is dimensionless time t/t_c . V, the linear velocity (equation 2.11) defined as;

$$V_l = \frac{L_c}{t_c} \tag{2.11}$$

Using this method, a smaller B_o or P_e implies good mixing whereas B_o greater than 20 is considered as a plug flow; where the shorter the mixing time the higher the liquid velocity.

Depending on the property of the fluid used, a higher circulating rate of slurry might not necessarily achieve superior or better mixing. For example in a low viscous fluid, less agitation (low circulation) would be required to accomplish mixing whereas; a highly viscous liquid might require a higher degree of agitation (higher circulation). Bello *et al.* (1984) compared mixing in an external loop reactor ($0.11 \le A_d/A_r \le 0.69$) and in a concentric tube ($0.13 \le A_d/A_r \le 0.56$) and found that the circulation liquid velocity (U_l) was correlated to the cube root of superficial gas velocity of the riser in both bioreactors. However, the A_d/A_r ratio had greater influence. The external loop had a higher U_l due to a higher differential gas holdup between riser and downcomer but the concentric tube bioreactor was better comparing the data on mixing time.

In a three phase system, the effect of draft tube length (1.1 and 1.7 m, A_r/A_d ratio of 1), gas velocity and introduction of solids (polystyrene and calcium alginate) in an internal loop airlift bioreactor was evaluated (Lu *et al.*, 1994) by measuring tracer response and the estimation of B_o using the time domain analysis introduced previously (Verlaan *et al.*, 1989). They reported that the mixing was better at the bottom (B_o is 10-20) but it was excellent in the top section (B_o is 10). Furthermore, B_o in the riser and the downcomer were 20-30 and 40-70, respectively. They also found that the overall axial dispersion co-efficient (E_z) declined with increasing draft tube length although increased with increasing gas flow rate. Liquid mixing time in the polystyrene phase (two phase system) was shorter than that in a system with Calcium alginate. This was due to the fact that, bubbles attached to the polystyrene particles camouflaged it as a two phase system and enhanced mixing (Lu *et al.*, 1994). The D_d/D_c or A_d/A_r ratio influenced pressure loss in the vessel and hence, the circulation velocity varied on the extent of gas disengagement. However, gas disengagement did not support high mass transfer rate. Generally, the effect of the ratio of D_d/Dc or A_d/A_r on the liquid circulation velocity is difficult to predict because of interactions between gas, liquid and solid particles. A number of publications exist in this area and Koide et al. (1983) showed that when D_d/D_c decreased, the volumetric mass transfer increased. This was due to a decreased area for the gas holdup, which inhibited the liquid circulation velocity. However, Weiland (1984) disagreed, as in his study; an increase in the D_d/Dc was correlated to an increase in volumetric mass transfer. Furthermore, liquid circulation was higher when diameter ratio decreased to 0.6. These discrepancies could be attributed to the locations (clearance) of the draft tubes.

Molina *et al.* (1999) characterized mixing in a split cylinder airlift bioreactor (A_d/A_r ratio of 1, sucrose solution with viscosity variations of $1.54\pm19.5\times10^{-3}$ Pa.s, and U_g of 0 ± 0.039 m/s). It was reported that viscosity had no influence on circulation time, which contradicted the theory (increase in viscosity reduces flow as a result of resistance). According to them, the driving force of circulation has increased with increasing viscosity for any gas flow rate. This was due to the fact that as viscosity increased more bubbles were coalesced with a magnitude of smaller bubbles, where most of these large bubbles were disengaged at the top and smaller ones went through the downcomer. This achieved a higher driving force for liquid circulation. Viscosity had little effect on mixing time, which suggests that mixing time was affected by differences in velocities between the gas and liquid phases.

Merchuk et al. (1998) carried out an extensive study in a concentric tube reactor with seven different spargers (four cylindrical and three porous plates) of varying pore sizes using sea water and NaCl. They reported that the sparger pore size had an impact on the gas holdup and liquid recirculation. The smaller the pore sizes the higher the gas holdup which implied a decrease in the liquid circulation velocity. At a higher gas velocity, mixing time was independent of sparger geometry although the geometry of the sparger and pore size had an impact at a low gas velocity. Three different flow regimes were identified using the cylindrical sparger. The homogenous bubbly flow occurred as gas holdup increased with increasing gas velocity, while the transition flow occurred as coalescence of gas bubbles began. Finally, the holdup was affected by coalescing and not by the geometry of the sparger used. For the use of plane sparger, the transition flow was not very significant. They presented a correlation for mixing using the axial

dispersion co-efficient (equation 2.12) as a function of equivalent diameter, superficial gas velocity in the riser and gas holdup.

$$E_Z = K_5 D_c \left(\frac{U_{gr}}{\varepsilon}\right)^{n_4} \tag{2.1.2}$$

where, D_c is column diameter, K_5 and n_4 are 0.81 and 1.34 respectively. This correlation predicted satisfactorily (Glen *et al.*, 1993; Aoyama *et al.*, 1968) although for different constants, implying further research on effect of geometry on the constants.

Miron *et al.* (2000) tested mixing in a bubble column and airlift (split cylinder and draft tube) with a dispersion height of 2 m and working volume of 0.06 m^3 using water and seawater. They reported that, at any gas flow rate the values of mixing parameters in the two fluid media were identical. In all reactors mixing time decreased with increased superficial gas velocity. However, the bubble column gave the shortest mixing time due to the bulk flow as opposed to the airlift where circulation was in a cyclic motion impeding the bulk flow. A higher or rapid circulation (decreased circulation time) also enhanced mixing, although, at a superficial gas velocity of 0.02 m/s, gas bubbles increased and were entrained into the downcomer decreasing the liquid circulation rate. Although existing correlations for bubble column were in agreement with Miron *et al.* (2000), they proposed a correlation for airlift in terms of Bodenstein number, axial dispersion co-efficient and mixing time (see equations 2.13 and 2.14).

$$B_{OLG} = \beta (F_r^{0.33})^\alpha \tag{2.13}$$

$$B_0 = k \left(\frac{t_m}{t_c}\right) \tag{2.14}$$

where, B_{OLG} is Bodenstein number based on the superficial velocity, F_r is the Froude number, t_m is the mixing time, t_c is the circulation time and k, b and g are constants that depend on geometry of the reactor and fluid used. ($k=9.2\pm0.2$ for their work).

Since in convection, the movement of molecules within fluids occurs through diffusion, the conventional method (the pulse response technique where conductivity probes are used) cannot differentiate the two processes, since as a result of diffusion equilibrium is reached quickly upon addition of a tracer. Furthermore, to avoid the interference by the positions of the probes with the flow field, Luo *et al.* (2008) used an advanced imaging technique, computer automated radioactive particle tracking (CARPT) to investigate mixing in a draft-tube airlift bioreactor. CARPT is an advanced imaging technique that measures the flow filed tracking a radioactive particle. This particle is made to have a density equal to the slurry phase to be studied and the tracking is usually done with NaI scintillation detectors placed in the vessel. Flow measurements provide information for the analysis of flow field, mixing time and liquid circulation velocity.

Luo *et al.* (2008) have also used a draft tube sparged internal loop of 0.13 m diameter and 1.5 m height and a draft tube of 0.09 m diameter and 1.05 m height in an air-liquid medium with varying superficial gas velocities (0.00076–0.05 m/s) and varying TC and BC. The residence time distribution from flow trajectories obtained using the CARPT system was analyzed. As the superficial gas flow increased, the gas holdup also increased causing a difference for faster fluid flow and this concurred with Lu *et al.* (1994). They reported an effect of superficial gas velocity on liquid circulation velocity in terms of the circulation time. As superficial gas velocity increased, the circulation time also decreased (meaning higher liquid circulation) for different TC, however, at a faster rate for no top clearance (i.e.TC=0 cm). This implies that liquid flow

and mixing was enhanced at as gas-liquid separation was improved inducing a higher liquid velocity for circulation. However, a decrease in the bottom clearance did not affect the fluid circulation, meaning the friction loss did not change significantly. They also confirmed that flow in the riser and downcomer could be modeled as plug flow as suggested previously (Chisti, 1998; Lu *et al.*, 1994) by estimating the P_e number.

2.4. Applications of Airlift Bioreactors

Airlift reactors are used in various industries and amongst them, they are vital in biodegrading pollutants in municipal and industrial wastewater. Traces of pollutants in wastewater cause environmental hazard to animals, plants and humans. Some traces found in wastewater are phenol, sulfide, nitrogen etc. which are generated by chemical industries such as petroleum, plastics, textile and dyes. Mohanty *et al.* (2008) designed a multi-stage external loop airlift reactor for the removal of phenol from wastewater by means of its adsorption onto the surface of activated carbons. To enhance the adsorption of the trace element onto the carbon sites, the reactor was designed for continuous bubble formation, breakup and regeneration which promoted recirculation of the slurry. This design was good enough to remove about 95% phenol at a time with a lower carbon loading (2 g/L) which is typically 5 g/L for other wastewater removal processes.

The biofilm airlift suspension (BAS) bioreactor is a three phase system for biological wastewater treatment which became popular due to its high efficiency and low energy consumption (Heijnen *et al.*, 1993; Shieh, 1989) as opposed to the conventional activated sludge process. The BAS is made up of two concentrically placed columns with ceramic materials as carriers to support microorganisms (Zhou *et al.*, 2003). This system was used to treat domestic wastewater with an

effluent COD of 160-327 mg L^{-1} . In their experiments, two reactors with different sizes of carriers were used. Moreover, it was found that the reactor with a smaller diameter carrier achieved a higher concentration biomass of removing higher content of organic matter. This was due to the fact that smaller sized ceramic fillings provided a large surface area for biofilm attachment and biomass concentration. Generally a 95% COD has been removed by this method.

Nitrate removal from wastewater also employs the BAS, the first two processes ammonification and nitrifications are aerobic requiring oxygen whiles dentrification is anaerobic as performed by (Vilchez and Vega, 1995) using alginate beads as carriers to entrap *chlamydomonas reinhardtii*.

In the bioprocesses, for fermentation, microorganisms in an airlift bioreactor yielded better as compared to the achievement in conventional stirred tank bioreactor. In the stirred tank, the challenge of maintaining sterile condition for large cultures during scale up is enormous, whereas in the airlift reactors without the mechanical agitation had better aseptic conditions and oxygen supply. Ichii *et al.* (1993) developed a commercial scale internal loop airlift bioreactor of a volume of 145 m³ to grow *Candida utilis* for the production of RNA. For a higher production rate, the airlift fermenter was designed to increase the dilution rate by supplying more oxygen to the microorganisms. The cross sectional area ratio of the riser to downcomer was 0.8 with gas sparged in a draft tube containing perforated baffle plates. This design achieved an oxygen transfer rate of 9.9 kg-O₂/m³/h with yeast production of 9.79 kg-dry cell/m³/h which was greater than that achieved in a stirred tank reactor.

Aleksieva *et al.* (2000) successfully operated a fungus (*Humicola lutea* 120-5) without contamination for acid proteinase production under batch and continuous culture conditions. Acid proteinase has been predominantly used in the food industry as an enzyme for fermentation

in cheese production. The continuous system produced three times proteinase as compared to the batch system. Production data indicated comparable values for both stirred reactor and the airlift, although the cost was less in the case of the airlift process due to the low power requirement. Nakoa *et al.* (1997) used three reactors (internal loop, external loop and bubble column) in the production of gluconic acid with immobilized glucose oxidase (GO). During this process H_2O_2 was formed and deactivated glucose oxidase. Under optimal operational conditions and the reactor design, calcium alginate gel beads were used to entrap GO and H_2O_2 , while controlling the deactivation and the accumulation of H_2O_2 . They reported that the internal loop and bubble columns gave a higher production of gluconic acid with lower GO deactivation due to a higher K_La . They further developed a few models for the deactivation of GO which can be used to choose an optimum ratio of both M_nO_2 and GO for any gluconic production (Bao *et al.*, 2004).

Modified airlift reactor (static mixers) was also used to produce ethanol by Vincente *et al.* (1999). An increase of 30% production rate of ethanol was achieved in a three phase concentric tube reactor using *Saccharomyces cerevisiae* during glucose fermentation. This was due to smaller floc sizes produced with less gas flow rate, which promoted the dilution rate for the fermentation process.

The use of airlift has not been spared in the production of microalgae, which are useful biochemical substances as feed for human and aquaculture. Algae have been used for the production of biofuels subsequently contributing to reduce global warming. This organism utilizes light and CO_2 for its photosynthesis. In conventional reactors, since the light is introduced at a focal point, cells at the surface of the medium may capture higher photon flux density (PFD) while it tends to decrease away from the surface to the bottom. This inhibits the uniform distribution of light lowering the production. Ogbonna *et al.* (1996) proposed an

internally illuminated expensive stirred tank reactor. In light of this, Merchuk *et al.* (2000) used three reactors illuminated externally through the reactor wall (bubble column, an airlift reactor and a modified airlift reactor with helical flow promoters) for the cultivation of red microalgae *Porphyridium sp.* According to them, both airlift and reactor modified with helical flow had a higher production of biomass as compared to the bubble column. The airlift with its fluid circulation pattern encouraged an even distribution of light at a lower cost for CO₂. Earlier Merchuk *et al.* (1998) produced similar results for the same microalgae cultivation in both bubble column and airlift bioreactors, produced similar results at a higher gas input and a lower photon flux density (PFD). However, at a higher PFD and lower gas flow the airlift had better results.

Furthermore, Miron *et al.* (2000) studied the algae *Phaeodactylum tricornutum* using photobioreactors of bubble column, split-cylinder airlift and concentric draft-tube all with working volume of 0.06 m³ using Mediterranean seawater. There was no preference of one reactor over the other since in this case as they all produced an equal biomass concentration of about 4 kg/m³ after 260 h. Degen *et al.* (2001) investigated the production of *Chlorella vulgaris* in a rectangular airlift photobioreactor (with an illuminated front area of 0.084 m²). This airlift achieved 1.7 times higher production than the bubble column of similar dimensions.

Another area of growing interest is the production of materials and chemicals using different cell culture techniques. This would provide a reliable source of materials for production of drugs in the pharmaceutical industry, flavor for the food industry, and fragrances for the cosmetic industry in addition to other related chemical industries. Generally, airlift reactors provide uniform shear in cell cultures (though high shear would damage membrane and change the morphology), and provide good mixing. A glass airlift bioreactor with a working volume of 2.3L

was evaluated by Kim and Pedersen (1991) for cell suspension of *Thalictrum rugosum* which produced berberine. Berberine, which can also be derived from several plants, is used as an antimalarial drug to prevent or cure malaria. A production of berberine was much lower in the airlift bioreactor compared to the shake flask, although the cell yields were identical. Addition of ethylene and CO₂ in gas sparging increased the production level of berberine according to Kim and Pedersen, 1991. Wu et al. (2007) also produced caffeic acids derivatives in a high density of Echinacea purpurea without a product loss in 500 L and 1000 L airlift bioreactors as compared to a 20 L bioreactor using Murashige and Skoog (MS) medium. E. purpurea have been widely used in the production of caffeic acid derivatives (Wu et al., 2007). E. purpurea is a species of a medicinal plant having antiviral, antifungal, antioxidant and antibacterial substances that can be used for the treatment of various ailments. Its derivatives such as chichoric acid have been known to inhibit the type 1 HIV integrase and replication (Lin et al., 1999). Caspeta et al. (2005) tested the production of compounds having antifungal properties by Solanum chrysotrichum in shake flask, draught internal loop and novel modified meshed draught tube with mesh. A difficulty with this was that the hairiness and branching encouraged root entanglement that hindered mass and energy transfer processes. The airlift reactor was designed as such that the downcomer had a large space for root growth where the downcomer cross sectional area (A_d) was three times greater than the riser (A_r) . The modified reactor with helix provided superior performance. The mesh in the modified reactor allowed growth of root without flow obstruction, while the helixes promoted the even distribution of roots. With growing interest in genetic engineering, airlift reactors were exploited for protein recovery application. Dahman and Margaritis (2008) designed and used a draft tube fluidized bed reactor for protein bioseparation from bovine serum albumin (BSA) and bovine hemoglobin (BHb) solution. They achieved a

higher BSA adsorption from the solution due to the homogenous shear created by the airlift, thus reducing the compressibility of BHb. Bioseparation of BSA in a conventional adsorption processes was limited due to high adsorption of BHb.

The use of airlift bioreactor extends to other chemical and allied industries, such as petroleum in which toxic SO₂ and N₂ are released causing environmental pollution. In this case, biological processing of petroleum biodesulphurization (BDS) has been proposed which is not only less expensive (in terms of energy requirement) but also produces valuable products from the by product (Pachecoet *et al.*, 1999; Monticello, 2000). Mehrnia *et al.* (2004) simulated the chemical properties of fermentation broth used in BDS, using water-in-kerosene microemulsion system in a draft tube airlift bioreactor. They analyzed operating and geometric parameters on oxygen transfer and mixing characteristics. In BDS, where microorganisms were used, high oxygen levels for higher mass transfer rates were necessary for gas-liquid contacting. Mehrnia *et al.* (2004) reported an increase in A_d/A_r had superior mixing and oxygen transfer rate as opposed to an increase in the top section which depleted the system of oxygen.

2.5. Airlift Modification

Several modifications have been proposed to improve the conventional airlift bioreactor, and some of them have already been adapted. Modified airlift bioreactors include the inverse fluidized airlift, reactors with static mixers, helical flow promoters, and perforated draft tube; refer to Figure 6a, b, and c respectively. The principle for the inverse fluidized bed utilizes particles with densities lighter than liquid phase. These particles are entrained to the downcomer against the buoyancy of gas sparged. This inverse fluidized reactor has been used for the biological treatment of wastewater (Garnier *et al.*, 1990; Faraq *et al.*, 1997). In the process of biodegradation in the fluidized bed, the solids (microorganisms trapped in polymeric gel) are suspended in aqueous phase and gases are in contact with each other to provide oxygen for biodegradation.

Gluz and Merchuk (1996) investigated the helical flow promoter (HFP; Figure 6 (c)), which is consisted of fins and baffles that were placed either in the riser or downcomer. This caused fluid to flow in a helical pattern down the downcomer as opposed to the conventional type with straight downward flow. As the helical flow was generated, the content of the vessel began to swirl enhancing radial mixing in the downcomer. This contributed in better distribution of shear in addition to light and heat in the case of photobioreaction. Other characteristic of HFP is its capacity for fluidizing solid particles at lower gas flow rates. This is particularly favorable for operating dense cells immobilized on solids. Furthermore, mass transfer rates can be enhanced about 50% due to higher relative velocity between particles and liquids in such type of bioreactor (Merchuk *et al.*, 1993).

Since the mass transfer rate in the bubble column continued to be greater than in the airlift, the concept of integrating them has been exploited by the introduction of perforated or net draft tubes in concentric airlift reactors (Fig.2.6 (b)). Results of its hydrodynamic and process performance indicated an enhanced performance when compared with the conventional airlift.



Fig.2.6. Modifications of the airlift bioreactor.

This modified reactor had better mixing performance and higher oxygen transfer (Bando *et al*, 1992; Jong *et al.*, 1995). Cheng *et al.* (2002) used a draft tube with perforations in an airlift reactor to cultivate *Acetobacter xylinum* with an increased production level of bacterial cellulose nanofibers as compared to the bubble column. Cellulose nanofibers will be produced in the form of gels, which limits the amount of oxygen transfer inside the bioreactor. Furthermore, the shear

sensitive nature of the cells makes it a challenging task to produce this bacterium in the conventional stirred tank or in the bubble column. The superiority of the modified airlift with the draft tube was attributed to the ability of the draft tube to promote high oxygen transfer, which lead to improvement in the production of bacterial cellulose.

Wu *et al.* (2001) used the fungus *Absidia coerulea* CCRC 30897 to produce chitosan directly in a perforated draft tube and bubble columns. Chitosan is a polysaccharide produced by deacetylation of chitin and has unique properties that allow for its use in the biomedical industry. The modified airlift with perforated draft tube proved to be superior as it produced more chitosan with excellent oxygen transfer due to the proximity of chitosan to the cell walls of the fungus which enhanced the production of biomass. Wu *et al.* (2000) produced *Monascus purpureus*, a fungus that has an application in the production of certain fermented foods in China and Japan. Recently it has been discovered that properties of this mold lowers cholesterol in human bodies. Production levels of *M. purpureus* in the net draft tube far exceeded that in a bubble column as reported by Wu *et al.* (2000).

In view of cell damaged due to high shear, Garnier *et al.* (1990) proposed a concentric tube which combined both the fluidized bed and internal loop reactor. The setup ensured that, the gas, liquid and solid contact was somewhat limited to reduce shear damage to cells. Solid particles (i.e., polystyrene beads) were put in the annulus and gas was spared through the inner tube, whereas liquid flowed through the downcomer expanding the bottom, resulting in an inverse fluidized bed. However, lighter particles in addition to bursting of gas bubbles entrained into the downcomer can damage cells. In retrospect, Guo *et al.* (1997) used an external loop inverse fluidized bed airlift bioreactor (EIFBAB) to treat wastewater. In this type of bioreactors gas was sparged in the middle of the riser to aid fluidization for heavy particles in the bottom of the riser.

Loh and Liu (2001) employed a similar setup in a 4 L EIFBAB (shown in Fig. 2.7) to eliminate high contents of phenol in wastewater. However, a globe valve and stainless steel screen were used to prevent solid particles from getting into the downcomer. Polystyrene solid particles were used as support for immobilizing of *Pseudomonas putida*. The effect of the valve provided a range of gas holdup at a fixed gas input rate to the reactor. This system successfully biodegraded phenol up to 3000 mg/L. More on that, Loh and Ranganath (2005) integrated the EIFBAB with granular activated carbon (GAC) and successfully achieved cometabolic biotransformation of 4-chlorophenol (4-cp) in the presence of phenol.



Fig.2.7. The EIFBAB; external loop inverse fluidized bed airlift bioreactor

Another variation to enhance the performance of the conventional airlift was the introduction of static mixer in the riser. In such a vessel, liquid circulation was decreased as a result of an increased resistance from the static mixer.

Large gas bubbles were broken into smaller ones increasing the interfacial surface for higher mass transfer rates. Chisti *et al.* (1990), Potucek (1990), and Gavrilescu *et al.* (1997) have studied the influence of the static mixer in airlift bioreactors and observed an increase in K_La as compared to reactors without mixers.

Other modified types of airlift bioreactors include the convergent-divergent airlift (Fig.2.8a). In this type of airlift bioreactors the draft tube contracts and expands as the name suggests with continuous renewal of gas-liquid interfacial area to promote mass transfer. Wei *et al.* (1999) studied a three phase airlift bioreactors of air, water and resin. They reported an increase in both ε_g and $K_L a$ by 8% and 10%, respectively as compared to the conventional draft tube bioreactors with a lower liquid circulation due to the geometry of the convergent divergent vessel. The increase in ε_g even in the presence of solid loadings was related to the effect of the solids breaking up the gas bubbles into smaller ones to enhance mass transfer rates. They have further confirmed a viscous media (containing CMC) which also produced higher gas holdup and thus higher $K_L a$ (Wei *et al.*, 2000).



(a)Convergent - divergent

(b) multi stage airlift

Fig.2.8. Other modified airlift bioreactors

Mohanty *et al.* (2006) and Li *et al.* (2009) designed a novel multistage airlift system consisting of three vertically staged airlift reactors constructed to work in series (Fig. 2.8b). These designs are such that there is a continuous generation of bubbles, rupture and regeneration. This system was equipped with different spargers (multiple orifice and single orifice) together with a screen at various stages to encourage the continuous regeneration of bubbles. This system showed a 45% increment in gas holdup as compared to the single system. Mohanty *et al.* (2008) also applied this system to the removal of phenol in wastewater as have been mentioned previously.

2.6 Research Objectives

It is evident from literature surveyed that although several mixing experiments have been performed in the airlift bioreactors, discrepancies still exist mainly because of the various measuring techniques employed. With the relative ease of using the conventional methods where conductivity probes are inserted into the bioreactors, its downside however lies in their inability to visualize flow characteristics, disrupting flow and sometimes causing damage to sensitive cells thereby affecting accurate measurements in the bioreactor. Other visualization techniques may disrupt flow processes as well as alter the accuracy of measurements taken during an experiment. With growing interest and use of computational fluid dynamics (CFD) to simulate and validate processes, it is crucial that measurements taken during experiments are as precise as possible.

The electrical resistance tomography (ERT) has gained popularity over the last decade for investigating mixing especially in the stirred tank reactors and bubble columns. Quiet recently the ERT was employed to investigate the effect of sparger on gas holdup in bubble columns (Haibo et al., 2006) and also to study the hydrodynamics and flow in a three phase fluidized bed (Razzak et al., 2007). The main attractive feature of this measuring technique is that it is non invasive and non intrusive to flow processes, in addition to that it provides information on flow characteristics. The ERT technique measures conductivity variations within the reactor.

In view of the projected advantages of the ERT technique, this research work evaluated mixing performance in a draft tube airlift bioreactor with the following objectives:

• To investigate the feasibility of using ERT in a draft tube airlift bioreactor to enhance the knowledge of the characteristics of mixing inside airlift bioreactors.

- To determine the mixing time and circulation times which would be used in the calculation of liquid circulation velocity and shear rate in the draft tube airlift bioreactor.
- To evaluate the effect of the bioreactor geometry such as bottom clearances and draft tube diameter or ratio of cross sectional areas of downcomer to riser $A_d/A_{r,}$, in addition to gas flow rate, sparger configuration and liquid viscosity on the mixing characteristics.

CHAPTER 3

PROCESS TOMOGRAPHY

In an attempt to understand process related issues, several measuring techniques have been exploited. One of such is the development of computerized axial tomography (CAT) developed by Allan Cormack and Godfrey Hounsfield in 1970 as body scanners (Williams and Beck, 1995). This imaging technique has saved lives and proved reliable by visualizing the internals of the human body. To improve quality of products in industry; scientist and engineers have adapted and improved this technique over the last two decades to provide potentially detailed information on flow in chemical process units. The tomography technique measures signals from electrode located around a process vessel (Williams and Beck, 1995). The output of data obtained by these electrodes is then fed to the host computer to show variations of conductivity within a vessel. These variations of conductivity provide information on distribution of phases and mixing zones in a vessel.

Process tomography is a fast imaging tool and its sensing techniques or method of data collection is based on electromagnetic radiation, acoustic and electrical distribution inside the vessel. The daily or continual use of electromagnetic radiation is not only expensive but hazardous to our health. On the other hand, the cheaper and less expensive acoustic and electrical methods such as light, sound and electricity are alternatives for tomography techniques although the use of light for tomography is restricted to non-opaque vessels.

The electrical tomography provides distribution of electrical properties such as conductivity or resistance, magnetic inductance and capacitance (Table 3.1). Electrical impedance tomography (EIT), electrical inductance (magnetic) tomography (EMT) and electrical capacitance

tomography (ECT) produce images based upon the variations in the conductivity and permittivity. The electrical resistance tomography (ERT) is an example of EIT. The electrical tomography remain popular due to their simplicity, high speed capacity and most importantly can be used for real online imaging during manufacturing processes (Mann *et al.*, 1997).

3.1 Electrical Resistance Tomography (ERT) Configuration

The basic structure of the ERT system consists of the electrode hardware fitted around the periphery of a vessel, data acquisition system (DAS) and a host computer in which data processing software is imbedded for conductivity variation measurements (Fig 3.1).



Fig.3.1 Structure of a typical electrical resistance tomography system

Table 3.1	Electrical	tomography	technique
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Electrical methods	Principle realization	General remarks	Typical industrial applications
Capacitance (Dyakowski <i>et al.,</i> 1999)	Capacitance sensing electrodes which are non-invasive if separated from the process fluid by a 10 mm thick plastic liner	Electrode may need to be of 10 cm ² area to give sufficient capacitance change	Electrically insulating systems gas/oil, gas/solids two phase flow imaging fluidizing bed imaging
Resistance (Mann <i>et al.</i> , 1997)	Resistivity sensing, which are invasive but non-intrusive.	Very small electrodes can be used, Similar to system used for medical imaging.	Electrically conducting systems hydrocyclone imaging mixing study geophysical prospecting
Impedance (Lyon <i>et al.</i> , 1995)	Similar to above but without current injecting electrodes. Current is induced by coils surrounding the vessel.	Similar to above.	Ground water monitoring and soil remediation

Source: Williams and Beck (1995)

3.1.1 Electrodes

The electrodes for the ERT system are fitted inside the vessel non invasively for contact with process fluids to allow for the continuous detection of conductivity changes without invading the process flow. This hardware is usually made up of 16 equally spaced rectangular electrodes. The arrangement of the electrodes can either be circular or vertical. For reliable measurements, the electrode hardware must be more conductive than the process fluid (Williams and Beck, 1995). Also to withstand harsh operating conditions such as temperature and pressure the electrodes are made of metals (stainless steel, silver, gold, platinum etc). These electrodes are connected to the data acquisition system by means of co-axial cables to reduce the noise generated by electromagnetic transmissions associated with typical operation processes (Dickin and Wang, 1996).

3.1.2 The Data Acquisition System (DAS)

Fig.3.2 shows a schematic diagram of the ERT data acquisition system, this is a basic necessity for an ERT system. The DAS is attached to the electrodes and connected to the host computer to display signals of conductivity variations. It is responsible for applying current and recording its corresponding voltage output revealing conductivity distribution in a vessel (Williams and Beck, 1995). The DAS incorporates signal measurements, de-modulation and control, waveform generation and synchronization, multiplexer control and power supply (Dickin and Wang, 1996).

This system starts with the sine wave generator; here current is applied to produce a sine wave to probe the material under investigation (Brown *et al.*, 1994). An electrical current application is

preferred for sine generation because a good current source has high output impedance whereas a voltage source has a low one (Dickin and Wang, 1996).

Digital function generators in the form of EPROM- based staircase with lower harmonic distortions have replaced monolithic generators (Brown and Seagar 1985). For further filtration of harmonics, the digital waves are then converted to an analog- voltage by a high speed digital-to- analogue converter (DAC) and fed into a voltage current converter source (VCCS).



Fig.3.2. Components of a typical data-acquisition system (Holden et al., 1998)

The multiplexers share the current source and voltage measurement stages between any numbers of electrodes. Depending on the strategy used to apply electric current, a wide range of voltages maybe produced which the voltage measurements might not be able to accommodate. The amplifier solves this problem by rejecting the common mode signals as electric noise.

This occurs through the grounded (load) floating measurement (GFM) technique. Here a GFM lead is attached to one of the unused electrodes as a ground electrode on the vessel. The other end of the lead is attached to the output of the common-mode voltage. The effect of this arrangement is to produce a zero potential in the conducting region inside the vessel and therefore the high common-mode voltage is removed from the current-driven electrodes (Wang *et al.*, 1993). An optimum signal-to-noise (SNR) is obtained through the digital demodulation; finally the waves go through an analogue-to-digital converter (ADC).

3.1.2.1 Data Collection Strategies

With a reliable DAS system in place, the next step is the interrogation protocol for probing the conductivity distribution within the vessel. These include the adjacent, conducting boundary, diagonal and opposite measurements strategies (Fig.3.3). The interrogation protocol involves the application of electric current through a set of electrodes and the detected voltage measured from the other electrode combinations. For a typical ERT system, the most common measuring protocol is the adjacent method. The adjacent protocol involves the application of electric current through a pair of adjacent electrodes and the detected voltages measured from successive pairs of neighboring electrodes (Fig.3.3a). The procedure is repeated for all other combinations of adjacent pairs of electrodes for the entire ring of electrodes (Seagar *et al.*, 1987).

This protocol yields M independent measurements which can be calculated from:

$$M = \frac{N(N-3)}{2}$$
(3.1)

where N is the number of electrodes. The adjacent measuring protocol remains popular because



Fig.3.3. Data collection strategies: (a) The adjacent measurement strategy, (b) The conducting boundary measurement strategy, (c) The diagonal measurement strategy, (d) The opposite measurement strategy. Adapted from Dickin and Wang, 1996.

it requires minimal hardware for implementation with a relatively quick conductivity signal output recording (Abdullah, 1993). A downside however is the non uniform current distribution as a result of measurements at the periphery making this technique very sensitive to measurement errors and noise (Brown *et al.*, 1994).

3.1.3 Host Computer

Following acquisition of the voltage measurements from the electrode, an algorithm imbedded in the host computer is used to display the average conductivity distribution within the vessel. Average conductivity variations generated from the algorithm determines the internal distribution of conductivity within the vessel. The ITS 2000 uses a non-iterative algorithm known as the linear back projection (Barber *et al.*, 1983 and Kotre, 1989) which converts voltage measurements to conductivity values. This algorithm is quick, simple with low computational time as average conductivity readings or signal are generated by simply multiplying the measurements by a single, pre-calculated matrix (Holden *et al.*, 1998).

3.2 Applications of ERT

Introduction of body scanners in the medical industry have enhanced diagnostics and treatment. Other industries have exploited this possibility using the ERT without the multimillion dollar price tag to it. In other chemical engineering applications, the ERT has been applied to evaluate mixing performance in various reactors (Fransolet *et al.*, 2001; Razzak *et al.*, 2007; Pakzad *et al.*, 2008). Mixing evaluation is usually quantified by the mixing time and this is determined by the variation of conductivity values in the ERT system. Mixing efficiency or performance may depend on time, reactor geometry, fluid viscosity and impeller type. ERT technique enhances knowledge during mixing process. This knowledge can be used to improve the design of equipment and processes. GlaxoSmithKline a pharmaceutical company used the ERT for the online control of an active pharmaceutical ingredient (API) synthesis for paracetamol production (Ricard *et al.*, 2005). This technique enabled them evaluate the mixing performance of several reactor geometries and concluded that the ERT was valuable for development of API processes.

For batch processes, ERT was used to monitor the progress during nylon polymerization (Dyakowski *et al.*, 2000). To withstand harsh operating conditions involved in this process (i.e. high temperatures and pressures), specially designed electrodes were fabricated using plasma technology and fitted into the walls of the vessel. Results obtained using the ERT techniques provided useful information on the various stages of nylon polymerization. The ERT showed the material distribution during the heating up stage, the effect of opening the pressure control valve and the permittivity distribution at the last stage of the polymerization process (Dyakowski *et al.*, 2000).

In the past, interpretations of the mechanisms of separating fluids in hydrocyclone were inadequate due to; swirling, turbulence and the lack of on-line measuring techniques to provide reliable experimental data for modelling (Dyakowski and Williams, 1998). To monitor the separating process in hydrocylones, ERT was employed to investigate clay refining (Williams *et al.*, 1999). Experimental measurements were performed in a unit containing multiple 50 mm diameter hydrocylone separators. The electrode hardware consisted of 8 planes of 16 electrodes. These electrodes were disc shaped and engineered to be flush with the inner walls of the vessel for continues measurements. Results indicated the possibility of identifying different (real) fault conditions based on variations in conductivity readings. This information provides a rich source of data for model development in fluid separation (Williams *et al.*, 1999).

Pressure filtration is a process for separating of liquid from solid phase. Processes involved are lengthy and can extend to a number of hours or days. Due to the slow dynamics of the reaction, modest data acquisition rates of order one frame per minute of the ERT system are adequate. Vlaev *et al.*, 2000 used ERT, to monitor progress during washing and drying stages of pressure filtration at Zeneca. Their aim was to identify the end point of filtration such that processing times would be improved and waste reduced. Results obtained with the ERT technique showed movement of liquid level during filtration. Also any tilt of the filter which causes a malfunction of this process was also detected. They concluded that ERT sensing electrodes could be retrofitted to a large scale filter without modifying the internals of the unit.

CHAPTER 4 EXPERIMENTAL

4.1 Experimental Setup

Fig. 4.1 shows a schematic diagram of the bioreactor setup with a constant volume of 0.0578 m^3 . The draft tube airlift bioreactor consists of two co-axially mounted concentric tubes. The first concentric tube has an inside diameter of 0.40 m with a height of 0.60 m. The draft tubes examined have inner diameters of 0.22, 0.29 and 0.34 m. For mixing time studies, the BC (distance from bottom of the draft tube to base of the bioreactor) were adjusted to 0.003, 0.006 and 0.009 m. A sparger was placed just at the entrance of the draft tube (Chisti *et al.*, 1989) and the air flow controlled by a rotameter attached to the side of the bioreactor. The superficial gas velocity varied between 0.00165 to 0.00807 m/s, which is the typical range used for cell culture cultivation in industry (Grima *et al.*, 1997). Two sparger configurations shown in Fig.4.1 were used in this work: the cross shaped and circular shaped configurations.

4.1.1 Airlift Bioreactor

In order to evaluate the performance of the airlift bioreactor with different geometries, three sizes of draft tubes were used in this experiment. They consist of a bigger cylindrical column with internal diameter of 0.4 m and a height of 0.6 m. These draft tubes were of internal diameters of 0.22, 0.29 and 0.34 m, which is equivalent to A_d/A_r ratios of 2.31, 0.9 and 0.38 respectively. Geometrical parameters were calculated from:

Cross sectional area of downcomer;
$$A_d = \pi/4 (D_c^2 - D_d^2)$$
 (4.1)
Cross sectional area of riser; $A_r = \pi/4 (D_d^2)$ (4.2)

$$A_d / A_r = \frac{\left(D_c^2 - D_d^2\right)}{D_d^2}$$
(4.3)



Fig. 4.1 Bioreactor set up

4.1.1.1 Sparger Design

To prevent weeping (back flow) in the sparger, Mersmann (1978) proposed that the weber number (W_e) based on the hole diameter (1-5 mm) of the sparger to be 2. For the calculation of number of holes on the sparger, the weber criterion is used to calculate the superficial gas velocity through the holes (Ruff *et al.*, 1978). i.e.;

$$W_e = \frac{\rho_G d_0 U_G^2}{\sigma_L} = 2 \tag{4.3}$$
$$U_G^2 = \frac{2\sigma_L}{\rho_G d_o} \tag{4.4}$$

where d_o is diameter of hole on sparger, ρ_G is the density of gas, U_G superficial gas velocity in riser. The highest gas velocity (0.00807m/s) was calculated to find the maximum number of holes needed on the sparger based on the following parameters:

 $d_o = 2$ mm, $\rho_G = 1.206$ kg/m³ and $U_{gr} = 0.00807$ m/s.

$$U_G = \sqrt{\frac{2 \times 0.0724}{1.206 \times 0.002}}$$
$$U_G = 7.75 \, m/s$$

Since Overall Volumetric flow rate of air = Volumetric flow rate of air through holes in sparger

$$\frac{\pi}{4}D_c^2 U_{gr} = \frac{\pi}{4}d_o^2 U_G N$$

Number of holes, $N = \frac{D_c^2 U_{gr}}{d_o^2 U_G} = \frac{0.39^2 \times 0.00807}{0.002^2 \times 7.75} = 39$ holes

4.1.2 ERT System

The ERT setup for this experiment consists of the electrodes, ITS P2000 data acquisition system (DAS) and a host computer (Fig.4.2).

4.1.2.1 Electrodes

The wall of the bioreactor was fitted with four planes of electrodes (Fig.4.2). Each plane of electrode consisted of 16 equally spaced stainless steel electrodes. Each rectangular electrode was dimensioned as 20 mm high by 30mm wide with 1 mm thickness. The planes were

numbered from top to bottom (1-4) and at intervals of 8.5 cm for the first three planes and the last plane was 4.5 cm from the bottom of the bioreactor. Since the electrodes were fitted on the periphery of the bioreactor, changes in conductivity within the bioreactor were observed only in the downcomer (annulus). The electrodes were connected to the ITS P2000 data acquisition system via co-axial cables.

4.1.2.2 Data Acquisition System (DAS)

The DAS was responsible for applying current and measuring the corresponding voltage output. Data collection was performed quickly to track changes in conductivity variations in the bioreactor after the injection of a tracer. The ITS P2000 used the adjacent measurement protocol by applying electrical current between an adjacent pair of electrodes and measuring the voltage difference between all other adjacent electrode pairs.



Fig.4.2 ERT setup for experimental work

The procedure was repeated for other combinations of adjacent electrode pairs for a full rotation of the vessel. The main specifications of the DAS used in this study were: frequency, 9600 kHz and injecting current 75 mA.

4.1.2.3 Host Computer

The DAS was connected to a host computer (Fig.4.2) through a communication port for data collection and storage. It used a non-iterative algorithm (Linear Back Projection-LBP). The algorithm shows distribution of conductivity inside the annulus of the bioreactor. This algorithm requires voltage difference measurements before and after any change in conductivity to convert the voltage measurements into average conductivity readings or signals on the various planes of electrodes. The LBP is easy, quick and simple due to its low computational requirement (Barber and Brown, 1984).

4.2 Materials and Supplies

Xanthan gum (*XG*) powder was used for the preparation of aqueous xanthan gum solution. Xanthan gum is a highly viscous polysaccharide solution (Davidson, 1980) produced by fermentation of *Xanthomonas campestris bacterium* and has found its application in food, petrochemical, pharmaceutical and oil recovery industries. Food grade *XG* powder used was obtained from Archer Daniels Midland Company (Decatur, IL). Household salt (NaCl) was purchased for use as the tracer; sugar and tap water were also used.

The concentrations of sugar solution and xanthan gum solutions were chosen to approximate those found in fermentation media (Onken and Weiland 1980; Fields *et al.*, 1984). 34.5% sugar solution was prepared by dissolving the right amount of sugar into water. To prepare 0.2% and

0.5% xanthan solution for a volume of 0.0579 m³, appropriate masses of *XG* powder was weighed. The mixing tank was then filled with some tap water. With the stirrer on, xanthan gum powder was slowly added to the water in the tank. Water was added up to the desired volume and the resulting solution was stirred continuously for about 6 hours to achieve a homogenous solution. The density of the resulting solution was measured using a pycnometer. Rheological properties of *XG* were determined (Table 4.1) using a Bohlin CVOR Rheometer 150 (Malvern instruments, USA) and correlated according to the Hercshel-Bulkley model (Herschel and Bulkley, 1926) to determine viscosity. All measurements were taken using the Bohlin Mooney-Cell and Cup geometry (25mm diameter, 29mm height, stainless steel) at 22°C. The rheology of *XG* solutions was modelled best by the Hercshel-Bulkley model as shown in Fig.4.3. Then the viscosity is given by:

$$\mu = \frac{\tau_y}{\dot{y}} + K |\dot{y}|^{n-1} \tag{4.5}$$

The average shear $\dot{\gamma}$ for viscosity calculation is evaluated based on the correlation proposed by Nishikawa *et al.* (1977);

$$\dot{\gamma} = 1000 U_a^{0.5} \tag{4.6}$$

Concentration (%)	Consistency index, K (Pa s ⁿ)	Power-law index, <i>n</i>	Yield stress, τ _y (Pa)	Density,p (kg/m ³)	Viscosity, µ (Pa.s)
0.2 <i>XG</i>	0.0849	0.9509	1.864	1002	0.0889- 0.1167
0.5 XG	0.3408	1.169	4.506	1006	0.7482- 0.7790
34.5 Sugar	-	-	-	1089	0.004
Water	-	-	-	998	0.001

Table 4.1. Rheological properties of xanthan gum solution



Fig.4.3. Shear stress versus shear rate for xanthan solutions

4.3 Experimental Procedure

4.3.1 Mixing Time Measurements

The bioreactor with draft tube diameter of 0.22 m was first fitted with the cross shaped sparger and the bottom clearance (BC) set at 0.003 and filled with tap water to a volume of 0.0579 m³. The conductivity of the water was then measured using a conductivity probe and this value was fed to the ERT system. Air was introduced through the system by the sparger at a superficial velocity of 0.00165 m/s for about 25 seconds till steady state was reached. Mixing time measurements were performed in the downcomer (annulus of the bioreactor) by the addition of 50 ml saline tracer (NaCl). Tracer injection was performed in the draft tube diameter at about 2 cm above the liquid level. Mixing was monitored by means of variations in average conductivity signals that are generated by the ERT system. Once these changes in average conductivity of all four planes of electrodes were less than 10% (Fig.4.4) data collection was discontinued. Gas flow into the bioreactor was shut off and water was drained out. For the same superficial gas velocity, the above procedure was repeated for two other bottom clearances (0.006 and 0.009 m). This process was then repeated for four other superficial gas velocities (0.00331, 0.00471, 0.00667 and 0.00807 m/s) all in the 0.22 m diameter draft tube. For the second and third draft tubes with internal diameters of 0.29 m and 0.34 m respectively, the above procedure was repeated for all gas velocities and bottom clearance of 0.009 m using the cross shaped sparger. The effect of the circular sparger was also evaluated by fitting it to the 0.29 m diameter draft tube with a constant bottom clearance of 0.009 m for all gas velocities.

Mixing time measurement for 34.5% sugar, 0.2% and 0.5% XG solutions were performed in draft tube with internal diameter of 0.29 m for all gas velocities but at a bottom clearance of 0.009 m using the cross shaped sparger configuration.

Mixing time is the time required to reach a certain steady state conductivity reading after the injection of a tracer, usually with less than 10% deviation (Onken and Weiland, 1980).

$$t_m = \frac{c - c_\infty}{c_\infty} \times 100\% \tag{4.7}$$

Mixing time (degree of homogeneity) after the injection of a tracer into the riser is obtained in the downcomer by the plot of ERT data (Fig.4.4).

Fig.4.4 shows the changes in the water conductivity with time at $u_{gr} = 0.00807$ m/s in bioreactor with draft tube diameter of 0.22 m and bottom clearance of 0.009 m after the injection of tracer. Examining Fig.4.4 shows that there is no change in conductivity readings from 0 to about 26 seconds in the absence of a tracer.



Fig.4.4: The change in conductivity with time at $U_{gr} = 0.00807$ m/s in draft tube diameter of 0.22 m with a bottom clearance of 0.009 m (air-water system).

However, after the injection of tracer at about 26 seconds, electrodes of the ERT system picks up the changes in conductivity readings until there is a certain level of homogeneity at time 55 seconds (less than 10% deviation of the conductivity readings) to determine the mixing time.

4.3.2 Gas Holdup Measurements

Due to the large diameter of the bioreactor used, a flat circular cover of about 0.385 m was constructed to fit directly into the bioreactor. This cover had an extension of a cylinder up directly in the middle of the cover to make readings of dispersion easier (Fig. 4.4). Gas Holdup measurements were conducted using the height of dispersion method (Merchuk, 1990). The bioreactor with draft tube of 0.22 m fitted with a cross shaped sparger was first used with the bottom clearance set at 0.003 m. The bioreactor was then filled with tap water to the required

volume of 0.0579 m³ and the level or height of the liquid (h_L) was recorded (Fig.4.5). Air was then introduced through the system by the sparger at a superficial velocity of 0.00165 m/s till steady state was reached. By visual inspection, the height of dispersion or liquid level (h_D) due to aeration was recorded and the gas flow was shut off. The above procedure was repeated for two more bottom clearances (0.006 and 0.009 m). This process was then repeated for four other superficial gas velocities (0.00331, 0.00471, 0.00667 and 0.00807 m/s) all in the 0.22 m diameter draft tube. The draft tube diameters of 0.29 m and 0.34 m were fitted with the cross shaped sparger, and the above procedure was repeated for all gas velocities and constant bottom clearance of 0.009 m. Again for the second draft tube of 0.29 m the same set of experiments were performed only this time using the circular shaped sparger.



Fig.4.5 Bioreactor setup for gas holdup measurements

Gas holdup measurements using 34.5% sugar solution, 0.2% and 0.5% XG solutions were performed in the draft tube of 0.29 m fitted with the cross shaped sparger with the bottom clearance set at 0.009 m for all five superficial gas velocities.

The overall or total gas holdup measurement is:

$$\varepsilon_{gT} = \frac{h_D - h_L}{h_D} \tag{4.8}$$

According to Chisit, 1989, the total gas holdup is related to the holdups in the riser and downcomer, thus;

$$\varepsilon_g = \frac{A_r \varepsilon_{gr} + A_d \varepsilon_{gd}}{A_r + A_d} \tag{4.9}$$

For internal loop airlift reactors without gas-liquid separator, the gas holdup in the downcomer is reported (Chisiti, 1989) as;

$$\varepsilon_{gd} = 0.9\varepsilon_{gr} \tag{4.10}$$

Further, with the cross sectional areas of the riser and downcomer known (i.e. for draft tube diameter of 0.22 m, $A_d = 0.0878$ and $A_r = 0.038$), equations (4.7-4.9) could be combined to yield the following expression for gas holdup in the riser;

$$\varepsilon_{gT} = \frac{0.0380\varepsilon_{gr} + 0.0878(0.9\varepsilon_{gr})}{0.0380 + 0.0878}$$
(4.11)
$$\varepsilon_{gr} = \frac{0.1258\varepsilon_{gT}}{0.1170}$$

Gas holdup in the downcomer is then calculated using equation (4.10) from Chisit's correlation (1989).

With experimental data obtained from mixing time measurements and gas holdup, the following parameters were calculated;

4.4. Calculation of Liquid Circulation Velocity

Liquid velocity was calculated using data obtained from circulation time t_c (Fig4.6). The circulation time is defined as the time taken for the tracer to circulate the draft tube (Blenke, 1979) or the time difference between adjacent conductivity peaks recorded (Onken and Weiland, 1980). The liquid circulation velocity was then calculated as follows (Schlotelburg *et al.*, 1999);



$$U_{ld} = \frac{2n_d}{t_c} \left(1 - \varepsilon_{gd}\right) \tag{4.12}$$

Fig.4.6 Conductivity versus time at U_{gr} =0.00165 m/s in draft tube diameter of 0.29 m at 0.009 m bottom clearance on plane 1 of ERT system (air-water system) using circular shaped sparger.

Fig.4.5 shows the change in water conductivity with time at $U_{gr} = 0.00165$ m/s in draft tube diameter of 0.29 m for a bottom clearance of 0.009 m after the injection of tracer. The graph

represents the circulation time of the tracer. The 4 planes response time for the ERT electrode system was short and gave constant readings. Therefore, only plane 1 of the ERT electrode system was used for circulation time measurements. A similar method was used by Shi *et al.*, 1990.

4.5 Calculation of Shear rate

The definition of shear stress previously employed (Merchuk and Berzin, 1995) was modified for airlift (Grima *et al.*, 1997). Knowledge of energy dissipation rate and fluid residence time in each zone is used to calculate the shear rate in that zone.

$$\tau_d = \frac{E_d t_{cd}}{h_{ld} V_d a_d} \tag{4.13}$$

where τ_d , E_d , h_{ld} , V_d , and a_d values of shear stresses in the downcomer, energy dissipation rate downcomer, height of liquid level, volume of liquid and specific interfacial area for bubbles all in the downcomer respectively.

$$E_d = \rho_l g h_D A_d u_{ld} \varepsilon_{gd} \tag{4.14}$$

Yomoshimoto *et al.*, 2007 proposed the following equation for specific interfacial areas for bubbles; where μ_L is the viscosity of the fluid

$$a_d = 323\varepsilon_{gd}\mu_L^{-0.13} \tag{4.15}$$

Finally the shear rate for Newtonian fluids (water and sugar solution) is calculated as;

$$\dot{\gamma}_d = \frac{\tau_d}{\mu_L} \tag{4.16}$$

Shear rate calculations were performed using average values of the various parameters needed (ie. gas holdup, circulation time, liquid circulation velocity).

In the case of the non Newtonian xanthan gum solutions which obey the Herschel Bulkley model (Herschel Bulkley, 1926), the shear rate is calculated using the following equation;

$$\tau = \tau_v + K \dot{\gamma}^n \tag{4.17}$$

All experiments were repeated three times (expect for xanthan gum solutions) and each result is an average from three mixing time values. The error bars representing standard deviations indicate the spread of the mixing time about the mean value for water and 34.5% sugar solutions. The maximum standard deviation was 2, confirming the high accuracy of the ERT system for mixing time measurements.

Generally, the standard deviation is defined as:

$$s = \sqrt{\frac{1}{N} \sum_{i=1}^{N} (x_i - \bar{x})^2}$$
(4.18)

Where s is standard deviation; N denotes the number of samples taken; x represents the value of the mixing time; and \bar{x} is the average value of the mixing time.

4.6 Experimental Conditions

The experiments were performed at room temperature between at 21 (±1 °C). As summarized in Table 4.2, the working fluids were Newtonian fluids represent by tap water ($\rho = 998 \text{ kg/m}^3$; $\mu = 0.001 \text{ Pa} \cdot \text{s}$, $\sigma = 0.0724 \text{ kg/s}^2$; Okada *et al.* 1995), salt solution (5M NaCl), 34.5% coalescing sugar solution and non-Newtonian fluids represented by *XG* solutions of 0.5% and 0.2%. The tracer

was 50 mL of 5M NaCl solution for each test. The tracer injection was performed by a single syringe and the injection point was in the riser (inside the draft tube) whiles results from ERT were obtained in the downcomer (annulus).

	Setup 1	Setup 2	Setup 3
D_d (m)	0.22	0.29	0.34
A_d/A_r	2.31	0.90	0.38
BC (m)	0.003,0.006,0.009	0.009	0.009
System	Air-water	Air-water 34.5% Sugar 0.2, 0.5% <i>XG</i>	Air-water
Sparger Configuration	Cross shaped	Cross shaped Circular shaped	Cross shaped

Table 4.2 Experimental conditions

CHAPTER 5

RESULTS AND DISCUSSION

5.1 Mixing Time

The effect of gas flow rate, bottom clearance, draft tube diameter, sparger configurations and fluid viscosity on mixing time, liquid circulation velocity, gas holdup and shear values were examined. Experiments were conducted using three different bottom clearance (BC= 0.003, 0.006 and 0.009 m), and five different superficial gas velocity in riser, ($U_{gr} = 0.00165$ -0.00807 m/s). The ratio between downcomer and riser cross sectional areas (A_d/A_r) were varied between 2.31 - 0.38 for draft tube with internal diameter of 0.22, 0.29 and 0.34 m respectively.

5.1.1 Effect of Gas Flow Rate and Bottom Clearance

Mixing time studies is of particular importance in a typical fermentation media to prevent formation of products outside cells within the fermentation broth which compromises product quality. Supply of nutrients and oxygen takes place within cells in the fermentation broth therefore; once mixing is completed products begin to form out of the cells. And also for the mix of solids, segregation may occur as a result of excessive mixing. Fig.5.1.1 shows the effects of superficial gas velocity in the riser (U_{gr}) on mixing time at different bottom clearances (0.003, 0.006, and 0.009 m) with a draft tube diameter of 0.22 m. Results in Fig.5.1.1 demonstrates that the mixing time decreased with the increase in superficial gas velocity in the riser for all different bottom clearances examined. The decrease in t_m with increasing U_{gr} occurs as a result of the increase in turbulence in the bioreactor. This may be attributed to the fact that an increase in U_{gr} corresponds to an increase in energy generated, resulting in an increase in gas holdup which creates a higher driving force for liquid circulation velocity. It is interesting to note that at a higher gas flow rate, the effect of U_{gr} on mixing is not as efficient since the mixing time becomes a plateau (Fig.5.1.1). It has been established that the specific power input increases when superficial gas velocity increases (Chang *et al.*, 1993). An increased specific power input without significantly improving mixing suggests that selecting a proper aeration rate is important for bioreactor. A higher gas flow rate consumes more energy but does not necessarily improve mixing. On the other hand, higher gas flow rate may cause higher shearing which perhaps cause damages to microorganisms during fermentation.

Fig.5.1.1 also shows mixing time (t_m) at various bottom clearances (BC). According to this Figure, t_m decreases as the BC increases. It is within the bottom clearances that the moving liquid from the downcomer makes a 'u-turn' into the riser. This can be explained based on the fact that as these distances shortens; the greater is the impedance to the liquids momentum caused by a loss of kinetic energy and hydraulic friction (Petersen and Margaritis, 2001; Luo and Dahhan, 2008). This observation is in agreement with results obtained by Gavrilescu and Tudose, 1999; Gouveia *et al.* (2003); Luo and Dahhan, 2008. The above results are clearly in agreement with Bando *et al.* (1998) who also reported that mixing time decreased as the bottom clearance increased. However, for a larger bottom clearance, part of the gas from the sparger can get directly into the downcomer if the sparger is not placed properly which reduces the driving force and liquid velocity.



Fig.5.1.1 Effect of bottom clearance (BC), on mixing time for all three bottom clearances in draft tube diameter of 0.22 m (air-water system) using cross shaped sparger configuration.

5.1.2 Effect of Draft Tube Diameter

In Fig.5.1.2, comparisons of mixing time using different draft tube diameters (0.22, 0.29, 0.34 m) are shown as a function of superficial gas velocity in the riser (U_{gr}) with bottom clearance fixed at 0.009 m using the cross shaped sparger. The results in Fig.5.1.2 clearly demonstrate decrease in mixing time with the increase in draft tube diameter. The increase in draft tube diameter results in lower ratio of A_d/A_r , i.e. a decrease in cross sectional area of the downcomer with higher circulation rates. This indicates that an increase in draft tube diameter improves the mixing time and circulation in the downcomer, which enhances the entrainment of gas at a faster rate. This could be attributed to the fact that restrictions of flow are encouraged by small A_d/A_r

ratio (Gavrilescu and Tudose, 1999). These results are in agreement with results previously published by Mehrnia *et al.*, 2004.



Fig.5.1.2. Effect of draft tube diameter on mixing time at BC = 0.009 m (air-water system) using cross shaped sparger configuration.

5.1.3 Effect of Sparger Configurations

Fig.5.1.3 represents the effect of sparger configurations on mixing time for different superficial gas velocity (U_{gr}) in a draft tube with diameter of 0.29 m fixed at bottom clearance of 0.009 m. As shown in Fig.5.1.3, mixing time was lower for the cross shaped sparger configuration than the circular sparger. Mixing time decreased to about 40% in air-water media using the cross shaped sparger. It should also be noted that mixing time shows a steeper decrease in the low

range of superficial gas velocity, followed by a milder decrease as U_{gr} gets higher for both sparger configuration.



Fig.5.1.3. Effect of sparger configuration on mixing time at BC = 0.009 m in draft tube diameter of 0.29 m (air-water system).

Nonetheless, as U_{gr} increased from 0.00607 m/s the effect of the configuration was minimal and would probably diminish in the heterogeneous regime. In the heterogeneous regime, mixing is turbulent consisting bubbles of varying sizes. In this case, the gas holdup is low, since, too many large bubbles tend to occupy the entire volume of the reactor and would encourage bubble coalescence subsequently limiting mass transfer rates. This shows that the cross shaped sparger had produced a greater density difference between the riser and downcomer resulting in greater mixing efficiency at lower superficial gas velocities.

5.1.4 Effect of Fluid Viscosity

The influence of fluid viscosities on mixing time in a 0.29 m draft tube diameter using cross shaped sparger is shown in Fig.5.1.4. Different fluid systems of different viscosities were examined here, which include Newtonain solutions (water and solution containing 34.5% sugar), and non- Newtonian solutions (0.2% and 0.5% xanthan gum). It is evident from Fig.5.1.4 that mixing time in less viscous Newtonian fluids (water and 34.5% sugar solution) is lower than that in highly viscous non-Newtonian fluids (0.2% and 0.5% xanthan gum solutions). The effect of viscosity on mixing time is significant especially in the case of the viscous non-Newtonian xanthan gum solutions as shown in Fig.5.1.4. Although results of mixing time for water and sugar solution are not as appreciable as in the case of xanthan gum, there is a reduction in mixing time observed for 34.5% sugar solution. This is attributed to the occurrence of bubble coalescence behavior of sugar solution (Prince and Blanch, 1990). As gas is introduced into the bioreactor, bubbles begin to form and coalesce reducing gas holdup leading to a reduced driving force for liquid circulation velocity and mixing time. For xanthan gum solutions of 0.2% and 0.5%, mixing time significantly decreased as a result of increased viscosity. This increased fluid viscosity drag hinders the liquid circulation (Fields et al. 1984) which results in an increase in mixing time. The friction factor is substantially lower and this drag reduction (friction) usually increases with the flow rate and xanthan gum molecular weight (Fields *et al.* 1984).



Fig.5.1.4. Effect of liquid viscosity on mixing time at BC = 0.009 m and draft tube diameter of 0.29 m for cross shaped sparger configuration.

5.2 Liquid Circulation Velocity

5.2.1 Effect of Gas Flow Rate and Bottom Clearances

The liquid circulation velocity is the primary hydrodynamic parameter differentiating the draft tube airlift bioreactors from bubble columns due to the continuous recirculation of liquid. The liquid circulation velocity in the airlift bioreactors depends on the geometry of the bioreactor. Liquid circulation velocity versus superficial gas velocity data for three bottom clearances (BC= 0.003, 0.006, 0.009 m) in draft tube diameter of 0.22 m using cross shaped sparger configuration is shown in Fig.5.2.1.



Fig.5.2.1 Effect of bottom clearance on liquid velocity for all three clearances in draft tube diameter of 0.22 m (air-water system) using cross shaped sparger configuration.

The results in Fig.5.2.1 show liquid circulation velocity increases with increasing gas velocity. This can be explained based on that an increase in gas flow increases the difference in the gas holdup which is the driving force for liquid circulation velocity. Trends of the liquid velocity have similar shapes but different magnitudes (Fig.5.2.1). Also, increasing the BC increases the liquid velocity (Fig 5.2.1), as a result of the fact that by varying BC, the pressure drop at the bioreactor bottom is strongly affected. When the free area for liquid flow between the bottom plate and the draft-tube is very constrained, changes in the bottom clearance have strong influence on pressure drop.

5.2.2 Effect of Draft Tube Diameter

Fig.5.2.2 shows the comparison between the different liquid circulation velocities at different superficial gas velocities for three different draft tube diameters (0.22, 0.29, 0.34 m) fixed at a bottom clearance of 0.009 m using cross shaped sparger configuration. Fig.5.2.2 clearly demonstrates that liquid circulation velocity is increased for draft tubes of higher diameters. An increase in draft tube diameter results in a lower ratio of A_d/A_r (i.e., higher cross sectional ratio in the riser with less circulation velocity). With a ratio of less than 1 of A_d/A_r , (for draft tube diameter of 0.29 m), the downcomer cross sectional area is less than that for the riser. Thus, the liquid velocity in the riser is less due to its wider cross sectional area; therefore there is higher gas residence time in the riser.



Fig.5.2.2 Effect of draft tube diameter on liquid circulation velocity for BC = 0.009 m (air-water system) using cross shaped sparger configuration.

The downcomer liquid velocity is increased due to the decrease in cross-sectional area of the downcomer. Tobajas et al. (1999) employed a rectangular split airlift and found that the liquid circulation in the downcomer and mass transfer coefficient increased with decreasing cross sectional ratios.

5.2.3 Effect of Sparger

In Fig.5.2.3 the influence of sparger configuration on liquid circulation velocity versus superficial gas velocity in a draft tube diameter of 0.29 m, at a bottom clearance of 0.009 m is shown. An increase in liquid circulation velocity is observed between the data corresponding to the cross shaped sparger configuration. This is because the cross shaped sparger produced a uniform distribution of the gas in the riser than its counterpart the circular shaped sparger. This is attributed to a higher disengagement (due to lower rise velocity of gas bubbles) of gas into the downcomer due to even distribution of gas creating a higher driving force for fluid to circulate in the bioreactor. By visual inspection, the circular shape sparger creates a poor distribution of the gas in the riser with a tendency to coalescing bubbles, resulting in a reduced driving force for liquid recirculation in the bioreactor.



Fig.5.2.3 Effect of sparger configuration on liquid circulation velocity, BC = 0.009 m in draft tube diameter of 0.29 m (air-water system).

5.2.4 Effect of Fluid Viscosity

In Fig.5.2.4 the influence of different fluids at varying viscosities (water, 34.5% sugar solution, 0.2% and 0.5% xanthan gum solutions) on liquid circulation velocity versus superficial gas velocity in a draft tube diameter of 0.29 m, fixed at a bottom clearance of 0.009 m using cross shaped sparger configuration is compared. Fig.5.2.4 shows that the liquid circulation velocity decreases with the increase in fluid viscosity. This phenomenon could be attributed to increased fluid viscous drag which hinders liquid circulation of xanthan gum (Fields *et al.* 1984). The flow behavior of xanthan gum is different from that of water and sugar solutions. The friction factor is substantially lower and this drag reduction (friction) usually increases with the flow rate and xanthan gum molecular weight (Fields *et al.* 1984). An increase in viscosity diminishes the

bubble rise velocity, hence hinders the escape of bubbles in the downcomer (Onken and Weiland, 1981). This leads to a high gas hold-up in the downcomer resulting in a reduced difference in mean density between the riser and downcomer. With the above factors influencing the driving force for liquid circulation, less energy is left to accelerate the fluid.



Fig.5.2.4. Effect of liquid viscosity on liquid circulation velocity at BC = 0.009 m in draft tube diameter of 0.29 m (air-water system) for cross shaped sparger configuration.

5.3 Gas Holdup

5.3.1 Effect of Gas Flow Rate and Bottom Clearance

Fig.5.3.1 represents the effects of gas flow and bottom clearances in draft tube diameter of 0.29 m. It is evident from this Figure that gas holdup values significantly increased with increasing gas flow rate. Results obtained are in agreement with previously reported results in the literatures (Chisti, 1998; Gavrilescu and Tudose, 1999; Gouveia *et al.*, 2003). As the gas flow rate is increased more bubbles are produced which results in an increase in gas holdup. At lower gas input (i.e. 0.00165-0.00331 m/s); homogenous gas bubbles existed resulting in a lower gas holdup accounting for a bubbly flow regime. Moreover, as the gas flow rate increased, this homogenous dispersion turns into a heterogeneous regime with the formation of larger gas bubbles. For a typical aerobic fermentation, operations in the heterogonous regime is not desirable as the gas-liquid interfacial area for mass transfer rate is reduced and the gas residence time shortened due to the increasing velocity of larger bubbles.

Gas holdup in the system is largely controlled by the liquid circulation velocity as it carries the bubbles down into the downcomer. The recorded gas holdups indicate that the gas holdup increases as the bottom clearance decreases (Fig 5.3.1) as a result of the decreasing liquid velocity (Fig.5.2.1). This may be due to the fact that a decrease in the liquid velocity results in more gas being retained in the downcomer as opposed to being disengaged at the top of the bioreactor.



Fig.5.3.1 Effect of gas flow rate and bottom clearances on gas holdup in a draft tube diameter of 0.22 m and BC = 0.009 m (air-water system) using cross shaped sparger.

5.3.2 Effect of Draft Tube Diameter

A comparison of gas holdup values for three draft tube diameters fixed at a bottom clearance of 0.009 m is presented in Fig.5.3.2. The results in Fig.5.3.2 demonstrate increases in gas holdup with decreasing A_d/A_r ratio. With the increase in the draft tube diameter, (i.e. decreasing A_d/A_r ratio), there is a decrease in flow resistance which enhances more gas entrainment to the downcomer. The larger the draft tube diameter, the less resistance and more flow will entrain into the downcomer. The decrease in flow resistance with decreasing values of A_d/A_r ratio improves the entrainment of gas bubbles into the downcomer and thus the values of gas holdup in the downcomer increases. These results show the influence of A_d/A_r ratio as a principal

geometry which influences the friction in the airlift bioreactor (Chisti, 1989; Gavrilescu and Tudose, 1999; Gouveia *et al.*, 2003).



Fig.5.3.2 Effect of draft tube diameter on gas holdup for BC = 0.009 m (air-water system) using cross shaped sparger configuration.

5.3.3 Effect of Sparger

In Fig.5.3.3 the combined effects of gas flow and sparger configurations in a draft tube diameter of 0.29 m fixed at a bottom clearance of 0.009 m are shown. As expected, an increase in gas flow confirms the production of more bubbles for gas holdup values. The results also show increase in gas holdup values for the cross shaped sparger configuration. The differences in the gas holdup values for the two sparger configurations can be explained by visual observation, where the cross shaped sparger configuration of the gas in the riser, as

compared to the circular shaped sparger. For an increase in gas holdup, the liquid circulation velocity gets lower. The lower liquid circulation velocity in the downcomer with the circular shaped sparger (Fig.5.2.3) should result in an increase of bubble entrainment into the downcomer. However, this was not the case; which can be explained by the coalescence of bubbles due to uneven distribution by the circular shaped sparger. Moreover, the cross shaped sparger with higher liquid circulation velocity produced evenly distributed bubbles enhancing the entrainment which resulted in higher gas hold up values.



Fig.5.3.3 Effect of sparger configuration on gas holdup at BC = 0.009 m in a draft tube diameter of 0.29 m (air-water system).

5.3.4 Effect of Liquid Viscosity

The effects of fluid viscosity on gas hold up in draft tube diameter of 0.29 m are shown in Fig.5.2.4. It can be seen that viscosity has a significant influence on gas holdup results. The highest gas holdup values were recorded in water and sugar solutions because of their low viscosity. For such Newtonian fluid with low viscosity, numerous small bubbles with low ascending velocity were produced, leading to an increased residence time of gas in the bioreactor. Consequently, the gas hold-up was higher at lower viscosities.



Fig.5.3.4. Effect of liquid viscosity on gas holdup at BC = 0.009 m in draft tube diameter of 0.29 m (air-water system) using a cross shaped sparger configuration.

Conversely, in the viscous non Newtonian coalescing fluids (i.e., 0.2% and 0.5% xanthan gum solutions), intensive bubble coalescence increases the bubble rise velocity leading to a shorter gas residence time, and consequently, gas hold-up decreased with increasing viscosity.

The bubble coalescence results from inefficient gas radial dispersion as reported by Poggeman *et al.* (1983). At any superficial gas velocity in viscous fluids, gas from the sparger does not disperse uniformly radially. Therefore the bubbles tend to accumulate in the space above the sparger resulting in high bubble concentration which increases the bubble coalescence (Poggeman *et al.*, 1983).

In a real fermentation using xanthan gum (Olivier and Oosterhuis, 1988), the bubble coalescence was strongly enhanced by increasing xanthan concentrations similar to the results in the present work. From Fig.5.3.4, it is clear that in the range of fluid viscosities examined, bubble coalescence is enhanced at higher viscosity, probably by poor radial gas dispersion and a decrease in liquid circulation velocity (Fig.5.2.4). These observations and results are in agreement with results reported by Wen *et al.* (2005), Onken and Weiland (1980) and Fields *et al.* (1984).

5.4 Shear Rate

In most chemical processes, the shear rate is of no significance by itself except as a means of increasing the heat and mass transfer properties (Cerri *et al.*, 2007). However, for any fermentation processes, the shear rate is important. Higher shear stress may damage fragile cells leading to the loss of productivity. Shear rate has had important implications in cultures involving filamentous microorganisms, animal and plant cells (Chisti and Moo Young, 1986).

5.4.1 Effect of Gas Flow Rate and Bottom Clearance

The combined effects of bottom clearances and gas velocity on shear rates in draft tube diameter of 0.22 m at bottom clearances of 0.006, 0.009 m are shown in Fig.5.4.1 using the cross shaped sparger. Generally, the shear rates increased as the gas velocity is increased as shown in Fig.5.4.1. This is mainly because more energy is dissipated which can be turbulent to reticulate the fluid in the bioreactor.



Fig.5.4.1 Effect of gas flow on shear rate for draft tube diameter of 0.22 m at two bottom clearances (air-water system) using cross shaped sparger configuration.

Also results in Fig.5.4.1 indicate that an inverse relationship between the bottom clearances and shear rate can obviously be observed. According to this Figure, a decrease in bottom clearance corresponds to an increase in shear rate values. This is due to impedance to flow restriction

caused by the shorter distance between the bioreactor bottom and draft tube. At lower superficial gas velocities, the effect of bottom clearance was insignificant on the shear rate. The above observations are in agreement with those reported by Merchuk and Berzin (1995) and Luo and Dahhan (2008).

5.4.2 Effect of Draft Tube Diameter

Fig.5.4.2 displays the effects of superficial gas velocity in the riser (U_{gr}) on shear rate in different draft tube diameters (0.22, 0.29, 0.34 m) fixed at a bottom clearance of 0.009 m using cross shaped sparger. Results in Fig.5.4.2 indicate that the shear rate is higher for a lower A_d/A_r ratio. For a decrease in cross sectional area (i.e. and increase in draft tube diameter) it is known that liquid circulation velocity is higher (Gavrilescu and Tudose, 1999; Gouveia *et al.*, 2003) thus increasing the shear rate. However, in draft tube diameters of 0.22 m and 0.29 m, shear rate values are almost the same. This could be because the increase in liquid circulation velocity in these bioreactor setups did not vary as much as that of the 0.34 m draft tube diameter.



Fig.5.4.2 Effect of draft tube diameter on shear rate at BC = 0.009 m (air-water system) using cross shaped sparger configuration.

5.4.3 Effect of Sparger

Fig.5.4.3 compares the influence of sparger configuration on the profile of shear rates versus superficial gas velocity in a draft tube diameter of 0.29 m, at a bottom clearance of 0.009 m. The result shows an increase in shear rate for both configurations as gas flow is increased. Moreover, the circular shaped sparger resulted in the higher shear rate attributed to an increase in bubble coalesce produced in the circular shaped sparger. Results show 20% reduction in shear rate values using the cross shaped sparger configuration. The results suggest that for a sparger configuration which encourages bubble coalescence, the interfacial area for mass transfer is reduced. The shear in airlift bioreactors occurs mainly in the vicinity of the gas liquid interface.

Therefore, the shear rate is negatively affected by the bubble coalescence as gas velocity increases. Therefore, the shear rate is negatively affected by the bubble coalescence as gas velocity increases. Contreras *et al.*, 1999 reported similar results for shear rate values using different spargers.



Fig.5.4.3 Effect of sparger configuration on shear rate at BC = 0.009 m in draft tube diameter of 0.29 m (air-water system)

5.4.4 Effect of Liquid Viscosity

Fig.5.4.4 demonstrates the effect of varying fluid viscosities (water, 34.5% sugar solution, and 0.2% and 0.5% xanthan gum solutions) on shear rate versus superficial gas velocity in a draft tube diameter of 0.29 m. According to Fig.5.4.4, the shear rate decreased with increasing fluid viscosity. By increasing the fluid's viscosity the liquid circulation velocity can be lowered (Fig.5.2.4). It is also clear that the energy dissipation is decreases with increasing fluid viscosity,

causing to decrease the shear rate. Trends or increase in shear rate values in the downcomer is usually minimal. This observation is in agreement with results obtained by Grima et al., 1997. Results obtained from xanthan gum solutions (shear rate 50-500 s⁻¹) are well within the range for animal cell culture cultivation (Grima et al., 1997).



Fig.5.4.4. Effect of liquid viscosity on shear rate at BC = 0.009 m in draft tube diameter of 0.29 m using cross shaped sparger configuration.
CHAPTER 6

CONCLUSIONS AND RECOMMENDATIONS

6.1 Conclusions

ERT provides a powerful means for non-invasive measurements of fluid mixing parameters in the airlift bioreactor. The investigations performed for all parameters reveal that increasing the superficial gas velocity, (U_{gr}) corresponds to an increase in energy generated. This results in an increase in gas holdup which creates a higher bulk density difference (higher driving force) for liquid circulation velocity. This improves the gas-liquid separation at the top of the bioreactor and therefore, enhances turbulence which further enhances mixing.

The experimental results of mixing parameters such as: the mixing time, liquid circulation velocity, gas holdup, and shear rate revealed that geometric parameters are important factors that affect these mixing characteristics. Experimental results for bottom clearances and ratios of cross sectional areas of downcomer and riser (A_d/A_r) revealed that at these geometries, and depending on their flow resistance, the loss of kinetic energy and friction impacts the mixing characteristics. For a shorter bottom clearance, the greater impedance to the fluids movement results in higher friction losses.

The effect of sparger configuration is significant for mixing time and hydrodynamic parameters (gas holdup and liquid circulation velocity). Apparently, this is attributed to an even bubble distribution produced by the cross shaped sparger, resulting in an increase in driving force for liquid circulation velocity and hence shortened and enhanced mixing times. The even

distribution of bubbles prevents bubble coalescence which reduces mass transfer rates. However, the effect of the sparger on shear rate is not as significant.

The effects of fluid viscosity on mixing time, liquid circulation velocity, gas holdup, and shear rate were significant. Increased viscosity caused an increase in the circulation path flow resistance and increased the fluid viscous drag, thus reducing liquid circulation velocity and hence increasing mixing time. For both sugar and xanthan gum solutions, larger numbers of gas bubbles were formed with low bubble rise velocities resulting in a decrease for driving force for liquid circulation velocity. Shear rate values corresponding to typical shear rate range up to 6000s⁻¹ for viscous fluids were recorded, although much higher values were noted for less viscous media (water).

6.2 Recommendations

It is worthy to point out that all measurements were made in the downcomer (annulus of the bioreactor), for future studies to enhance our knowledge of the bioreactor system using this non invasive ERT technique, experiments could be performed in the split cylinder airlift bioreactor. This type of bioreactor affords ERT measurements both in the riser and downcomer regions. Although it is known that various mixing parameters and shear rates are of magnitudes higher in the riser than in the downcomer, ERT measurements of both regions would support these findings and predictions.

To validate the ERT system, further studies to include the churn and heterogeneous regimes could be performed and results could then be correlated for the various liquid circulation velocities. Computational fluid dynamics (CFD) modeling can be recommended. CFD analysis can be used to simulate and validate the mixing processes.

NOMENCLATURE

 A_b cross sectional area for flow under baffle or draft tube (m²)

- A_d cross sectional area of downcomer (m²)
- A_r cross sectional area of riser (m²)
- A_d/A_r ratio of cross sectional area of downcomer and riser
- B_o Bodenstein number
- c initial tracer concentration (kg/m³)
- c_{∞} mean or final concentration of tracer (kg/m³)
- D_c column diameter (m)
- D_d draft diameter (m)
- D_s gas separator diameter (m)
- E_z axial dispersion coefficient
- F_r froude number (-)
- g gravitational acceleration (m/s^2)
- *H* height of bioreactor (m)
- h_L height of gas free liquid (m)
- h_d height of draft tube (m)

h_D	height of gas liquid dispersion (m)
h_t	draft tube height (m)
K	consistency index (Pa s ⁿ)
K _B	frictional co-efficient at the bottom
K _T	frictional co-efficient at the top
K _L a	volumetric mass transfer coefficient (m ⁻¹)
L _c	circulation length (m)
Mo	Morton number
n	flow behavior index
Pe	peclet number
S	standard deviation
<i>t</i> _c	circulation time (s)
<i>t</i> _m	mixing time (s)
TC	top clearance (m)
U_g	superficial gas velocity (m/s)
Ugr	superficial gas velocity in riser (m/s)
U_l	circulation liquid velocity (m/s)

- U_{sb} bubble swarm velocity (m/s)
- U_{sp} particle swarm velocity (m/s)
- z axial dispersion distance in reactor (m)

Greek Symbols

- ε_g gas holdup
- ε_{gr} gas holdup in riser
- ε_{gd} gas holdup in downcomer
- ε_s solid loadings (-)
- μ_L viscosity of liquid (Pa.s)
- ρ_s density of solids (kg/m³)
- ρ_L density of liquids (kg/m³)
- au_m dimensionless time
- τ shear stress (N/m²)
- $\dot{\gamma}$ shear rate (s⁻¹)
- σ surface tension (N/m)

ABBREVIATION

- ADC analogue to digital converter
- ADM axial dispersion model
- ALR airlift reactor
- API active pharmaceutical ingredient
- BAS biofilm activated sludge
- BC bottom clearance (m)
- BDS biodesulphurization
- BHb bovine hemoglobin
- BSA bovine serum albumin
- CAT computer axial tomogoraphy
- CARPT computer automated radioactive particle tracking
- CFD Computational fluid dynamics
- CMC carboxyl methyl cellulose
- COD chemical oxygen demand
- DAC digital to analogue converter
- DAS data acquisition system

DTAB draft tube airlift bioreactor

EIFBAB external loop inverse fluidized bed airlift bioreactor.

- ECT electrical capacitance tomography
- EIT electrical impedance tomography
- EMT electromagnetic tomography
- ERT electrical resistance tomography
- GAC granular activated carbon
- GFM grounded floating measurements
- GO glucose oxidase
- HFP helical flow promoter
- SNR signal to noise ratio
- VCCS voltage current converter source
- *XG* xanthan gum

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APPENDICES

Appendix A

ERT DATA FOR MIXING TIMES

Table 1a. Mixing time for draft tube diameter 0.22 m bottom clearance 0.009 m (air-water) using cross shaped sparger configuration.

$U_{gr} \times 10^3 (m/s)$	1.65	3.331	4.75	6.69	8.07
$t_{m1(s)}$	44	40	37	32	29
$t_{m2(s)}$	43	40	37	31	30
$t_{m3(s)}$	45	41	39	33	31
mean	44	40.33	37.67	32	30
standard	1	0.58	1.15	1	1
deviation					

Table 1b. Mixing time for draft tube id 0.22 m bottom clearance 0.006 m (air-water) using cross shaped sparger configuration.

$U_g \times_r X10^3 (m/s)$	1.65	3.331	4.75	6.69	8.07
<i>t_{m1(s)}</i>	49	46	40	35	34
$t_{m2(s)}$	48	44	41	36	34
<i>t_{m3(s)}</i>	47	45	40	36	35
mean	48	45	40.33	35.67	34.33
standard deviation	1	1	0.58	0.58	0.58

Table 1c. Mixing time for draft tube diameter 0.22 m bottom clearance 0.003 m (air-water) using cross shaped sparger configuration.

$U_{gr} \times 10^3 (m/s)$	1.65	3.331	4.75	6.69	8.07
$t_{m1(s)}$	53	48	43	42	39
$t_{m2(s)}$	51	47	44	41	40
$t_{m3(s)}$	52	46	44	40	38
mean	52	47	43.67	41	39
standard	1	1	0.58	1	1
deviation					
deviation					

$U_{gr} \times 10^3 (m/s)$	1.65	3.331	4.75	6.69	8.07
	26		20	25	2.6
$t_{m1(s)}$	36	32	29	27	26
$t_{m2(s)}$	34	34	30	26	24
	25	24	20	27	26
$t_{m3(S)}$	35	34	29	27	26
mean	35	33.33	29.33	26.67	25.33
standard	1	1.15	0.58	0.58	1.15
deviation					

Table 2a. Mixing time for draft tube diameter 0.29 m bottom clearance 0.009 m (air-water) using cross shaped sparger configuration.

Table 2b. Mixing time for draft tube diameter 0.34 m bottom clearance 0.009 m (air-water) using cross shaped sparger configuration.

$U_{gr} \times 10^3 (m/s)$	1.65	3.331	4.75	6.69	8.07
$t_{m1(s)}$	31	26	22	15	14
$t_{m2(s)}$	31	28	22	15	14
<i>t_{m3(S)}</i>	30	28	23	16	13
mean	30.67	27.33	22.33333	15.33	13.67
standard	0.58	1.15	0.58	0.58	0.58
deviation					

Table 2c. Mixing time for draft tube diameter 0.29 m bottom clearance 0.009 m (air-water) using circular shaped sparger configuration.

$U_{gr} \times 10^3 (m/s)$	1.65	3.31	4.75	6.69	8.07
$t_{m1(s)}$	50	44	39	30	25
$t_{m2(s)}$	52	45	39	29	23
$t_{m3(s)}$	50	44	37	30	24
mean	50.67	44.33	38.33	29.67	24
standard devaiton	1.15	0.58	1.15	0.58	1

Table 3. Mixing time for draft tube diameter 0.29 m bottom clearance 0.009 m (sugar solution) using cross shaped sparger configuration.

$U_{gr} \times 10^3 (m/s)$	1.65	3.331	4.75	6.69	8.07
$t_{m1(s)}$	55	49	46	35	35
$t_{m2(s)}$	55	51	46	37	33
$t_{m3(s)}$	57	50	45	36	32
mean	55.67	50	45.67	36	33.33
standard	1.15	1	0.58	1	1.53
deviation					

Table4a. Mixing time for draft tube diameter 0.29 m bottom clearance 0.009 m (0.2% xanthan gum solution) using cross shaped sparger configuration.

$U_{gr} \times 10^3 (m/s)$	1.65	3.331	4.75	6.69	8.07
$t_{m(s)}$	699	514	430	289	201

Table4b. Mixing time for draft tube diameter 0.29 m bottom clearance 0.009 m (0.5% xanthan gum solution) using cross shaped sparger configuration.

$U_{gr} \times 10^3 (m/s)$	1.65	3.331	4.75	6.69	8.07
$t_{m(s)}$	2397	1905	1276	924	829

Appendix B

ERT DATA FOR CIRCULATION TIME

Table 1a. Circulation time for draft tube diameter 0.22 m bottom clearance 0.009 m (air-water) using cross shaped sparger configuration.

$U_{gr} \times 10^3 (m/s)$	1.65	3.331	4.75	6.69	8.07
$t_{c1(s)}$					
	10	9	8	7	7
$t_{c2(s)}$					
	10	10	9	8	7
$t_{c3(s)}$					
	11	9	8	7	6
mean					
	10.33	9.33	8.33	7.33	6.67
Standard					
deviation	0.58	0.58	0.58	0.58	0.58

Table 1b. Circulation time for draft tube diameter 0.22 m bottom clearance 0.006 m (air-water) using cross shaped sparger configuration.

$U_{gr} \times 10^3 (m/s)$	1.65	3.331	4.75	6.69	8.07
$t_{c1(s)}$	11	9	9	9	7
$t_{c2(s)}$	11	9	10	7	6
$t_{c3(s)}$	10	10	8	8	8
mean	10.67	9.33	9	8	7
standard					
deviation	0.58	0.58	1	1	1

Table 1c. Circulation time for draft tube diameter 0.22 m bottom clearance 0.003 m (air-water) using cross shaped sparger configuration.

$U_{gr} \times 10^3 (m/s)$	1.65	3.331	4.75	6.69	8.07
$t_{c1(s)}$	13	11	10	9	9
$t_{c2(s)}$	12	11	10	9	9
$t_{c3(s)}$	9	9	9	9	8
mean	11.33	10.33	9.67	9	8.67
standard					
deviation	2.08	1.15	0.58	0	0.58
$U_{gr} \times 10^3 (m/s)$	1.65	3.331	4.75	6.69	8.07
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$t_{c1(s)}$	8	8	6	5	4
$t_{c2(s)}$	8	7	7	5	4
$t_{c3(s)}$	9	7	6	6	5
mean	8.33	7.33	6.33	5.33	4.33
standard					
deviation	0.58	0.58	0.58	0.58	0.58

Table 2a. Circulation time for draft tube diameter 0.29 m bottom clearance 0.009 m (air-water) using cross shaped sparger configuration.

Table 2b. Circulation time for draft tube diameter 0.34 m bottom clearance 0.009 m (air-water) using cross shaped sparger configuration.

$U_{gr} \times 10^3 (m/s)$	1.65	3.331	4.75	6.69	8.07
$t_{c1(s)}$	7	6	5	4	3
$t_{c2(s)}$	7	6	5	4	3
$t_{c3(s)}$	7	6	5	4	4
mean	7	6	5	4	3.33
standard					
deviation	0	0	0	0	0.58

Table 2c. Circulation time for draft tube diameter 0.29 m bottom clearance 0.009 m (air-water) using circular shaped sparger configuration.

$U_{gr} \times 10^3 (m/s)$	1.65	3.31	4.75	6.69	8.07
$t_{c1(s)}$	10	8	7	6	5
$t_{c2(s)}$	11	9	7	7	6
$t_{c3(s)}$	11	9	8	7	5
mean	10.66667	8.666667	7.333333	6.666667	5.33
standard					
deviation	0.58	0.58	0.58	0.58	0.57735

Table 3. Circulation time for draft tube diameter 0.29 m bottom clearance 0.009 m (sugar solution) using cross shaped sparger configuration.

$U_{gr} \times 10^3 (m/s)$	1.65	3.331	4.75	6.69	8.07
$t_{c1(s)}$	11	10	10	7	7
$t_{c2(s)}$	12	9	8	8	7
$t_{c3(s)}$	12	10	8	8	6
mean	11.67	9.67	8.67	7.67	6.67
standard					
deviation	0.58	0.58	1.15	0.58	0.58

Table 4a. Circulation time for draft tube diameter 0.29 m bottom clearance 0.009 m (0.2% xanthan gum solution) using cross shaped sparger configuration.

$U_{gr} \times 10^3 (m/s)$	1.65	3.331	4.75	6.69	8.07
$t_{c(s)}$	31	26	22	18	16

Table 4b. Circulation time for draft tube id 0.29 m bottom clearance 0.009 m (0.5% xanthan gum solution) using cross shaped sparger configuration.

$U_{gr} \times 10^3 (m/s)$	1.65	3.331	4.75	6.69	8.07
$t_{c(s)}$	44	36	28	23	17