HYDRODYNAMIC ASPECTS OF AIRLIFT CONTACTORS

by

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A thesis submitted to the University of London
in fulfillment of the thesis requirement
for the degree of
Doctor of Philosophy
in
Chemical Engineering

July 1996

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To my parents
ABSTRACT

This study deals with gas void fraction, liquid circulation velocity and liquid mixing in airlift contactors.

Experimental measurements of gas holdup, liquid velocity, mixing and circulation times, Bodenstein number and liquid phase axial dispersion coefficient were obtained with various Newtonian and non-Newtonian fluids using different reactor configurations. The experimental data were obtained in two distinct laboratory scale draft tube internal loop airlift devices (52 L) and an external loop airlift contactor (35 L).

The effect of superficial gas velocity (low to intermediate range), liquid properties (salt concentration and liquid viscosity), two immiscible liquid phases (oil/water) and reactor geometry on the gas holdup, liquid velocity and liquid mixing in airlift vessels were examined.

Gas holdup was observed to increase with increasing gas velocity, liquid phase electrolyte concentration and Newtonian and pseudoplastic non-Newtonian liquid viscosity. The effect of the two immiscible liquid phases on gas holdup was relatively complex. Initially, on addition of oil to the continuous phase (water) there was a sharp increase in gas holdup. As the concentration of oil-in-water emulsion increased the gas holdup was observed to decrease and then increase as the pure oil limit was reached.

The liquid velocity and liquid phase mixing in the airlift vessel improved with increasing gas flow rate. But, with an increase in electrolyte concentration, liquid viscosity and oil-in-water emulsion concentration the superficial liquid velocity decreased. Hence, the liquid mixing was found to deteriorate with increasing salt concentration, liquid viscosity and oil-in-water emulsion concentration. For a fixed gas velocity, the local Bodenstein numbers and axial dispersion coefficients indicated that the mixing in the airlift contactor was not uniform in the different regions of the vessel. The downcomer section gave considerably higher axial dispersion coefficient values compared to the riser section for a given gas flow rate. The riser and downcomer Bo numbers indicated that the liquid flow in the riser section approached plug flow, while the flow in the downcomer section was reasonably well mixed. The liquid mixing in the reactor as a whole was somewhere in-between the mixing in the riser and downcomer sections.
A hydrodynamic model was employed to estimate the riser gas holdup and liquid velocity in the airlift vessels. This model was developed using a drift flux model together with an energy balance over an airlift loop. Model predictions were compared with experimental data obtained in this study. In addition, suitable experimental data available in previous literature were used. The applicability of the model to a broad range of airlift reactor types (split-cylinder and draft tube internal loops; external loop) and scales ($0.033 \text{ m}^3 \leq V \leq 1.058 \text{ m}^3$; $1.220 \text{ m} \leq H_t \leq 5.076 \text{ m}$; $0.11 \leq A_d/A_r \leq 1.45$) and with various Newtonian and non-Newtonian liquids is demonstrated.

For non-Newtonian media in airlift vessels, a correlation for average shear rate was also developed. This relationship considers the effect of gas holdup, liquid velocity and flow behaviour index on shear rate. A higher apparent liquid viscosity gave a lower average shear rate in the vessel than a lower apparent viscosity, at a given superficial gas velocity.
ACKNOWLEDGEMENTS

I am sincerely grateful to my supervisors, Dr. Parviz Ayazi Shamlou and Dr. Andrew Ison, for their suggestions, constructive criticisms, encouragement and support during the course of this study.

The successful construction and operation of the laboratory scale airlift reactors would not have been possible without the help of the technical staff (electrical, laboratory, workshop and stores). I wish to express my thanks to Martin Town, Martyn Vale, Patrick Bloomfield, Samuel Okagbue and John Betts for providing outstanding service.

I am also thankful to all my colleagues in the Chemical Engineering Research group, both present and past, for the many useful discussions and suggestions. I am especially indebted to Bita Vafaei, Olivier Djezvedjian and Nadja König for their invaluable help.

The financial support of the Engineering and Physical Sciences Research Council of UK under Grant No. 93310766 is gratefully acknowledged.
The essential qualifications of a seeker of wisdom are a capacity to discriminate between the eternal and the non-eternal, detachment from the enjoyment of the fruits of action, terrestrial and celestial, self-control and an ardent desire for spiritual freedom.

Pythagoras (582 BC - 500 BC)

Mankind is engaged in an eternal quest for that “something else” he hopes will bring him happiness, complete and unending. For those individual souls who have sought and found God, the search is over: He is that Something Else.

Paramahansa Yogananda (1893 - 1952 AD)
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1 INTRODUCTION

1.1 Overview

In the past 40 years, since the patent for a concentric tube airlift reactor (Lefrancois et al., 1955), this family of gas-liquid contactors has received increasing attention. Initially, the device was mainly applied for large-scale microbiological processes, such as single cell protein production (SCP) from the carbon sources methanol and \( n \)-paraffins (Blakebrough et al., 1967; Onken and Weiland, 1983). While interest in SCP production has declined, the opposite is true of the airlift reactor. Today, the airlift is viewed as a potential reactor for a wide variety of biological and chemical processes, ranging from waste treatment (Moo-Young and Chisti, 1994) to fermentations (Arathoon and Birch, 1986).

Airlift reactors are gas-liquid or gas-slurry contacting devices, characterised by fluid circulation in a clear and defined cyclic pattern through conduits specifically designed for this purpose. This liquid flow pattern creates complex and unique hydrodynamic conditions, which for engineering purposes can be characterised through the parameters of gas holdup, liquid circulation velocity and mixing. Due to the interaction of these parameters, defining the behaviour of these reactors is complicated (Siegel et al., 1988).

In an airlift reactor, the contents of the vessel are pneumatically agitated by a stream of gas (usually air). In addition to agitation, this stream also has the important function of mediating exchange between the gas phase and the medium. An airlift can be considered as an extension of a bubble column since this is also pneumatically agitated. The main difference lies in the fluid flow which depends on the geometry of the system. The bubble column is a simple vessel where gas is injected, usually at the bottom, and mixing of the medium is caused by the ascending bubbles. Liquid circulation on a macroscale does not occur in the column. On the other hand, in the airlift contactor a well defined fluid circulation occurs and is determined by the reactor design (Merchuk, 1990).

Although airlift reactors are being increasingly used in commercial applications, there are still many unanswered questions about the design and hydrodynamic behaviour of these reactors (Chisti and Moo-Young, 1987; Siegel et al., 1988). Partly, this is due to the wide variations of reactor geometries, physio-chemical systems and experimental techniques employed by various researchers to study the airlift reactor. This has led to a wide variation of the data available in the open literature and to contradictions between researchers. For this reason, generally accepted design equations for the scaleup of the
airlift contactor is still not available and much work remains to be done to facilitate a better understanding of airlifts and exploit their potential for industrial uses.

1.2 Classification

Two basic classes of airlifts can be distinguished, from a morphological point of view:

1) the internal loop (baffled) airlift contactors where what would otherwise be a simple bubble column has been split into a riser and a downcomer by an internal baffle (Figure 1.1), and

2) the external (outer) loop airlift reactors where the riser and the downcomer are two quite separate tubes connected by horizontal sections near the top and the bottom (Fig. 1.1).

Internal and external loop reactors could be further subdivided. Internal loop airlifts, for example, may be of the split-cylinder type or they may have a concentric draft-tube configuration. In the latter either the draft-tube or the annulus may be gas sparged. The draft-tube and the vertical baffle (in split-cylinder mode) may themselves be divided into vertical sections to increase the link between the riser and the downcomer (Chisti and Moo-Young, 1987). Chisti and Moo-Young (1993) have shown several gas-liquid separator designs for draft-tube and split-cylinder airlift reactors. Internal loop airlift reactors can also be operated without gas-liquid separators (Chisti et al., 1995). Multiple concentric draft tubes (Margaritis and Sheppard, 1981) and multiple airlift loop reactors (Bakker et al., 1993) have also been studied. There are lesser variety in external loop reactors, but several designs of horizontal connections between the riser and the downcomer, particularly the head-space region, may be appropriate (Chisti and Moo-Young, 1993). In summary, countless possible variations on basic airlift design exist and may be advantageously employed for different applications.

Internal and external loop airlift reactors usually have circular cross-sections, but rectangular and square cross-sections have also been studied (Gasner, 1974; Siegel et al., 1986; Siegel and Merchuk, 1988; 1991).

Regardless of the configuration (external or internal loop), all airlift reactors consist of four distinct sections with different flow characteristics. Therefore, each section may have different characteristics from the point of view of hydrodynamics, mass and heat transfer, as well as yield. Hence, when designing or interpreting data on airlift contactors, it is important to recognize that the system is not a simple unit. The four connected sections are:
Introduction

(1) **Riser**: The riser is the section where both gas and liquid flow upwards, the gas being injected at the bottom of this section. In terms of reactor hydrodynamics, most researchers consider this section to be one of the most important and majority of existing hydrodynamic models predict either the riser gas holdup or the riser liquid velocity;

(2) **Downcomer**: this section is parallel to the riser and is connected to it at the bottom and at the top. The downcomer is the section where the liquid flows down from the top to the bottom of the contactor. It may be desirable to prevent or increase bubble recirculation depending on the specific process carried out. For example, in aerobic fermentations, to avoid recirculation of small bubbles with a low concentration of oxygen and a relatively high concentration of carbon dioxide, it is desirable to prevent bubble recirculation. But bubble recirculation is increased in systems that are not tall enough to assure satisfactory absorption of the gas component during a single pass;

(3) **Gas-liquid separator**: this section connects the riser to the downcomer, at the top of the reactor, allowing liquid recirculation and gas disengagement. The gas-liquid separator can be designed to change the fluid residence time in this region, giving either nearly complete gas disengagement or high gas entrainment and recirculation;

(4) **Base**: usually, in most airlift designs, the bottom connection zone between the riser and the downcomer is very simple. The base is usually not considered as significantly affecting overall reactor behaviour (Merchuk, 1990). Therefore most researchers have ignored this section altogether (Bello et al., 1984; Chisti and Moo-Young, 1993; Joshi et al., 1990; Merchuk, 1986; Verlaan et al., 1989).

Since all the four elements are interconnected, the design of each section (apart from the base) will affect the performance characteristics of each of the other sections.

The airlift reactor design determines the flow patterns in the reactor. The flow patterns will have a direct influence on the gas holdup and an indirect influence on the liquid recirculation velocity and mass transfer (Siegel et al., 1988). Hence, the yield of a process can be significantly changed by the design of the airlift reactor employed. The airlift reactor can be designed to maximise different properties (such as gas-liquid mass transfer rate) depending on the specific process objectives. For example, in animal cell cultivation the primary objective in reactor design would be the control of shear stress acting on the cell suspensions; but in the case of wastewater treatment the reactor would be designed to maximise gas utilization, while minimising energy input.
1.3 The Advantages of Airlift Reactors

The main advantages in using airlift vessels compared to other conventional bioreactors (e.g. stirred tanks) are: (1) their construction is simple (2) less maintenance is necessary due to the absence of moving parts (3) airlifts have a well defined flow pattern (4) high values of effective interfacial areas and overall mass transfer coefficients possible (5) high heat transfer rates per unit volume of the reactor can be achieved (6) they have relatively low shear fields - useful for shear sensitive cultures (7) solids can be handled without any erosion or plugging problems (8) airlifts occupy less floor space and are also less costly (9) slow reactions can be carried out due to high liquid residence time (10) comparatively low power inputs for requisite transport rates (11) in airlift reactors the distribution of the energy dissipation rate is fairly uniform in contrast to stirred tank reactors (for example) and (12) extended aseptic operation is possible due to the elimination of stirrer shafts, seals and bearings (Chisti, 1992a).
There are some specific advantages of airlift contactors over other pneumatic reactors (for example, bubble columns). Firstly, airlift devices have a better defined liquid flow pattern than do bubble columns and they also have slightly improved mixing capability. Finally, in some instances, higher mass transfer coefficient is possible in airlifts because of the very high gas velocities which may be employed.

Airlift reactors also possess some disadvantages over other types of bioreactors. These are: (1) they are less flexible than the stirred tank reactor, for example (2) there is usually a high pressure drop in very tall industrial airlifts (3) considerable backmixing and bubble coalescence can be disadvantageous in some cases and (4) the design of airlift reactors is more exacting because the fluid dynamics of these vessels are often difficult to model.

1.4 The Objective of the Study

Although airlift reactors have been employed for various industrial applications (Allen, 1985; Birch et al., 1987; Chisti and Moo-Young, 1991; Margaritis and Wallace, 1984; Ottewell, 1994; Smart, 1984; Smith, 1992; Wood and Thompson, 1986), widespread use of pneumatic reactors commercially is still lacking [airlift reactors and bubble columns only accounted for about 11% of the bioreactors currently used for industrial production (Scragg, 1991)]. This is principally due to a lack of knowledge needed for their design and scale up (Chisti and Moo-Young, 1987; Guy et al., 1986; Siegel et al., 1988; Siegel and Robinson, 1992) and also partially due to the still strongly entrenched industrial tradition of employing mechanically stirred reactors (Guy et al., 1986). The limited information available in the open literature shows a wide variation and contradictions between researchers. Also many of the correlations presented in the literature are restricted in their validity to the same reactor type, size and gas-liquid system employed in their development (Young et al., 1991). Hence, a lot of work still remains to be done to promote a better understanding of the airlift contactors.

In view of the foregoing, the principal objective of this work is to enhance the understanding of the hydrodynamics and mixing of gas-liquid airlift contactors. This dissertation is also concerned, to a lesser extent, with the hydrodynamics and mixing of gas-liquid-liquid airlift reactors.

A detailed set of experimental data of the hydrodynamics and mixing of airlift reactors were obtained with various fluids (Newtonian and non-Newtonian) and with different airlift geometries (draft tube internal and external loop). The hydrodynamic
parameters investigated were gas void fraction and liquid velocity. Liquid mixing (mixing and circulation times and axial dispersion) was also examined. These parameters affect the transport processes (such as mass and heat transfer), shear rate and the intensity of turbulence in the reactor. The effect of gas velocity, electrolyte concentration, liquid viscosity and reactor geometry on the hydrodynamics and liquid mixing were also examined.

Additionally numerical models to study the hydrodynamics and liquid mixing of airlift contactors were employed. A hydrodynamic model was used to estimate the riser gas holdup and liquid velocity in the airlift vessels. Another objective was to develop a correlation for shear rate, which could be used to predict the shear rate in an airlift reactor containing non-Newtonian fluids. A further purpose of the research was to study the liquid mixing by calculating the axial dispersion coefficient and the Bodenstein number in different parts of the airlift contactor, namely the riser and the downcomer.
2 LITERATURE SURVEY

2.1 General

In this section only the basic concepts and background knowledge relevant to the present work is given. A comprehensive literature review on airlift contactors will not be presented in this thesis. There are many good reviews of airlift technology in the published literature (BHRA International Conferences on bioreactor fluid dynamics, 1986, 1988; Blenke, 1979; Chisti, 1989, 1992; Chisti and Moo-Young, 1987; Joshi et al., 1990; Kossen, 1984, 1984a; Merchuk, 1986a; Merchuk and Siegel, 1988; Moo-Young and Chisti, 1988; Onken and Weiland, 1983; Scragg, 1991; Siegel et al., 1988; Siegel and Robinson, 1992); these provide an extensive coverage of the existing literature on airlift reactors. For more detailed information on airlift devices, the reader is referred to these publications.

2.2 Dispersion Characteristics of Airlift Reactors

2.2.1 Introduction

Airlift reactors are pneumatic reactors. In these reactors, the introduction of a gas into the riser section of the reactor produces a pressure difference between the riser and the downcomer sections. This provides the driving force for the circulation of the dispersion. In addition to providing momentum transfer, the gas also sustains mass and oxygen transfer to the liquid.

2.2.2 Bubble Dynamics

Bubble formation, bubble size distribution, bubble breakage, bubble coalescence and bubble rise velocity have a direct bearing on the performance of airlift reactors. Bubble size distribution is dependent on bubble coalescence and bubble formation, and in this section we shall look at these in more detail. The bubble rise velocity is discussed in Section 3.1.2.3.

2.2.2.1 Bubble Size Distribution

Dispersion of the gas entails the generation of bubbles in the riser, of different sizes, which are then carried through the airlift by the fluid circulation. Hence, the mean and the variance of the bubble distribution have a considerable influence on the hydrodynamics and the oxygen transfer in the reactor (Ho et al., 1977). Generally, small
bubbles in the riser provide better mass and oxygen transfer than larger bubbles. This is because small bubbles have a lower rise velocity than larger bubbles, hence their residence time in the riser is usually longer, allowing for a longer period of oxygen transfer to occur (Russell, 1989). Small bubbles also have a larger specific gas-liquid interfacial area for mass transfer than larger bubbles (Bhavaraju et al., 1978). A disadvantage of small bubbles would be that they become depleted of oxygen sooner than larger bubbles (Bhavaraju et al., 1978). Another drawback is that the lower rise velocity of small bubbles will have an effect on molecular diffusion between the gas and liquid phases, reducing the liquid side film coefficient for mass and oxygen transfer (Russell, 1989). In general, the beneficial effects of small bubbles outweigh their shortcomings.

Besides influencing the hydrodynamics and mass transfer, the bubble size distribution also determines the amount of gas recirculation in the downcomer. When a bubble reaches the gas-liquid separator from the riser section, it can either disengage from the liquid or can be recirculated with the liquid flow into the downcomer. At low liquid velocities, only small bubbles are recirculated and they will be depleted of oxygen once they reach the downcomer; for a bubble to be entrained in the liquid downflow, the velocity of the liquid must be greater than the rise velocity of the bubble in the stagnant liquid (Weiland, 1984). At higher liquid velocities, larger bubbles (containing significant oxygen content) will also be recirculated. Therefore bubble size distribution is an important factor in directly influencing gas recirculation.

Many techniques are available to determine bubble sizes. Photographic methods, due to their simplicity, are employed widely (Akita and Yoshida, 1974; Burckhart and Calderbank, 1975; Schugerl et al., 1977). Other techniques involve light scattering and reflection (Calderbank et al., 1960), depolarization (Rodionow and Patzelt, 1969) and optical and electrical probes (Burgess and Calderbank, 1975; Calderbank and Pereira, 1977; Koide et al., 1979; Serizawa et al., 1975; Yamashita et al., 1979). To measure the bubble distribution, various investigators (Sriram and Mann, 1977; Vermeer and Krishna, 1981) have employed a dynamic bubble disengagement technique. Yamashita et al. (1979) have pointed out that though the original bubble size distributions obtained from various techniques differ markedly, the volume-to-surface mean bubble diameters (usually termed the Sauter mean diameter, $d_{sv}$) estimated consequently differ only slightly. They also state that photographic results are in reasonable agreement with that obtained from the two-point electrical probe method. Shah et al. (1982) reported that photographic, electrical and
optical probe techniques give reliable results only in the homogeneous bubbly flow regime (see Section 2.2.3).

2.2.2.2 Bubble Formation and Coalescence

Bubble creation and bubble coalescence directly affect the bubble size distribution in airlift reactors.

Bubble Formation

Bubble formation depends on the type of sparger employed and the size of the sparger orifices. Numerous designs of spargers exist which can be used for gas dispersion in airlift reactors (Chisti and Moo-Young, 1987; Heijnen and Van’t Riet, 1984). Two types of spargers which may be used for gas injection into airlift reactors are: (1) static and (2) dynamic.

Examples of the static type are perforated plates, rings and pipes, porous plates and single hole spargers. Dynamic spargers are depicted by injector nozzles and venturies which disperse gas due to the kinetic energy of a liquid jet (Deckwer, 1992). There are several factors which determine the choice of sparger for a particular application. Chisti and Moo-Young (1987) reported that perforated plates, rings and pipes are cheap to install and operate, while porous plates are more expensive and have higher operating costs due to greater pressure drops through them. Furthermore, porous plates are prone to blockages and can be a source of contamination. However, Chisti and Moo-Young (1987) stated that they have successfully used porous plate aerators (100 μm pore size) in a two litre airlift vessel during cultivation of mammalian cells for monoclonal antibody production. They also reported that cell growth on aerator surface or into pores posed no problems.

Dynamic gas spargers are not commonly used in fermentation practice (Chisti and Moo-Young, 1987). They are more complex to design and build and require some external liquid circulation mechanism, usually a pump (Deckwer, 1992). High shear rates in the dynamic gas spargers rule them out for shear sensitive microorganisms and tissue cultures. Another factor which can determine the type of gas sparger used, is the size of the gas bubble required in the dispersion.

If the bubble size formed from the sparger, in the riser, has an influence on the liquid velocity in that section, then the type of gas distributor would be expected to have an effect on the overall liquid velocity. But, Onken and Weiland (1983) reported that if the gas
phase is distributed uniformly across the riser cross-section, then the type of gas sparger has only a small influence on liquid velocity.

**Bubble Coalescence**

Bubble coalescence is basically the collision and joining of two small bubbles to form one large bubble. Blenke (1985) has elucidated that bubble coalescence occurs due to the thinning of liquid lamella between neighbouring gas bubbles. Chisti and Moo-Young (1987) suggested that in airlift reactors the bubble size in dispersion is usually independent of the size at birth and it is controlled by the equilibrium between the dynamic pressure forces which work to break the bubble and the surface tension force which attempts to preserve its size and shape. There are exceptions to this rule. For example, in strongly non-coalescing media aerated with porous plates the very fine bubble size originating at the sparger persists in dispersion (Chisti and Moo-Young, 1987). Erickson and Desphande (1981) have stated that the bubble size distribution in a given gas-liquid mixture is a result of a balance between the processes of coalescence and breakage.

The rate of bubble coalescence is primarily affected by the composition of the liquid phase and its properties. Erickson and Deshpande (1981) reported that in pure liquids, such as distilled water, the rate of coalescence is high. But the presence of surfactants, electrolytes, sugar solutions, alcohol and surface active agents have been shown to reduce the rate of bubble coalescence (Erickson and Deshpande, 1981; Kelkar et al., 1983; Kennard and Janekeh, 1991; Lessard and Zieminski, 1971; Onken and Weiland, 1981; Posarac and Tekic, 1987; Schugerl et al., 1978; Shah et al., 1982; Snape et al., 1992; Voigt and Schugerl, 1979; Zahradnik et al., 1987).

Coalescence is most common at high bubble densities, because the frequency of collision will be high. Deckwer (1992) has suggested that the rate of bubble coalescence is also influenced by the rate of gas flow into the airlift reactor. Hence, at different gas flow rates different flow regimes prevail due to the rate of coalescence.

**2.2.3 Flow Regimes**

In an airlift reactor, beginning with the introduction of gas into the reactor and as gas flow is increased, several different flow regimes may be observed. Wallis (1969) has characterised the upward movement of the bubble swarms into different separate flow regimes. This definition is incomplete but it is simple and often used (Guy et al., 1986). Four distinct flow regimes occur successively as the gas flow rate increases:
Homogeneous bubbly flow or quiescent bubbling (Figure 2.1a)

At low superficial gas velocities (typically < 5 cm/s; Fair, 1967) homogeneous bubbly flow is observed. This regime is characterised by almost uniformly sized bubbles with equal radial distribution. Levich (1962) has stated that the rise velocity of the bubbles, in this regime for water, lies between 0.18 and 0.30 m/s.

Homogeneous coalesced bubble flow regime (Figure 2.1b)

As the gas flow rate increases (but still < 5 cm/s), the bubble density in the fluid gradually rises and there is greater interaction between them (bubble collision frequencies increase). A transitional regime, known as the coalesced bubble flow regime, characterized by greater turbulence develops.

Churn turbulent or heterogeneous regime (Figure 2.1c)

At higher gas velocities (usually > 5 cm/s; Joshi et al., 1990), the homogeneous gas-liquid dispersion cannot be maintained and unsteady flow pattern with channelling occurs. This heterogeneous flow regime is characterized by large bubbles moving with high rise velocities in the presence of small bubbles (Hills and Darton, 1976). The large bubbles take the form of spherical caps with a very mobile and flexible interface. These large bubbles can grow up to a diameter of about 0.15 m (Shah et al., 1982).

Slug flow (Figure 2.1d)

In small diameter airlift vessels, at high gas flow rates (> 5 cm/s), large bubbles are stabilised by the vessel wall leading to the formation of bubble slugs. Bubble slugs can be observed in columns of diameters up to 0.15 m (Hills, 1976).

The type of flow regime in the reactor is not only determined by the superficial gas velocity but also by other parameters such as fluid properties, sparger design and the velocity of the liquid (Chisti and Moo-Young, 1987). For example, homogeneous flow can be attained at higher gas flowrates if the liquid velocity is increased by some means (Deckwer, 1992). Chisti and Moo-Young (1987) stated that earlier transition to slug flow occurs in pipes of small diameter than in vessels of larger size and spherical cap bubbles form more readily in highly viscous fluids than in water-like systems. The transition from one flow regime to another can also be influenced by the presence of suspended solids (Vance et al., 1981).
2.2.4 Gas Velocity in Airlift Reactors

The amount of energy in a pneumatic reactor is a function of the power imparted to the fluid in addition to the momentum transport characteristics of the fluid itself. The power input to an airlift reactor containing a batch of liquid results solely from the gas supplied to it. In an airlift operated batchwise with respect to liquid all the power supply originates from two sources:

1) isothermal expansion of the gas as it moves up the reactor, and
2) the kinetic energy of the gas injected into the reactor.

Hence, the total power input into the reactor may be shown to be (Chisti and Moo-Young, 1987)

\[ P_z = \frac{Q_m}{M} R_v \left( \ln \left( \frac{P_b}{P_h} \right) + 0.5 \eta U_{or}^2 \right) \]

where \( P_b \) and \( P_h \) are the bottom pressure and headspace pressure of the fluid in the vessel, respectively. \( U_{or} \) is the gas velocity in the orifice of the sparger which relates to the gas velocity just above the orifice through the efficiency \( \eta \) (Moo-Young and Blanch, 1981 have given \( \eta \) to be about 0.06). \( Q_m \) is the molar gas flow rate and \( M \) is the molar mass of the gas. It should be pointed out that even when the molar gas flow into a reactor is
constant the power input to the fluid declines with increasing reactor head-space pressure and the reactor becomes markedly less turbulent.

Chisti (1989) has shown that the primary source of power in the reactor is the isothermal expansion of the injected air. Furthermore, he stated that the kinetic energy contribution to power input is below 1.5% of the total power. Joshi and Sharma (1979) have also stated that under normal operating conditions, the kinetic energy of the gas is negligible. Therefore the kinetic energy contribution can be neglected in the calculation of specific power input (i.e. the second term on the right hand side of Eqn 2.1). The specific power input into pneumatic reactors can be obtained via the following equations (Chisti, 1989):

For bubble columns having a constant cross section

$$\frac{P_g}{V} = \rho_l g U_g$$  \hspace{1cm} 2.2

and for airlift reactors for which the superficial gas velocity ($U_{sg}$) is usually based on the riser cross-section, Eqn 2.2 modifies to

$$\frac{P_g}{V} = \frac{\rho_l g U_{sg}}{1 + \frac{A_g}{A_r}}$$  \hspace{1cm} 2.3

where $P_g$ is the power input due to gassing and $V$ is the volume of liquid in the column. It should be noted that the specific power input for a given fluid, is directly dependent on the gas velocity (Eqns 2.2 and 2.3). Therefore, $P_g/V$ and $U_{sg}$ can be interchangeably used in correlations of reactor performance with appropriate correction factors (Chisti, 1989).

The average superficial gas velocity for an airlift reactor with unaerated liquid height ($H$), operating at headspace pressure ($P_h$) is given by (Chisti and Moo-Young, 1987):

$$U_{sg} = \left( \frac{Q_a R_t t_i}{A H \rho_l g} \right) \ln \left[ 1 + \left( \frac{\rho_l g H}{P_h} \right) \right]$$  \hspace{1cm} 2.4

where $t_i$ is the reactor temperature and $A$ its cross-sectional area.

The location of the gas sparger

The gas sparger in airlift reactors is normally located at the base of the riser, as in bubble columns. But Chisti and Moo-Young (1987) have shown that this is not the best location, because the recirculating fluid flowing from the downcomer over the gas stream...
emerging from the sparger causes a maldistribution of the gas to the far wall of the riser. The gas thus concentrated along the wall of the riser coalesces into large bubbles and channels up the wall for some distance before the turbulent forces, in the reactor, redisperse it over the entire riser cross-section. This poor gas distribution has a marked detrimental effect on the performance in small reactors (used commercially for the production of speciality products) and in tall airlift devices, when low gas flow rates are employed (for animal tissue culture and plant cell fermentations), the effect of gas distribution is still important (Chisti and Moo-Young, 1987).

Gas distribution can be improved by placing the sparger slightly inside the riser (Chisti, 1989; Chisti and Moo-Young, 1987). Another approach would be to position the gas sparger some distance up the riser. This would reduce power consumption (Chisti and Moo-Young, 1987). The region of the riser below the sparger could be aerated either by recirculating gas from the downcomer or by supplementary spargers located either in the downcomer or further down in the riser.
2.3 Design and Operating Parameters of Airlift Contactors

The design and performance evaluation of gas-liquid or gas-slurry airlift reactors depend on a knowledge of the following principal parameters: gas holdup, liquid circulation rate and liquid mixing. These parameters are strongly dependent on the gas flow rate, chemical and physical properties of the system, and reactor design and geometry. The mass and fluid-wall heat transfer processes in the airlift device are determined by the three principal parameters. A thorough knowledge of these interdependent parameters is also necessary for a proper scaleup of these reactors. This section considers these design parameters.

2.3.1 Momentum Transfer

2.3.1.1 Gas Holdup

Gas holdup (or gas void fraction) is one of the most important and frequently used parameters characterising the hydrodynamics of airlift reactors. Hence it is the most studied parameter of airlift columns. It can be defined as the percentage by volume of the gas in the two or three-phase mixture in the reactor. The importance of gas holdup is multifold. It is used to compute the residence time of each phase and to evaluate the interfacial area and subsequently the mass and heat transfer rates between the gas and the liquid phases. Holdup also influences the liquid velocity and mixing in the reactor.

A number of investigators (Merchuk and Stein, 1981; Onken and Weiland, 1983; Verlaan et al., 1986) have found that radial gas holdup profile across the riser is rather flat, but the holdup at the wall is slightly higher than in the rest of the column. Lippert et al. (1983) have reported the variation of radial gas holdup in the riser section of the airlift, while Chisti (1989) stated that axial variations of gas holdup in the riser and downcomer occur and higher gas holdups near the core of the column than closer to the wall have been observed. Generally, gas holdup values can vary within a reactor and hence average values are usually reported (Merchuk, 1990).

The techniques for the measurement of gas holdup can be classified into two categories: local or global measurements.

Local measurement

The gas holdup can be measured locally by measuring the pressure drop along the flow path. The distribution of gas holdup along the column is given by the local holdup (Merchuk and Stein, 1981). The average volumetric gas holdup, for example in the riser
section of the airlift reactor, can be determined by measuring the differential hydrostatic pressure at two points along the riser section of the column.

Among the local measurement methods, the manometric method is the most used. This method is discussed below in detail. Another local method is based on the measurement of the electrical conductivity which depends on the concentration of each phase (Guy et al., 1986). Conductivity probes have been employed in bubble columns by Hills (1974), and Ueyama and Miyauchi (1977). One of the disadvantages of using conductivity probes is that the liquid has to be electrically conductive. A new promising technique using optical probes can avoid this difficulty (Guy et al., 1986).

**Manometric method**

The determination of the individual riser and downcomer gas holdups can be achieved by the manometric methods. Either U-tube or inverted U-tube manometers can be used. The U-tube manometers are usually filled with mercury and the inverted U-tube manometers are filled with the liquid in the column.

For the U-tube manometer the gas holdup (riser) can be calculated from the following equation (Chisti, 1989):

$$
\varepsilon_s = \frac{\left( \rho_M - \rho_L \right) dh}{\left( \rho_L - \rho_G \right) dz}
$$

and for the inverted U-tube manometer, the gas holdup (riser) can be calculated as follows (Chisti, 1989):

$$
\varepsilon_s = \frac{\rho_L \ dh}{\left( \rho_L - \rho_G \right) dz}
$$

The reliability and the simplicity of the manometric methods is certified by the frequency with which they have been used by many investigators (Bello, 1981; Chisti, 1989; Glennon et al., 1993; Hills, 1976; Hsu and Dudukovic, 1980; Kawase and Moo-Young, 1986; Kemblowski et al., 1993; Siegel and Merchuk, 1987; Young et al., 1987).

**Global measurement**

In the global method, the overall gas holdup is determined by measuring the difference between the ungassed and gassed liquid levels, rather than from pressure measurements. This method does not differentiate among the different sections of the airlift reactor and hence is not as useful as local measurements (Merchuk, 1990).
The overall gas holdup, refers to the reactor as a whole, is given by the following equation:

$$\varepsilon_o = \frac{V_g}{V_g + V}$$

where $V_g$ and $V$ are the gas and the liquid volumes in the reactor, respectively. In airlift reactors, individual riser gas holdup ($\varepsilon_g$) and downcomer gas holdup ($\varepsilon_d$) can also be identified and are related to the overall holdup (Chisti, 1989) via

$$\varepsilon_o = \frac{A_r \varepsilon_g + A_d \varepsilon_d}{A_r + A_d}$$

The more traditional and still most frequently used global method is the volume expansion technique. Another way of measuring the overall gas holdup is by the gamma ray attenuation technique.

**Volume expansion**

In this method, the measurement of unaerated, static, liquid height ($H$) and the height of gas-liquid dispersion ($H_D$) upon aeration can be used to calculate the overall average value of the holdup ($\varepsilon_o$) as follows:

$$\varepsilon_o = \frac{(H_D - H)}{H}$$

Chisti (1989) reported that, using this method, the gas holdup was reproducible to within ±10% even at the highest gas flows at which there were pronounced fluctuations in the dispersion level. This method makes no allowance for such end effects as the presence of a foam cap (Hills, 1974).

**Gamma ray absorption**

Measurements of the attenuation of a beam of gamma radiation enable the mean holdup along the path between the source and the detector to be determined. This method has been successfully used in a bubble column by Lockett and Kirkpatrick (1975) and has been thoroughly discussed by Hewitt (1978).

The gas holdup depends primarily on the rate of gas input into the reactor (Siegel et al., 1988). In general, the gas holdup in an airlift reactor increases with increasing superficial gas velocity. At a given gas velocity the holdup in the riser can be influenced by the bubble size distribution, the gas recirculation rate, the liquid velocity and the liquid physical properties (Onken and Weiland, 1983). Gas holdup has also been shown to vary
Literature Survey

with time of gas flow in viscous liquids (Philip et al., 1990). The geometric design of the airlift reactor also has a strong effect on the riser gas holdup (Merchuk, 1990).

The gas recirculation rate, in an airlift, will determine the downcomer gas holdup and hence the overall reactor gas holdup. The gas-liquid separator size and configuration influence the gas recirculation rate (Chisti and Moo-Young, 1993; Siegel et al., 1986). Siegel et al. (1986) reported that the gas recirculation is dependent on the fluid residence time in the separator. They also observed that the liquid level in the separator was found to have an influence on the recirculation rate. Higher liquid levels led to longer fluid residence times and decreased gas recirculation.

The gas holdup in the downcomer is always less than that in the riser due to gas disengagement in the separator. In the external loop airlift only a small fraction of the riser gas is recirculated into the downcomer whereas in the internal loop a much larger fraction of the riser gas is entrained in the downcomer. Bello et al. (1985) reported that gas holdup in a concentric draft tube internal loop reactor can be about 80-95% of the riser holdup, while Frohlich et al. (1991) stated that 85% of the riser gas holdup was recirculated into the downcomer of a 4 m³ tower loop reactor.

The downcomer gas holdup is strongly dependent on the geometry of the top section, bubble size distribution, liquid properties and liquid velocity (Siegel et al., 1988). The velocity of the liquid must be greater than the terminal rise velocity of the bubble in stagnant liquid for the bubble to flow downward. Hence, a certain liquid velocity must be reached before any gas is recirculated. A number of investigators (Bello et al., 1985; Chakravarty et al., 1973; Siegel et al., 1986) have stated that, for the bubble sizes normally found in the downcomer, the liquid velocity must exceed about 20-30 cm/s. As the liquid velocity increases, recirculation of larger bubbles into the downcomer begins to occur. The presence of larger bubbles is advantageous as they have more oxygen than smaller ones. Hence, it may be desirable to maintain a high liquid velocity in the downcomer (Russell, 1989).

As the riser liquid velocity in the reactor increases, the riser gas holdup decreases due to the reduction in the residence time of bubbles in the riser section (Onken and Weiland, 1981). The downcomer holdup also decreases with increasing downcomer liquid velocity.

The physical properties of the liquid in the column, such as viscosity and ionic strength, can also affect gas holdup. This is because liquid viscosity and ion strength (electrolyte concentration) influence bubble breakup and coalescence which, in turn, alter
bubble size. It has been shown by a number of workers (Marrucci and Nicodemo, 1967; McManamey et al., 1984; Onken and Weiland, 1980; Shah et al., 1982; Snape et al., 1992) that gas holdup in a coalescent liquid (e.g. pure water) is lower than that in a non-coalescent liquid (e.g. salt solution) under identical conditions. This is primarily due to large bubbles occurring in the former; since large bubbles, due to their large rise velocity and shorter residence time, tend to reduce gas holdup. The addition of antifoam (coalescence promoting substance) to a medium reduces gas holdup, due to bubble coalescence and increased average bubble size (Moo-Young et al., 1987; Schugerl et al., 1978; Suh et al., 1992).

Most researchers agree that the viscosity of the medium can also influence gas holdup. But in the literature, there are a lot of contradictions between investigators on the influence of viscosity on gas holdup. Philip et al. (1990) reported that in their experiments the gas holdup, at a given superficial gas velocity, increased with liquid viscosity. Deckwer (1992), Fields et al. (1984), Godbole et al. (1984), Haque et al. (1986), Heijnen and Van't Riet (1984), Kawase et al. (1987), Kawase and Moo-Young (1990), Popovic and Robinson (1984), Schumpe and Deckwer (1982) and Shah et al. (1982) have stated an opposite effect. Deckwer (1992) and Heijnen and Van't Riet (1984) explained that an increase in viscosity leads to the formation of large bubbles which reduce gas holdup. However, a high liquid viscosity may reduce bubble rise velocity (due to viscous drag) and this should enhance holdup; but they reported that the former seems to be predominant over the latter effect. Other investigators (Guy et al., 1986; Onken and Weiland, 1980) found their gas holdup data to be independent of liquid viscosity. Furthermore, with CMC solutions in a bubble column Godbole et al. (1982) found, for increasing apparent liquid viscosity, the gas holdup to increase to a maximum and then decrease. From the above, it seems that the effect of viscosity on gas holdup in pneumatic reactors is still not clear and requires further study (Russell, 1989).

The gas holdup has been shown to vary with time in columns containing viscous Newtonian and non-Newtonian liquids (Philip et al., 1990). Philip et al. (1990) reported that two classes of bubbles occur in columns containing viscous Newtonian and non-Newtonian liquids. These were large bubbles with ellipsoidal caps or slugs having an equivalent diameter of 20 mm and nearly spherical small bubbles with diameters in the range 100 μm to 1 mm. The interaction between the two groups of bubbles and their combined effect on gas holdup and liquid velocity in the reactor were complex. They
demonstrated that the holdup increased gradually to a steady state value over a period varying from 15 minutes to 1 hour and that the rate of increase in holdup was a function of liquid viscosity and superficial gas velocity. They explained that the holdup varies with time primarily due to the accumulation of small bubbles in viscous liquids. In viscous liquids, Philip et al. (1990) stated that when the gas flow is turned on, the immediate rise in the holdup is due to the large bubbles and that the small bubbles are almost absent. They further stated that with an increase in aeration time, there is a progressive increase in the concentration of small bubbles. Some of the small bubbles coalesce and leave the system. The rate of coalescence and hence of disengagement, being dependent on the concentration of small bubbles, eventually attains equilibrium with the rate of generation. The gas holdup is then constant and has attained a stable value. When measuring the overall or local gas holdup value in an airlift with viscous liquids, care must be taken to ensure that only the steady state value of the holdup is recorded. This way 'erroneous' holdup values can be avoided. It is surprising that there is little mention of this in the open literature. Schugerl (1981) studied the oxygen transfer in highly viscous fluids. He reported that, after a long period of aeration, the accumulation of small bubbles in viscous liquids can have a large contribution to the gas holdup. Previous investigators, apart from Guy et al. (1986), who have measured gas holdup values with viscous liquids in airlift reactors have not distinguished whether they measured the stable or the unstable value of gas holdup. Guy et al. (1986) stated that they obtained the gas holdup just after turning on the gas flow. Hence, avoiding problems due to the evaporation of water, changes in concentration and accumulation of small bubbles due to bubble breakage. However, since Guy et al. employed viscous liquids they have recorded the unstable value of gas holdup.

The gas holdup has been found to be strongly influenced by some aspects of the geometric design of the airlift reactor. Changes in the ratio of the downcomer to riser cross sectional area ($A_d/A_r$) will change the gas and liquid residence times in each part of the reactor and hence their contributions to the holdup. A number of investigators (Bello et al., 1985; Chisti and Moo-Young, 1987; Siegel et al., 1986, 1988; Weiland, 1984) have stated that a decrease in $A_d/A_r$ will increase the overall, riser and downcomer holdups and decrease the riser and downcomer liquid velocity. There are several reasons for the influence of $A_d/A_r$ on gas holdup (Siegel et al., 1988): Firstly, since for a smaller $A_d/A_r$, the riser liquid velocity is lower this leads to an increased gas residence time in the riser. Also for smaller $A_d/A_r$, the liquid velocity in the downcomer is greater (which means the fluid residence time in the downcomer is decreased), hence more gas will be entrained in the
downcomer increasing both the riser and downcomer gas holdup. Thirdly, since for a smaller $A_d/A_r$, the downcomer holdup is greater this will lead to a reduction in the driving force for liquid circulation and therefore the overall liquid velocity will decrease. With decreasing $A_d/A_r$, the relative contribution of the riser to the overall holdup increases while the contribution of the downcomer (with its lower gas concentration) decreases. Finally, the smaller the $A_d/A_r$, the closer the airlift resembles a bubble column (where $A_d/A_r = 0$) and hence the greater the gas holdup.

It has been reported that the presence of a draft tube in a bubble column (which transforms the bubble column into an internal loop airlift) has no effect on the gas holdup (Bello et al., 1985; Chisti and Moo-Young, 1987). Chisti (1989) stated that this could be due to an increase in liquid velocity in the airlift which might have reduced the riser gas holdup because the bubble rise velocity had increased. A number of investigators (Chisti and Moo-Young, 1987; Erickson and Deshpande, 1981; Russell et al., 1994; Siegel and Merchuk, 1991) have shown gas holdup and liquid velocity to be independent of the aerated and unaerated liquid heights above the draft tube. This could be due to the two zone flow pattern where at a given gas velocity the height of the zone of the circulating flow does not increase above a critical height (Russell et al., 1994). Hence, the residence time of fluid in separator will not considerably change as the gas-liquid separator is enlarged. Russell et al. (1994) also claimed that the draft tube height has no effect on the overall gas holdup in a concentric tube airlift reactor. But they stated that the downcomer holdup at a given gas velocity increased slightly with draft tube height because the liquid velocity increased with draft tube height. They further explained that as the overall gas holdup remained constant and since the downcomer gas holdup increased with increasing draft tube height, the riser holdup must decrease. The changes in riser and downcomer gas holdups with increasing draft tube height will counter each other and therefore the overall gas holdup will not change.

To estimate the gas holdup, a number of correlations have been presented by different authors. Most of these relationships consider the holdup as a function of the superficial gas velocity. These correlations are discussed fully in Section 2.4.

2.3.1.2 Liquid Velocity

The liquid velocity in airlift contactors arises from the difference in the bulk densities of the fluids in the riser and the downcomer (de Nevers, 1968; Freedman and Davidson, 1969) and from the liquid-wake associated with the rising bubbles (Ayazi
The fluid circulates along a well defined path, upflow in the riser and downflow in the downcomer. Liquid circulation velocity is the characteristic which distinguishes airlift devices from other gas-liquid contacting reactors. The liquid velocity is a function of the superficial gas velocity, gas holdup and the pressure drop along the flow path (Siegel et al., 1988). The velocity is also influenced by properties of the circulating liquid, airlift reactor geometry and gas recirculation.

Liquid velocity, in turn, influences the gas holdup in the vessel, the prevailing flow regime, heat and mass transfer and the extent of mixing in the reactor. In addition the circulation affects the turbulence in the reactor. Hence the shear forces to which the microorganisms are exposed. In some cases, the morphology and/or physiology of the microorganisms are also affected (Siegel et al., 1988).

There are two ways of depicting the liquid circulation velocity: the superficial liquid velocity or the linear liquid velocity (also known as the interstitial velocity). The superficial liquid velocity is the volumetric liquid flow rate divided by the cross-sectional area of the section (riser or downcomer) through which it is flowing. This velocity does not take into account that the cross-sectional area available for liquid flow is slightly less than the riser or downcomer cross-sectional area due to the area taken up by the gas. Conversely, the linear liquid velocity is the true velocity of the liquid. The relationship between the interstitial liquid velocity and the superficial liquid velocity is shown below (Chisti, 1989):

\[
V_{lr} = \frac{U_{id}}{1 - \varepsilon_g}
\]  

\[
V_{ld} = \frac{U_{id}}{1 - \varepsilon_d}
\]

Generally, the linear liquid velocity increases markedly with increasing superficial gas velocity up to about 0.05 ms\(^{-1}\) (Gopal and Sharma, 1982; Kloosterman and Lilly, 1985; Merchuk and Stein, 1981; Verlaan et al., 1986; Weiland, 1984). At higher gas flow rates (above 0.05 ms\(^{-1}\)), the liquid velocity still increases with gas velocity but the rate of increase is reduced (Weiland, 1984). Several investigators (Chisti, 1989; Merchuk and Stein, 1981; Siegel et al., 1988) stated that this was probably due to the change in the prevailing flow regime at higher gas velocities. Merchuk and Stein (1981) also added that at higher gas velocities the presence of large bubbles give rise to greater turbulent energy dissipation in the reactor and therefore the proportion of energy input available for liquid motion will be lower.
The properties of the liquid medium can also influence the liquid velocity. Usually, the liquid velocity, at a fixed superficial gas velocity, decreases with increasing liquid viscosity (Allen and Robinson, 1989; Glennon et al., 1988; Koide et al., 1988; Li et al., 1995; Onken and Weiland, 1983; Philip et al., 1990; Popovic and Robinson, 1984; Weiland, 1984). This is probably due to the growing frictional resistance to flow around the loop with increasing liquid viscosity. However, some researchers (Fields et al., 1984; Kawase and Moo-Young, 1986b; Wachi et al., 1991) have claimed the opposite, for low viscosity pseudoplastic media (xanthan gum and CMC solutions). They stated that, apparently the liquid velocity increases with increasing liquid viscosity due to increased density difference (between riser and downcomer) and viscous drag reduction. For example, Fields et al. (1984) reported that the liquid velocity, in a concentric draft tube airlift, was higher with xanthan gum (0.1%) solution than with water, due to a reduction of viscous drag. Wachi et al. (1991) stated that, in their case, the liquid velocity was higher with CMC solution (0.15% and 0.3%) than with water because of increased density difference with the former. This was due to the fact that, with CMC solution, there was a reduction in downcomer holdup since the large bubbles in the riser disengaged in the separator. From the above, it seems that the effect of liquid viscosity on liquid velocity is not clear.

The presence of electrolytes and/or alcohols in the liquid medium also has an effect on the liquid velocity. A non-coalescent liquid will give a lower liquid velocity (Wachi et al., 1991; Weiland, 1984). This is because the inhibition of coalescence results in small bubbles in the riser which will recirculate into the downcomer giving a higher downcomer holdup. Higher holdup in the downcomer will result in a lower driving force for liquid circulation.

The liquid velocity can also be influenced by reactor geometry. In general, the downcomer to riser cross-sectional area ratio \((A_d/A_r)\) strongly affects the liquid velocity. The liquid velocity increases with increasing \(A_d/A_r\) ratio (Bello et al., 1984, 1985; Siegel et al., 1988; Weiland, 1984). This is due to the fact that an increase in \(A_d/A_r\) ratio results in a reduction in the magnitude of the liquid velocity in the downcomer relative to that in the riser. This means that fewer bubbles will recirculate into the downcomer giving a lower downcomer holdup relative to the riser holdup. Hence a larger density difference for liquid circulation is obtained.

Other reactor geometrical parameters which are known to affect liquid velocity are reactor height (Faust and Sittig, 1980; Siegel et al., 1988; Weiland, 1984) and the height of
the draft tube (Russell et al., 1994; Weiland, 1984). Due to the increase in the hydrostatic pressure difference between riser and downcomer, liquid velocity increases with reactor height and with draft tube height. The liquid circulation velocity is not usually affected by the type of gas sparger employed, providing that the riser cross section is uniformly sparged (Chisti, 1989; Merchuk, 1986; Onken and Weiland, 1983).

The magnitude of the liquid circulation velocity is one of the most important design and scale up parameter for airlift reactors (Merchuk, 1990). Numerous investigators have developed hydrodynamic models and correlations for the prediction of local liquid velocity in the riser and/or the downcomer section of the airlift. This is discussed more fully in Section 2.4. In this section, the calculation of the average overall liquid circulation velocity is considered. Blenke (1979) defined average circulation velocity \( V_i \) as

\[
V_i = \frac{L}{t_c}
\]

where \( L \) is the circulation path length and \( t_c \) is the average time for one complete recirculation. The mean circulation times may be obtained by employing tracer response techniques (Bello et al., 1984; Verlaan et al., 1986). A dampened or decaying sinusoidal type of response can be detected at some location downstream of the tracer injection point (Blenke, 1979; Chisti and Moo-Young, 1987). The time difference between adjacent peaks being equal to the circulation time (Weiland, 1984). It must be noted that values of liquid velocity either in the riser or downcomer are more useful than the average overall liquid circulation velocity (Chisti, 1989). Other methods which have been employed for the measurement of the liquid circulation velocity are:

1) flow-followers (Choi et al., 1995; Fields et al., 1984; Vatai and Tekic, 1986);
2) thermal method (Douek and Livingston, 1994; Ford et al., 1972);
3) conductivity method (Choi and Lee, 1993; Shi et al., 1990);
4) pH method (Verlaan et al., 1989);
5) magnetic flow meters (van der Lans and Smith, 1982; Verlaan et al., 1986);
6) ultrasonic flow measuring devices (Popovic and Robinson, 1988); and
7) paddle wheel meters (Siegel et al., 1986).

From the above, only the thermal, conductivity and pH methods can also be used for the measurement of mixing time and the liquid phase axial dispersion coefficient (Ford et al., 1972; Hoogendoorn and Hartog, 1967). These techniques are discussed in more detail in Section 2.3.1.3.
2.3.1.3 Mixing

The nature and degree of liquid phase mixing are important in airlift reactor design and operation. This parameter is particularly significant during the process of scale up from laboratory scale to industrial scale reactors, since the former exhibit a greater degree of mixing than the latter (Siegel et al., 1988). In an airlift, mixing is affected by turbulence and liquid velocity in the reactor (Chisti and Moo-Young, 1988). In turn, mixing influences the driving force for heat and mass transfer (Chisti, 1989). It also affects the biological parameters such as shear damage, growth rate, gas transfer and control of pH in the reactor (Pollard, 1995).

The overall mixing process of a fluid within a given phase (due to non ideal flow conditions) in an airlift reactor, is divided into two phenomena (Fields and Slater, 1984; Levenspiel, 1972; Shah et al., 1978; Smith, 1981): micromixing and macromixing.

Micromixing involves the mixing of different components on the molecular level. It includes all aspects of mixing such as molecular diffusion, which is not defined by the residence time distribution. The extent of micromixing is dependent only upon the time of association between various molecules. Complete segregation and maximum mixedness are the two extreme conditions of micromixing. In the first case mixing of the various entities occurs as late as possible, whereas in the latter case the mixing occurs immediately upon entering the system. To quantify micromixing it would be necessary to obtain the local velocity and mixing rate on the molecular level. But, this information is very difficult to acquire in an actual airlift (Siegel et al., 1988). Hence, usually all the experimental results on mixing in airlift contactors is concerned with macromixing.

The macromixing process considers the fluid as independent entities and provides information on the residence time experienced by each. The entities are infinitesimal compared with the liquid bulk but large enough to contain an enormous amount of molecules. Each entity behaves like an individual batch reactor, with its own identity. Macromixing occurs solely as a result of convective diffusion. Two extreme cases of macromixing which could possibly exist in a flow system are plug flow and fully backmixed (Levenspiel, 1972). In the first case, there is complete radial (perpendicular to the direction of flow) mixing but longitudinal (along the flow direction) mixing is nonexistent. Due to complete radial mixing all fluid components within the system have identical velocities and therefore identical residence times (Shah et al., 1978). In the fully backmixed state, the concentrations are uniform throughout the body of the fluid. In this state, the residence time distribution of the fluid follows an exponential decay, with the exit
stream composition being identical to that within the system (Shah et al., 1978). In practice, however, flows in airlift reactors deviate considerably from the above extreme cases of macromixing (Chisti, 1989; Shah et al., 1978). Shah et al. (1978) suggested that these deviations may be the result of nonuniform velocity profiles, channelling, fluctuations in velocity due to molecular and turbulent diffusion, reactor geometry, fluid backflow due to velocity differences between phases, and gas recirculation.

In general, under batch operation conditions micromixing and macromixing are similar with respect to conversion and product distribution. However, the two types of mixing in a continuous flow process will only be similar for events of first-order reaction kinetics with respect to conversion and product distribution (Levenspiel, 1972).

Mixing in an airlift reactor is primarily due to axial dispersion in the riser and downcomer sections, turbulent mixing in the gas-liquid separator and backmixing caused by liquid recirculation (Fields and Slater, 1983). Mixing may be characterised, in airlift reactors, in terms of a mixing time (\( t_m \)), and a liquid phase axial dispersion coefficient (\( D \)). Another important parameter for the design and operation of airlift reactors is the circulation time (\( t_c \)) (Onken and Weiland, 1981).

The mixing time (\( t_m \)) is defined as the time required by a mixed fluid to reach a specified degree of homogeneity after a tracer pulse has been added to the airlift reactor (Guy et al., 1986; Kawase et al., 1994). As the tracer pulse circulates around the loop, it mixes into the liquid phase and therefore the concentration of the pulse decreases to an equilibrium value in the vessel (Onken and Weiland, 1983). The degree of homogeneity is defined as the relative deviation of the tracer concentration of the pulse (C) at some time after injection from the equilibrium concentration after complete mixing (\( C_e \)) (Russell, 1989):

\[
Homogeneity = \frac{C - C_e}{C_e} \quad 2.13
\]

Usually, the mixing time is measured for 90 or 95% homogeneity (Siegel et al., 1988). Mixing time is one of the most relevant characteristics of gas-liquid and gas-slurry reactors, particularly in biotechnology (Kawase and Moo-Young, 1989; Petrovic et al., 1990). It is a direct measure of the homogeneity of the component concentration (such as dissolved oxygen and substrates, microorganisms) in a fermentation medium. Knowledge of mixing time gives useful information about the distribution of the concentrations of acids or alkalis which have been added into the reactor for the purpose of pH control (Kawase et al., 1994).
Another useful parameter for the characterization of a mixing efficiency of an airlift reactor is the circulation time. It is defined as the time necessary for a fluid element to complete a closed loop. In practice, the circulation time is taken as the time interval between two crossings, in the same direction, of an appropriate reference plane (Carreau et al., 1976). In bubble columns, the circulation time (or the related liquid circulation velocity) has been used as a correlating parameter for the mixing time (Pandit and Joshi, 1983). Hague et al. (1987) stated that the following correlation can be used to relate mixing time to the circulation time.

\[ t_m = N t_c \]  \hspace{1cm} 2.14

However, the relation between the mixing and circulation time has not yet been clearly established, especially in the case of non-Newtonian liquids.

The nature of mixing phenomena in airlift reactors is clearly shown (Fig 2.2) by the typical response to a pulse tracer input into an airlift reactor. As shown in Figure 2.2, in airlift reactors a definite liquid recirculation is superimposed on mixing by dispersion, which leads to the characteristic decaying sinusoidal tracer response pattern shown. The calculation of mixing time for 90% difference between the final and instantaneous tracer concentrations can be achieved directly from the tracer response curve. Similarly, the circulation time (the distance between adjacent tracer output peaks) can be read directly and converted to circulation velocity from a knowledge of the length of the circulation path.

The mixing time has been shown to depend on the superficial gas velocity, volume of the contactor, the column height to diameter ratio and the liquid properties (Pandit and Joshi, 1983; Weiland, 1984). Fields and Slater (1983) demonstrated that the presence of a stationary cloud of bubbles in the annulus can also affect the mixing characteristics of a draft tube airlift reactor. Pandit and Joshi (1983) expressed the mixing characteristics of the liquid in a bubble column by means of a hydrodynamic model. Pandit and Joshi (1983) studied the effect of various design parameters on mixing time and found that: (1) mixing time decreases initially with increasing superficial gas velocity (up to about 0.2 m/s) and then increases with further increase in gas velocity; (2) mixing time increases with increasing ratio of the column height to diameter; (3) when the column height to diameter ratio remains constant, the mixing time increases with increasing column diameter; (4) mixing time was found to increase in the presence of electrolytes (sodium sulphate and sodium chloride) and alcohols (ethyl and butyl alcohols); (5) mixing time increased with an
increase in Newtonian liquid viscosity (at lower gas velocities the effect of viscosity was much more pronounced than at higher gas velocities) and with non-Newtonian liquids, the mixing time also increased with increasing consistency index.

Other studies (Margaritis and Sheppard, 1981; Onken and Weiland, 1980; Rousseau and Bu’lock, 1980; Weiland, 1984) have found mixing time to decrease sharply with an increase in superficial gas velocity (up to about 0.04 m/s) and for further increases in gas velocity the mixing time decreased at a slower rate. Margaritis and Sheppard (1981) suggested that this was due to a change in the flow regime from bubbly to churn turbulent. Russell (1989) and Weiland (1984) proposed that this division of mixing times into two regimes indicated that the mixing process was dominated by the circulation of the fluid phase (giving dispersion at the end zones) rather than axial dispersion due to ascending bubbles. Velan and Ramanujam (1995) also found, for a draft tube internal loop airlift reactor, the mixing time to decrease with an increase in the superficial gas velocity.

Figure 2.2: An ideal response to a tracer pulse input in an airlift reactor.
Liquid mixing is affected by the geometry of the airlift reactor. The reactor height has been shown to have a pronounced effect on the mixing time (Onken and Weiland, 1983; Rousseau and Bu'lock, 1980). Mixing time increased with reactor height to the power of 1.7 (Rousseau and Bu'lock, 1980). Several investigators (Chisti, 1989; Popovic and Robinson, 1993; Russell et al., 1994; Weiland, 1984) have stated that the mixing time decreased with increasing liquid height above the draft tube, to a height above which no further improvement is noticed. Chisti (1989) suggested that this confirmed the existence of two distinct zones in the top section. Two zones existed, when the height of the fluid in the top section was above a certain critical height and the recirculating fluid passed through the lower region, bypassing the upper region. The recirculating flow moved through the entire top section when the liquid level above the draft tube decreased below the critical height.

Mixing time increases with an increase in draft tube height (Onken and Weiland, 1983; Russell et al., 1994; Sukan and Vardar-Sukan, 1987). This is due to the fact that an increase in the draft tube height increases the circulation path length and hence lengthens the distance the tracer pulse has to travel between the end sections of the device, where most of the dispersion occurs; thus increasing the mixing time (Onken and Weiland, 1983). Mixing in an airlift will improve as the ratio of the downcomer to riser cross-sectional area ($A_d/A_r$) decreases. This was due to a higher difference between the riser and downcomer liquid velocities and increased mixing at the end sections (Margaritis and Sheppard, 1981; Weiland, 1984). Mixing time was also found to decrease with an increase in the draft tube to column diameter ratio (Velan and Ramanujam, 1995).

Although mixing time usually increases with airlift reactor volume, some investigators (Rousseau and Bu'lock, 1981; Weiland, 1984) observed the specific mixing time (mixing time per unit liquid volume) to decrease as the working volume of the reactor increased.

The effect of liquid properties on mixing time in airlift reactors is not clear. Pandit and Joshi (1983) and Weiland (1984) reported that liquid properties influenced the mixing time. However, Onken and Weiland (1980) stated that mixing time was not affected by liquid properties. The mixing time was found to increase with an increase in the Newtonian and non-Newtonian liquid viscosity (Guy et al., 1986; Kawase et al., 1994; Li et al., 1995; Pandit and Joshi, 1983). But, Russell (1989) observed that the mixing time decreased with an increase in the non-Newtonian viscosity of the fermentation broth. Apparently, this was due to increased turbulence from the formation of large bubbles and increased liquid
velocity as the viscosity of the broth increased. Velan and Ramanujam (1995) also observed a decrease in the mixing time with an increase in the shear thinning of the media (CMC solutions). From the above, it seems that the influence of liquid properties (certainly liquid viscosity) on mixing time is not clearly understood and requires further work.

For the measurement of mixing time (or circulation time or liquid phase axial dispersion coefficient) different methods can be employed (Ford et al., 1972; Hoogendoorn and Hartog, 1967). All these methods consist in measuring the time between the moment when a characteristic property of the liquid is locally modified and the moment when it becomes homogeneous with regard to this property. Most of the previous researchers have used one of the following methods (Ford et al., 1972; Hoogendoorn and Hartog, 1967):

1) The conductivity or pH method
The conductivity technique involves the detection of a change in the electrical conductivity as mixing proceeds. Generally, a small volume of an electrolyte solution (of similar viscosity and density as the bulk fluid; the electrolyte should not influence the rheological properties of the bulk fluid either) is added to the reactor fluid and then the variation of conductivity within the vessel is measured. When there is a uniform reading between the conductivities measured, then mixing is complete. Either a conductivity probe or two similar cells can be used as the measuring instrument. The main disadvantage of this method is that eventually after each experiment the entire vessel contents have to be renewed. Nevertheless, this method has the advantage of giving objective results and can give an exact criterion for mixing (Ford et al., 1972).

2) The thermal method
This technique uses a change of temperature as the property to be measured. The method involves the injection of a small amount of heated fluid (or the generation of a heat pulse by an electrical heating element) into the vessel and then recording the fluctuations in temperature at one or more points in the vessel. Complete mixing is obtained when the temperature distribution throughout the vessel is uniform. The quantity of heat should be rapidly added and must be greater than that transferred to the fluid as a result of the viscous dissipation of energy (Ford et al., 1972). There are several practical requirements which must be met before using the thermal method: A high sensitivity temperature measuring probe (thermocouple) is required if the thermal pulse is not to be excessive and the probe must have a high thermal conductivity. The heat transfer to and from the probe should occur only by transfer between the probe and the surrounding liquid. The advantage of this method is that an unlimited number of experimental runs can be carried out with the same
Another benefit is that large scale work requiring reproducible quantitative data could be conveniently performed by this method (Ford et al., 1972). Hoogendoorn and Hartog (1967) demonstrated that the mixing time can be determined for a required degree of homogeneity with this technique.

3) The colour (dye) addition method
The colour addition technique involves the observation of the contents of the vessel upon the addition of a dye to the bulk fluid. Usually, the dye solution is prepared with some of the bulk fluid and has a viscosity and density similar to the bulk fluid. The mixing time is given by the time required for a uniform distribution of the dye throughout the vessel. Several methods are available to determine when the dye is uniformly dispersed. Visual observation is generally the simplest one. Another approach is the removal of samples from the vessel and then these samples can be analysed chemically, or colourometrically, or optically (Ford et al., 1972). An advantage of this method is that it provides for a simple procedure in which the initial stages of mixing may be closely observed. However, the disadvantage is that the final stages of mixing may be more difficult to observe and the method becomes very subjective. Another disadvantage is that the addition of colour requires the removal of the bulk liquid after each experiment. A further disadvantage is that the removal of samples for analysis can affect the mixing efficiency and disturb the flow patterns (Ford et al., 1972).

4) The decolouration (chemical reaction) method
This method is based on the principle that the bulk fluid either loses or changes its initial colour due to a chemical reaction with the injected fluid. For instance, the acid-base neutralisation reaction in the presence of phenolphthalein produces a solution colour change from red to colourless. The redox reaction between iodine and sodium thiosulphate, in the presence of starch indicator, causes a change in colour from dark purple to colourless. The experiment is started with a coloured batch of liquid, which gradually clears as mixing proceeds, stagnant zones appearing as coloured patches while the bulk is already clear. The main advantage of this method is that the final mixing stages may be closely observed. This technique is suitable for small scale work (Ford et al., 1972). A disadvantage of this technique is that it is subjective. Also, for opaque fluids difficulty may arise in the visual observation of the end point. Finally, the initial reactant must be homogeneously mixed with the bulk fluid; more uniformly than can be done with the agitator to be tested, because otherwise possible stagnant zones are not coloured and will
remain undetected in the subsequent decolourisation stage (Hoogendoorn and Hartog, 1967).

5) The tracer method
The tracer technique involves the observation of a tracer component within the bulk fluid. A fluid is completely mixed when there is a uniform composition of tracer component throughout the vessel. Different tracer methods, such as electrolytic and radioactive techniques, have been developed (Ford et al., 1972). This method has no obvious advantage over other techniques since either samples must be removed or only a small volume of fluid monitored.

The mixing characteristics of airlift contactors can also be reported in terms of liquid phase axial dispersion coefficients and/or Bodenstein numbers (Siegel et al., 1988). The longitudinal mixing in an airlift reactor can be mathematically formulated by Taylor’s (1953) axial diffusion flow model. This one dimensional axial dispersion model is a suitable model on which to estimate the axial dispersion coefficient (D) in an airlift loop reactor (Blenke, 1979; Chisti, 1989; Fields and Slater, 1983; Li et al., 1993; Mavros, 1992; Murakami et al., 1982; Obradovic et al., 1994; Scragg, 1991). This type of model has found wide use in representing small deviations from plug flow (Levenspiel and Fitzgerald, 1983). Such a model assumes plug flow with disturbances caused by molecular diffusion, small eddies, dead zones and the liquid velocity profile lumped in an axial dispersion coefficient. The dispersion, in this model, is represented by a dimensionless Bodenstein number (Bo). Some studies (Chisti, 1989; Mavros, 1992; Siegel et al., 1988; Verlaan et al., 1986) have used the Peclet number (Pe) to depict dispersion.

Some researchers (Choi, 1990; Merchuk, 1986a; Siegel et al., 1988) have inferred a difference between the Bodenstein number and the Peclet number. For the calculation of the Bo number they have employed a characteristic length while for Pe number estimation the diameter of the column was used. Siegel et al. (1988) gave the Bo number as in Eqn 2.16 and the Pe number was defined as \((V_l d_c)/D\). It was suggested by Choi (1990) that the Peclet number was proportional to the Bodenstein number when the \(L/d_c\) ratio remained constant.

A few investigators (Haque et al., 1986; Pandit and Joshi, 1983) have employed the ‘tank in series with interstage recirculation’ model instead of the axial dispersion model to study mixing. Chisti (1989) noted that in some cases the ‘tank in series with interstage recirculation’ model may better describe the fluid mixing.
A mass balance over a liquid volume part in the reactor, neglecting radial concentration gradients, gives (Fields and Slater, 1983; Lu et al., 1994; Scragg, 1991)

\[
\frac{\partial \hat{C}}{\partial \theta} = \frac{1}{Bo} \frac{\partial^2 \hat{C}}{\partial x^2} - \frac{\partial \hat{C}}{\partial x} \tag{2.15}
\]

where \( \hat{C} \) is the dimensionless concentration (= \( C/C_\infty \)), \( \theta \) is the dimensionless time (= \( t/t_c \)), \( x \) is the dimensionless axial coordinate (= \( z/L \)) and \( Bo \) is the dimensionless mixing parameter.

In this equation the dimensionless mixing parameter (Bodenstein number) is the ratio of convective to diffusive transport rates. It is defined as (Siegel et al., 1988)

\[
Bo = \frac{V_1 L}{D} \tag{2.16}
\]

where \( V_1 \) is the mean liquid circulation velocity, \( D \) is the dispersion coefficient and \( L \) is the loop length of the airlift (or the length of interest). The value of the Bodenstein number represents the degree of axial mixing. The solution of Eqn 2.15, for an initial Dirac pulse in an airlift reactor taking into account the recirculating flow, is (Levenspiel, 1972)

\[
\frac{C}{C_\infty} = \left( \frac{Bo}{4\pi \theta} \right)^{\frac{1}{2}} \sum_{r=1}^\infty \exp \left[ -\frac{Bo}{4\theta} (x - \theta)^2 \right] \tag{2.17}
\]

where \( C \) is the concentration at distance \( j = yL \) in the airlift loop and \( C_\infty \) is the concentration at infinite time in the airlift loop.

The overall Bodenstein number for the reactor may be obtained by fitting the above equation (2.17) to the experimental tracer response curve with Bodenstein number as the fitting parameter (Blenke, 1979; Fields and Slater, 1983; Scragg, 1991; Voncken et al., 1964). This pulse response technique has been commonly employed in the literature (Fields and Slater, 1983; Hatch, 1973; Voncken et al., 1964). However, since it is not possible to create an ideal Dirac pulse, an experimental error is already included when using this technique (Verlaan et al., 1989).

It must be remembered that \( y \) in Eqn 2.17 must be a whole number and that any value of \( y \) cannot be used (i.e. \( j = L, 2L, 3L \ldots yL \), where tracer pulse is at \( y = 0 \)). Slater and Merchuk (1985) have warned that only multiples of the circulation path are admitted. Although it is physically possible to insert any value for \( y \), since the sensing element can be located anywhere in the loop, there is one grave restriction to the validity of Eqn 2.17 in this case. This limitation is that the axial dispersion coefficient must be the same in every part of the reactor, so that the mixing taking place in the path between the injection point and the detector is proportional to the fraction of the total recirculation length that such
path represents (Slater and Merchuk, 1985). This restriction does not exist when the sensor is located exactly at the injection point (i.e. \( y = 0 \)). Since different zones (riser, downcomer, gas-liquid separator) of an airlift reactor are hydrodynamically different, the mixing in them are expected to be different (Chisti, 1989). Hence, the use of Eqn 2.17 is incorrect unless \( i = L \).

There are several other techniques, which have been employed to analyse the tracer response data. Methods such as the moment (Hatch, 1973), weighted moment (Anderssen and White, 1977), Laplace and Fourier transform domain analysis (Gangwal et al., 1971) and time domain analysis (Lu et al., 1994; Mills and Dudukovic, 1989; Verlaan et al., 1989) have been used. A general mathematical model for the description of the mixing behaviour of airlift loop reactors has been presented by Warnecke et al. (1985; 1985a). Mazer (1994) also presented a model to calculate tracer concentrations in an airlift loop reactor, while Schmidt et al. (1992) have developed a physical model to describe the local mixing behaviour within the riser section (two phase flow) of an airlift loop reactor. Most of these methods are complex to use compared to the relatively simple axial dispersion model.

Most of the existing studies have attempted to characterize the liquid phase mixing in terms of an overall axial dispersion coefficient \( (D) \) for the reactor as a whole (Blenke, 1979; Lin et al., 1976; Takao et al., 1982; Voncken et al., 1964). A major disadvantage of these studies is that they have grouped together the, otherwise different, mixing effects of the riser, the downcomer and the head-space section of the reactor (Chisti, 1989; Fields and Slater, 1983). The Bodenstein numbers so obtained are thus empirical parameters which are not related to flow properties (Fields and Slater, 1983). In general, the flow in the riser and the downcomer is closer to a plug flow regime and flow in the head-space for gas disengagement frequently approaches a more backmixed state (Chisti, 1989).

The effect of superficial gas velocity on Bodenstein number is controversial. For example, Verlaan et al. (1989) found that the Bodenstein number increases with an increase in superficial gas velocity in all zones of the external loop airlift, while Fields and Slater (1983) reported the opposite effect for the overall Bodenstein number in internal loop airlift reactor. Chisti (1989) stated that the Peclet number for the reactor as a whole increases with increasing gas flow rate and that the same trend is displayed to varying degrees by Peclet numbers for the individual zones of the reactor. However, Frohlich et al. (1991) found that Bodenstein number decreased with increasing gas velocity in tap water and ethanol solution. Obradovic et al. (1994) reported that, for a draft tube internal loop
airlift (gas-liquid-solid), the riser Bo number decreased with superficial gas velocity, while the downcomer Bo number was not significantly affected by the increase in the superficial gas velocity. Furthermore, they stated that the gas-liquid separator Bo number decreased with increasing gas velocity. Lu et al. (1994) studied the mixing of two and three phase draft tube internal loop airlift reactors. They found the overall and the downcomer Bo numbers to decrease with increasing gas flow rate. However, the increase in gas flow rate apparently had no effect on the riser, gas-liquid separator and the bottom section Bo numbers.

For Newtonian fluids, Chisti (1989) reported that the Peclet (Pe) numbers in the individual sections of an airlift were found to be in an order of Pe(downcomer)>Pe(riser)>Pe(separator). In water-like fluids, the usual range of Pe(overall) in internal and external loop airlift devices is 30-80 (Chisti, 1989). Chisti (1989) characterised the fully mixed state of continuous flow reactors as when Pe < 0.1 and characterised plug flow behaviour as when Pe > 20. Verlaan et al. (1989) showed that the Bodenstein (Bo) number for the riser of an external loop airlift reactor was between 30-40, while the Bo numbers in the downcomer and gas-liquid separator were 40-50 and 10, respectively. Obradovic et al. (1994) reported their Bo numbers for the riser and the downcomer sections to be between 5-50 and 20-60, respectively. Their gas-liquid separator approached ideal mixing and had a Bo number between 0.2-0.6. Apparently, their reactor as a whole also had a high degree of mixing with Bo numbers ranging from 0.3-0.9. Lu et al. (1994) found the Bo numbers in the riser to be between 20-30 and 40-70 in the downcomer. They also discovered the gas-liquid separator Bo number to be 10 and the bottom section Bo number to be between 10-20.

The effect of gas flow rate on the liquid phase axial dispersion coefficient in airlift reactors is also not clear. A number of studies (Li et al., 1993; Lu et al., 1994, 1994a; Obradovic et al., 1994; Verlaan et al., 1989) have observed the local or overall axial dispersion coefficient to increase with increasing gas velocity. However, Wu and Jong (1994) noticed their overall axial dispersion coefficient in a draft tube internal loop airlift reactor to increase to a maximum with increasing gas velocity and then decrease for further increases in gas velocity. They stated that this phenomenon was due to the Sauter mean diameter of bubbles. Very small bubble sizes, at low gas flow rates, led to violent eddies and therefore the axial dispersion coefficient increased with increasing gas flow rate in this region. But, at higher air velocities the dispersion coefficient values decreased due to the prevalence of bubbly flow.
2.3.2 Mass Transfer

In most applications, airlift reactors are employed in aerobic processes utilizing microorganism growth (Siegel et al., 1988). Because of the low solubility of oxygen in the liquid phase of these processes, fresh oxygen must be continuously transferred from the gas to the liquid phase. Hence, for these systems to be both physically and economically viable, a maximum rate of oxygen transfer must be achieved while minimising the power requirements. Thus, for biological processes, gas-liquid mass transfer is one of the most important design considerations in airlift reactor design.

Mass transfer, in airlift contactors, is directly influenced by the overall volumetric mass transfer coefficient \( k_{ba} \) and the liquid mixing (Chisti and Moo-Young, 1988). It is affected indirectly by the gas holdup and liquid velocity in the airlift vessel. Mass transfer is also dependent on the gas flow rate, reactor geometry, liquid properties and gas recirculation.

**Overall volumetric mass transfer coefficient**

The rate of mass transfer of a gas such as oxygen or carbon dioxide from gas to liquid phase (or vice versa) can be expressed in terms of a liquid-side mass transfer coefficient \( k_L \), the specific gas-liquid interfacial area \( a \), through which the transfer occurs, and a concentration driving force for the transfer. Mass transfer performance of airlift reactors has frequently been stated in terms of an overall volumetric mass transfer coefficient \( k_{ba} \) based on the unaerated liquid volume (Chisti and Moo-Young, 1987). The volumetric mass transfer coefficient is the rate of gas transfer across the gas-liquid interface, per unit volume of suspension and per unit of driving force (Merchuk, 1990). It is given by (Siegel et al., 1988):

\[
\frac{dC_L}{dt} = k_L a \left(C^* - C_L\right)
\]

where \( C_L \) is the instantaneous concentration of dissolved oxygen in the liquid phase and \( C^* \) is the maximum possible oxygen concentration (i.e. saturation value) in the liquid. Due to the low solubility of oxygen in the liquid phase, the liquid-film resistance at the gas-liquid interface is considered the limiting step in the mass transfer process (Siegel et al., 1988).

The overall mass transfer coefficient is influenced by the superficial gas velocity, reactor geometry, liquid velocity, liquid-phase properties, gas recirculation and bubble coalescence (Muller and Davidson, 1992; Shah et al., 1982; Siegel et al., 1988).

A number of techniques exist for the measurement of overall volumetric transfer coefficient. They are (Chisti, 1989; Sobotka et al., 1982): (1) sulphite oxidation method,
(2) carbon dioxide absorption technique, (3) hydrazine method, (4) pressure gauge procedure and (5) enzymatic oxidation method.

Generally, in airlift reactors, the liquid-side mass transfer coefficient \( (K_{\text{L}}) \) is considered as a constant and is independent of the hydrodynamics (Seipenbusch et al., 1980).

**Interfacial Area**

The gas-liquid interfacial area is an important design variable which depends on the geometry of the device, the operating conditions and the physical properties of the liquid media. The mass transfer rate, which is determined by the volumetric mass transfer coefficient \( (k_{La}) \), can be easily influenced by varying interfacial area \( (a) \). The interfacial area is defined by (Chisti and Moo-Young, 1987):

\[
a = \frac{6\varepsilon_o}{d_{\text{vs}}} \left(1 - \varepsilon_o\right)
\]

where \( d_{\text{vs}} \) is the volume-to-surface mean bubble diameter, often called the Sauter mean diameter.

Numerous techniques have been used to measure the interfacial area in gas-liquid dispersions. The methods fall into two main categories: (1) physical methods and (2) chemical methods. The physical methods involve determining the bubble size. Many methods are available to determine bubble sizes. Photographic techniques are used widely because of their simplicity (Akita and Yoshida, 1974; Burckhart and Calderbank, 1975; Popovic and Robinson, 1987). Other methods involve light scattering, light reflection (Calderbank et al., 1960) and depolarization. Though the original bubble size distributions obtained from various techniques differ markedly, the Sauter mean bubble diameter \( (d_{\text{vs}}) \), evaluated consequently differ only slightly. This is confirmed by Yamashita et al. (1979), who have reported that the photographic results are in reasonable agreement with those obtained from the two-point electrical probe method. To determine interfacial area, by a chemical method, a suitable reaction is applied. The widely applied reactions are sulphite oxidation (Popovic and Robinson 1987; 1987a) and carbon dioxide absorption in alkali (Prasher and Wills, 1973). Other suitable methods have been proposed in the literature, including the reactions in organic solvents (Bossier et al., 1973; Sharma and Danckwerts, 1970; Sridharan and Sharma 1976).

Interfacial area determined from the physical and chemical methods may differ by more than 100% (Kurten and Zehner, 1978; Voyer and Miller, 1968). It is extremely
difficult to measure the interfacial area accurately, hence usually only the overall volumetric mass transfer coefficient is experimentally measured (Chisti, 1989).

A number of investigators have studied the mass and oxygen transfer in airlift reactors (Ayazi Shamlou et al., 1995; Bello et al., 1985a; Chisti, 1989; Chisti and Moo-Young, 1988b; Ghosh et al., 1993; Halard et al., 1989; Kawase and Moo-Young, 1986a, 1986b, 1987; Kawase and Tsujimura, 1994; Koide et al., 1983, 1984; Lindert et al., 1992; Margaritis and Sheppard, 1981; Onken and Weiland, 1980; Pollard, 1995; Russell, 1989; Siegel et al., 1988; Zhao et al., 1994; Zwietering et al., 1992). Most of these studies have demonstrated that the airlift vessel provides high mass transfer at low power consumption.

2.3.3 Heat Transfer

Fluid-wall heat transfer is an important parameter in airlift reactor design and performance (Chakravarty et al., 1971; Siegel et al., 1988). However, it is one of the least studied phenomenon in airlift contactors (Chisti and Moo-Young, 1987; Siegel et al., 1988). Compared to airlift reactors, heat transfer in bubble columns has been the subject of many investigations (Deckwer, 1980, 1992; Fair et al., 1962; Hart, 1976; Heijnen and Van’t Riet, 1984; Joshi et al., 1980; Nishikawa et al., 1977; Shah et al., 1982). Siegel et al. (1988) noted that most of the existing knowledge on heat transfer in airlift reactors has been extrapolated from work done on bubble columns.

Generally, in the chemical industry, exothermic gas-liquid reactions carried out in pneumatic contactors necessitate means for heat removal. Other operations such as gas absorption accompanied by chemical reaction may also create heat dissipation problems.

In biotechnology, the growth rates of most microorganisms are temperature dependent with growth rate and productivity falling sharply if the temperature shifts out of the optimal range for the organism (Siegel et al., 1988). Most commercial fermentation processes are both mesophilic and exothermic (Siegel et al., 1988). The problem of heat transfer is further compounded by the fact that fermentations which utilise highly reduced carbon sources such as hydrocarbons and methanol generate much larger quantities of heat (Chisti and Moo-Young, 1987).

Heat transfer in airlift reactors is directly influenced by the liquid mixing and the heat transfer coefficient in the reactor (Chisti and Moo-Young, 1988). Gas holdup and liquid velocity in the airlift indirectly affects the heat transfer (Chisti, 1989). Heat transfer will also be affected by parameters that influence bubble rise velocities and liquid element flow paths such as the gas flow rate and the liquid phase properties (Siegel et al., 1988).
2.4 Hydrodynamic Correlations for Airlift Devices

In this section, we examine the existing hydrodynamic models and correlations for the prediction of gas holdup and/or liquid velocity. Specific emphasis is placed upon the studies which consider the liquid viscosity effects on gas holdup and/or liquid velocity.

Various researchers have derived empirical relationships for gas holdup and/or liquid velocity, usually based on the superficial gas velocity (Bello et al., 1984, 1985, 1985a; Chisti et al., 1987; Chisti and Moo-Young, 1988a; Hills, 1976; Kawagoe and Robinson, 1980; Miyahara et al., 1986; Onken and Weiland, 1980; Philip et al., 1990; Popovic and Robinson, 1984, 1988, 1989; Siegel et al., 1986; Vatai and Tekic, 1986). These correlations are often relatively simple and system specific (generally restricted to the same reactor size, type and gas-liquid system); hence usually they cannot be applied to any airlift reactor geometry.

Numerous investigators have also developed models of the fluid hydrodynamics which illustrate the theoretical relationship between gas holdup, circulation pressure drop and liquid velocity in the different types of airlift devices (Blenke, 1979; Chakravarty et al., 1974; Chisti et al., 1988; de Nevers, 1968; Fan et al., 1984; Freedman and Davidson, 1969; Garcia Calvo, 1989, 1992; Glennon et al., 1993; Hsu and Dudukovic, 1980; Jones, 1985; Joshi et al., 1990; Kawase, 1990; Kemblowski et al., 1993; Koide et al., 1984a; Kubota et al., 1978; Lee et al., 1987; Merchuk et al., 1980; Merchuk and Stein, 1981, 1981a; Onken and Trager, 1990; Russell, 1989; Verlaan et al., 1986; Young et al., 1991). These models were obtained employing either an energy balance on the gas input due to the isothermal expansion of the sparged gas, momentum balances and/or a steady state mechanical energy balance (usually in conjunction with the Zuber and Findlay (1965) drift flux model). Kawase et al. (1995) have even developed a gas holdup model based on the Kolmogoroff theory of isotropic turbulence. In most of these models a key parameter is the total frictional resistance to fluid flow in the airlift loop, including fluid reversals at the top and bottom. Generally, friction coefficients ($K_f$) have been estimated by applying standard relationships for single phase flow in pipes and bends (Verlaan et al., 1986), two phase flow (Joshi et al., 1990) or by treating $K_f$ as an adjustable parameter (Chisti et al., 1988).

Most of the correlations, in the literature, for gas holdup and liquid velocity in airlift contactors which specifically consider the frictional effects and/or the liquid viscosity effects are given in Table 2.1.

Most of the models or correlations stated above only applied to particular vessels and in several instances either the predicted holdup or velocity did not satisfactorily
compare with the observed value or the equation had a large amount of empiricism. The models also employed empirical values for friction factors, gas recirculation rates and/or slip velocities. Furthermore, experimental verification of these correlations generally has been restricted to laboratory or pilot scale airlifts and to the air-water or air-Newtonian fluid systems.

For example, the relatively simple equation reported by Jones (1985) (Eqn 18 of his paper) predicted significantly higher or lower values of liquid circulation velocity than the experimentally found ones; he also confined his work to draft-tube sparged concentric-tube airlifts. Jones (1985) also proposed a second theoretical equation (Eqn 21 of his paper) in an identical form to the theoretical relationship derived by Chakravarty et al. (1974). Both equations were similar to the correlation derived by Russell (1989). All three equations were third degree polynomials in liquid velocity and were difficult to use for other reasons as well (Chisti, 1989). Chisti (1989) has compared (Fig 3.5 of his book) the Jones's (1985) cubic equation prediction with Jones's (1985) experimental measurements obtained in air-water fluids. Chisti (1989) shows that even for a simple system (air-water) Jones's (1985) experimental results did not match within ±30% of his theoretical values. In view of the shortcomings of the model, Clark and Jones (1987) have attempted to improve the theory, but the analysis remained limited to particular reactor geometries.

Bello et al. (1984) have used an energy balance on the airlift loop to theoretically relate the riser linear liquid velocity to the riser superficial gas velocity. This term involved excessive empiricism (Bello, 1981). Their correlation also ignored the viscosity of the fluid, therefore it appears that wall friction seems to be not considered.

Lee et al. (1987) presented an analysis of the liquid circulation phenomenon in airlift contactors. The energy losses in the wakes behind bubbles in the riser as well as the losses in the downcomer due to upflow of large bubbles were also taken into account. The predicted circulation velocities were found to satisfactorily agree with the experimentally observed values, but the calculation required a knowledge of terminal bubble rise velocity. The investigators used a realistic value of 0.23 m/s for the terminal bubble rise velocity in most of their calculations. In some cases, however, to achieve satisfactory agreement between the model and the experimental results, they had to adjust the bubble rise velocity and consequently the bubble velocities used varied almost two-fold over the range 0.22-0.43 m/s, with little justification. Thus the bubble rise velocity was used as merely an adjustable parameter without any real physical meaning. Furthermore, these investigators
assumed quite arbitrary values of frictional loss factors, between two and four, for the various airlift vessels despite the wide variations in reactor geometries.

Chisti et al. (1988) have also given a model, in their paper, to predict the liquid circulation rate. Their model is based on an energy balance and requires a value for friction losses in the top and bottom connecting sections of an airlift, gas-liquid dispersion height and the value of the gas holdup in the riser and the downcomer. For the calculation of the top and bottom friction loss coefficients Chisti et al. (1988) have used an empirical equation. Furthermore, their model requires the riser and downcomer gas holdups to be known a priori. This might not be possible in all cases. If the riser and the downcomer gas holdups are not known, then the model can still be used but the procedure becomes more complex.

Another model, reported by Garcia Calvo (1989), estimates both the gas holdup and the liquid circulation velocity in airlift reactors for a given gas flow rate. This model requires a value for the friction losses of fluids against the airlift device and a value for the slip velocity. Garcia Calvo (1989) has shown the model to predict liquid circulation and gas holdup in airlifts to within ±20% of the experimental values, over a broad range of reactor geometries and scales of operation. Garcia Calvo and Leton (1991) extended this model to include bubble columns. However, the model has two uncertain parameters (slip velocity and total friction coefficient) and moreover the calculation procedure is not simple.

Joshi et al. (1990), Verlaan et al. (1986) and Young et al. (1991) have proposed rational models for two-phase flows in airlift reactors, but their models are very complicated and have only been applied to air-water systems.

Recently, Glennon et al. (1993) and Kemblowski et al. (1993) have derived models for gas holdup and liquid velocity in airlifts. Both models have only been tested on external loop airlift contactors. Glennon et al. (1993) have employed air-water system only and have empirical factors in the correlation for liquid velocity. The Fanning friction factors required in the model of Kemblowski et al. (1993) are calculated using empirical relationships.

In summary, it can be stated that numerous researchers have developed theoretical relationships to estimate the gas holdup and/or liquid velocity in airlift reactors. But a non-empirical model which can be applied to all the various geometries of airlifts and to different fluid systems has not yet been firmly established.
<table>
<thead>
<tr>
<th>Reference</th>
<th>Proposed correlations</th>
<th>System studied</th>
<th>Parameter range</th>
<th>Reactor type</th>
</tr>
</thead>
<tbody>
<tr>
<td>Chakravarty et al. (1973; 1974)</td>
<td>[ \varepsilon_g = \left( \mu_L - \mu_w \right)^{2.75} + 161 \left( \frac{73.3 - \sigma_L}{79.3 - \sigma_L} \right) \times 10^{-4} \times U_{rg}^{0.88} ] &lt;br&gt; [ aV_r^2 + bV_r^2 + cV_r + d = 0 ] where &lt;br&gt; [ a = \left( K_o + PL_d \right) + \frac{\varepsilon_g}{\varepsilon_d} ] &lt;br&gt; [ b = \frac{U_{rg}}{\varepsilon_g} - \frac{U_w A_d \varepsilon_d}{A_r \varepsilon_r} ] &lt;br&gt; [ c = 2gL_d \frac{\varepsilon_g - \varepsilon_d}{1 - \varepsilon_r} ] &lt;br&gt; [ d = \frac{2gL_d}{1 - \varepsilon_g} \left[ \frac{2U_w A_d \varepsilon_d}{A_r} - U_w \left( 1 - \varepsilon_d \right) \right] ] &lt;br&gt; [ K_o = K_1 \left( 1 + \frac{A_r^2}{A_d^2} \right) + K_2 ] and &lt;br&gt; [ P = 4 \left[ \frac{f_1}{T} + \frac{f_2 A_r^2}{d_d A_d^2} \right] ]</td>
<td>Gas - air; Liquids - water, sodium sulphate, isobutyl alcohol (1 and 2% v/v) and glycerol (47.7 and 52.7% w/w) solutions</td>
<td>[ U_{rg} = 0.015-0.24 \text{ m/s}; ] Reactor capacity = 0.038 m³; [ d_c = 0.10 \text{ m}; ] [ T = 0.045-0.074 \text{ m}; ] [ H_t = 0.73 \text{ m}; ] [ H = 0.486 \text{ m}. ]</td>
<td>Concentric draft tube internal loop (draft tube sparged)</td>
</tr>
<tr>
<td>Chisti et al. (1988)</td>
<td>[ U_{sl} = \left( \frac{2 g H (\varepsilon_g - \varepsilon_d)}{K_t + K_b \left( \frac{A_r}{A_d} \right)^2 \frac{1}{1 - \varepsilon_g}} \right)^{0.5} ]</td>
<td>[ \varepsilon_d = 0.89 \varepsilon_g ]</td>
<td>K_t and K_b are the frictional loss coefficients for the top and bottom sections, respectively. ( \varepsilon_g ) has to be either measured or estimated using an empirical correlation.</td>
<td></td>
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<tr>
<td>-------------------</td>
<td>-------------------------------------------------</td>
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</tr>
<tr>
<td>Garcia Calvo (1989)</td>
<td>[ \varepsilon_g = \frac{U_{sg}}{V_{lr} + V_r} ]</td>
<td>[ V_{ld} = \left( \frac{P_h U_{sg} \ln \left( \frac{P_h + \rho_l g H}{P_h} \right) - \varepsilon_g V_r \rho_l g H}{0.5 K_f \rho_l \left( \frac{A_d}{A_r} \right)} \right) ]</td>
<td>( V_r = 0.25 \text{m/s} )</td>
<td></td>
</tr>
<tr>
<td>Glennon et al. (1993)</td>
<td>[ \varepsilon_g = b U_{sl}^2 ]</td>
<td></td>
<td>Gas - air; Liquid - water</td>
<td></td>
</tr>
</tbody>
</table>
\[ U_{sl} = 1.017 \cdot b^{-0.409} U_{sg}^{0.420} ; \quad U_{sg} \leq 0.05 \text{ m/s} \]

\[ U_{sl} = 0.735 \cdot b^{-0.427} U_{sg}^{0.115} ; \quad U_{sg} > 0.05 \text{ m/s} \]

where

\[ b = \frac{\left(\sum K_f\right) A_r^2}{2 g H_D A_d^2} \]

This empirical correlation is simple to apply, but requires the accurate determination of the loss coefficients for the reactor.

| Hsu and Dudukovic (1980) | \( H_d - (1 - \varepsilon_o) H_r = \left( h_{lu} + \frac{4 f_{lu} L_{er}}{T} \right) \frac{U_{sl}^2}{2 g} (1 - \varepsilon_o) + \left( h_{ld} + \frac{4 f_{ld} L_{ed}}{d_d} \right) \frac{V_i^2}{2 g d_d} T^4 \) | Gas - air; Liquid - water, other liquids (not mentioned) with physical properties
\( \rho_L = 800-1500 \text{ kg/m}^3 \)
\( \mu_L = 0.0008-0.056 \text{ Pa s} \) | U_{sg} = 0.015 - 1.0 m/s; T = 0.02-0.045 m; H/T = 38 - 90 | External loop

| Joshi et al. (1990) | \[ \frac{dV_i}{dy} = \frac{\tau_w}{\mu_L} \left[ \eta - \frac{d_c \varepsilon_o \rho_L g \eta}{2 \tau_w} x \left(1 - \eta^2\right) \right] \] | Gas - air; Liquid - water | Reactor volume = 10 - 1000 m\(^3\); T = 0.255 - 2.492 m; d_d = 0.255 - 2.332 m; H_t = 5.40 - 102.45 | External loop

\[ T = 0.1-0.225 \text{ m}; d_d = 0.05-0.15 \text{ m}; A_d/A_r = 0.198-1; H_D = 2-10.5 \text{ m}; \sum K_f = 1.9-22.6 \]
\( \tau_0 \) is required. \( \varepsilon_0 \) has to be either measured or estimated using an existing correlation. Then for a given boundary condition the model is solved, by applying a 6-step iterative procedure, to obtain \( U_{gl} \).

| Kawase and Moo-Young (1986a) | Gas - air; Liquids - water and solutions of glycerine and CMC, silicone antifoam emulsion | m; 
\( A_d/A_r = 0.25 - 4.0 \) | Bubble column; draft tube internal loop |
|---|---|---|---|
| \( \varepsilon_0 = 0.24 n^{-0.6} \left[ \frac{U_{sp0}}{\sqrt{gd_c}} \right]^{0.84 - 0.14n} \left[ \frac{gd_c}{v_0^2} \right]^{0.07} \) | \( U_{slk} = 0.008-0.285 \) m/s; 
\( d_c = 0.14-0.35 \) m; 
\( T = 0.113 \) m; 
\( A_d/A_r = 0.33 \); 
\( n = 0.28-1.0 \); 
\( K = 0.00089-1.22 \) Pa s^n | |
| This is a relatively simple empirical correlation. | | |

| Kemblowski et al. (1993) | Gas - air; Liquids - water without and with added surfactants; solutions of glycol, sugar syrup and carboxymethyl cellulose | Reactor volume = 0.08-0.7 m³; 
\( T = 0.1-0.2 \) m; 
\( d_d = 0.1-0.2 \) m; 
\( H_l = 1.95-7.18 \) m | External loop |
<table>
<thead>
<tr>
<th></th>
<th></th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>( \varepsilon_g = 0.203 \frac{Fr^{0.31}}{Mo^{0.012}} \left( \frac{U_{g0} A_r}{U_{d0} A_d} \right)^{0.74} )</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>( U_d = \left( \frac{2 g H_D (\varepsilon_g - \varepsilon_d)}{K_t \left( \frac{1 - \varepsilon_g}{A_r} \right)^2 + K_b \left( \frac{A_r}{A_d} \right)^2 + 4 H_D \left( \frac{f_r + f_d}{T} \frac{A_r}{A_d} \right)^2} \right) )</td>
<td></td>
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<td></td>
</tr>
</tbody>
</table>
| The Fanning friction factors (\( f \)) are: for laminar flow (Re<1000) 
\( f = \frac{16}{Re} \) for turbulent flow (Re>2000) | | | |
\[ f = \frac{0.0792}{Re^{0.25}} \]

Fr and F_d are the friction factors for two-phase flow inside the riser and downcomer, respectively.
An iterative procedure is followed to predict \( \varepsilon_g \) and \( U_{sl} \) values.

### Lee et al. (1987)

\[

E_g = A_d \rho_d g H \varepsilon_g U_{sl} + \sum_{j=\text{riser}} 0.5 \rho_j V_j K_f A_j (1 - \varepsilon_j) + A_d \rho_d g H \varepsilon_d V_d
\]

and

\[
V_1 A_r (1 - \varepsilon_r) = V_{id} A_d (1 - \varepsilon_d)
\]

\( V_1 \) and \( V_{id} \) can be obtained from the two equations above; providing \( \varepsilon_g \) and \( \varepsilon_d \) are known (either estimated or measured from experiment).

### Merchuk and Stein (1981)

\[
\varepsilon_g = \frac{2U_s}{gT} \left[ \left(1 + \frac{U_s}{U_d}\right) K_{fm} + K_f \left(1 + \frac{L_e}{H}\right) \left(\frac{A_r}{A_d}\right)^{2.5}\right]
\]

The above equation is used to calculate \( U_{sl} \), for a given value of \( U_{sg} \) and \( \varepsilon_g \).

\[
K_{fm} = 0.046 N_{Re}^{-0.2}
\]

where

\[
N_{Re} = \frac{\rho_s U_s d_e}{(1 - \varepsilon_g) \mu_L}
\]

\[
K_f = 0.0791 Re^{-0.25}
\]

where

<table>
<thead>
<tr>
<th>Component</th>
<th>Details</th>
</tr>
</thead>
<tbody>
<tr>
<td>Reactor volume</td>
<td>Gas - air; Liquid - water; salt water (0.6% NaCl); fermentation medium</td>
</tr>
<tr>
<td>Total reactor volume</td>
<td>Gas - air; Liquid - water</td>
</tr>
<tr>
<td>Draft tube and split cylinder</td>
<td>Draft tube and split cylinder internal loop; external loop</td>
</tr>
<tr>
<td>Literature Survey</td>
<td>Internal loop; external loop</td>
</tr>
</tbody>
</table>
\[
\text{Re} = \frac{\rho_l U_{sl} d_e}{\mu_l}
\]

K\text{fm} and K\text{fl} are the friction coefficients for the mixture and the liquid, respectively. N\text{Re} is the mixture Reynolds number. The riser holdup is either measured or calculated using an empirical correlation.

<table>
<thead>
<tr>
<th>Authors</th>
<th>Formula</th>
<th>Conditions</th>
<th>Reactor volume</th>
<th>External loop</th>
</tr>
</thead>
<tbody>
<tr>
<td>Onken and Trager (1990)</td>
<td>[\varepsilon_g = aU_{ig}^b]</td>
<td>Gas - air; Liquids - water; solutions of sodium chloride (0.4M), i-propanol (0.22% w/w), glycerol (45.7-62.0% w/w) and carboxymethyl cellulose (0.25% w/w)</td>
<td>Reactor volume = 0.056-3.1 m(^3); T = 0.1-0.6 m; d(_d) = 0.05-0.2 m; H(_t) = 4.4-9.5 m</td>
<td>External loop</td>
</tr>
<tr>
<td>Popovic and Robinson (1984)</td>
<td>[\varepsilon_g = 0.02 U_{ig}^{0.6504} \left(1 + \frac{A_d}{A_r}\right)^{-1.0516} \mu_a^{-0.1039}; \mu_a \text{ based on shear rates calculated using } \gamma = 50 U_{sg} (\text{Nishikawa et al., 1977}); U_{sg} \text{ is in cm/s.} ]</td>
<td>Gas - air; Liquid - Carboxymethyl cellulose solutions</td>
<td>T = 0.152 m; d(_d) = 0.05 - 0.1 m; A(_d)/A(_r) = 0 - 0.44</td>
<td>External loop</td>
</tr>
<tr>
<td>Vatai and Tekic (1987)</td>
<td>[\varepsilon_g = 0.128 \left(\frac{U_{sg}}{U_{sl}}\right) \text{Re}^{0.062} \text{We}^{0.164}; ]</td>
<td>Gas - air; Liquids - water, carboxymethyl cellulose solutions</td>
<td>T = 0.10 m; d(_d) = 0.10 m; A(_n)/A(_d) = 0-1.0; n = 0.805-1;</td>
<td>External loop</td>
</tr>
</tbody>
</table>
\[ U_{st} = 2.858 \left( \frac{A_n}{A_d} \right)^{0.416} U_{ng}^{0.481} \mu_a^{0.0105} ; \]

where
\[ \mu_a = KY^n \quad ; \quad Y = 5000U_{ng} \quad \text{and} \]
\[ A_n \quad \text{is the cross-sectional area of the orifice plate opening (m}^2). \]

Verlaan et al. (1986)

\[ U_{st} = \left[ \frac{2g}{K_f} \left( \int_{e}^{l} \varepsilon_{g}(z)dz - \int_{0}^{l} \varepsilon_{r}(z)dz \right) \right]^{\frac{1}{2}} ; \]

\[ \varepsilon_{g}(z) = \frac{U_{ng}(z)}{\left[ C_o \left( U_{st}(z) + U_{nr}(z) \right) + U_{nr} \right]} ; \]

\[ U_{ng}(z) = \frac{U_{ng}(0)P_b}{P(z)} ; \]

\[ P(z) = \left[ P_b - \rho_L g z - P_b e_T(0) \right] \ln \left( \frac{1 - \rho_L g z}{P_b} \right) ; \]

\[ U_{ng}(z) = \frac{Q_g}{A_r} ; \]

\[ Q_{mr} = \frac{Q_{mi}}{(1-q)} ; \quad Q_{md} = \frac{Q_{mi} q}{(1-q)} \]

Gas - air; Liquid - water

Reactors volume = 0.165-0.6 m³; Ht = 3.23-10.5 m; T = 0.2-0.225 m; \( d_d = 0.1 \) m; \( K_f = 0.1-11.36 \)

External loop
Where

\[ P(z) \] is the pressure at a distance of \( z \) (Pa); \( q \) is a constant; \( Q_{md} \) is the mass flow rate of gas in the downcomer (m\(^3\)s\(^{-1}\)); \( Q_{mi} \) is the mass flow rate of injected gas in the sparger (m\(^3\)s\(^{-1}\)); \( Q_{mr} \) is the mass flow rate of gas in the riser (m\(^3\)s\(^{-1}\)); \( z \) is a coordinate.

This model requires an interactive calculation procedure to acquire the riser and downcomer gas holdups and the riser liquid velocity.

<table>
<thead>
<tr>
<th>Young et al. (1991)</th>
<th>Gas - air; Liquid - water</th>
<th>Reactor working volume = 160 L; T = 0.19 m; ( d_d = 0.089 ) and 0.14 m; ( U_{sg} = 0.0096 - 0.082 ) m/s; ( \frac{A_d}{A_r} = 0.47 ) and 0.73</th>
<th>External loop</th>
</tr>
</thead>
<tbody>
<tr>
<td>[ \frac{d}{dz} \left[ (\rho_L)(1 - \epsilon_p)(U_{sg}) \right] = 0 ]</td>
<td></td>
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<td></td>
</tr>
<tr>
<td>[ \frac{d}{dz} \left[ (1 - \rho_L)\epsilon_p (U_{sg}) \right] = 0 ]</td>
<td></td>
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<td></td>
</tr>
<tr>
<td>[ -\frac{d}{dz} \left[ (P_h)(1 - \epsilon_p) \right] - \epsilon_p (\rho_L)(g) - \frac{1}{T} \frac{1}{2} \rho_L (V_i)^2 + ]</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>[ F_{dfc} \frac{3\epsilon_p}{R_b} = 0 ]</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>[ -\frac{d}{dz} \left( \epsilon_p (P_h) \right) - F_{dfc} \left( \frac{3\epsilon_p}{R_b} \right) = 0 ]</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>[ \frac{V_{dfc}^2}{2g} \left[ \frac{3}{2} \left( \frac{d_d}{(1 - \epsilon_p)} \right)^2 \left[ 1 - \left( \frac{T}{d_d} \right)^2 \right] \left( \frac{d_d}{T} \right)^2 - 1 \right] + ]</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>[ \frac{\langle P_h \rangle - P_{bd}}{\rho_L g} = 0 ]</td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>
\[
\frac{V_{lbh}^2}{2g} \left[ \frac{(d_d)}{T}^2 - 1 \right] + \frac{P_{hr} - P_{hd}}{\rho_l g} - l_{sr} + f_d \left( \frac{L_{slid}}{d_d} \right) \frac{V_{lbh}^2}{2g} + \\
\frac{V_{lbh}^2}{2g} \left\{ 1.3 + \left[ 1 - \frac{(d_d)}{T} \right]^2 \right\} + 15 \frac{(V_{l})^2}{2g} = 0
\]

Where
\( \epsilon \) is the phase volume fraction; \( f \) is the friction factor; \( L_{sl} \) is the axial distance from sparger to riser; \( z \) is the axial component; subscripts: \( r \) and \( d \) represent the riser and the downcomer, respectively; \(< > = \text{volume average.}\)

The first four differential equations describe the riser hydrodynamics. The bottom two algebraic equations describe the flow through the separator and the downcomer. The riser equations must be integrated along the column. The resulting values at the riser entrance and exit are then substituted into algebraic equations and consistency between the two sections is checked. Newton-Raphson iteration and the Runge-Kutta technique are used to solve the above equations.

Table 2.1: Literature correlations of gas holdup and liquid velocity for airlift reactors, which consider liquid viscosity effects.
2.5 The Industrial Applications and Uses of Airlift Reactors

Airlift reactors are used commercially in the manufacture of pharmaceuticals, bulk chemicals, foods, feeds and in wastewater treatment. Continuous production of vinegar, citric acid, beer and biomass from yeast, bacteria and fungi has been carried out in airlift vessels at different working capacities (Moresi, 1981).

Eastern Europe and Russia have extensively utilized airlift devices for single cell protein (SCP) yeast cultivation (Blakebrough et al., 1967); in United Kingdom, Imperial Chemical Industries Ltd. (ICI) has operated an airlift fermenter with a working volume of 1500 m$^3$ for the PRUTEEN process (Westlake, 1986). Malfait et al. (1981) claimed a more than 18% (by weight) enhancement in yield of a filamentous mould Monascus purpureus in an external loop airlift relative to that in a stirred tank. This yield improvement was achieved with a 50% reduction in power input leading to a more than 50% reduction in the cost of biomass produced in the airlift. Improved productivity of the airlift reactor relative to the stirred vessel was associated with the higher overall mass transfer coefficient obtained in airlift (Malfait et al., 1981). Similarly, Erickson et al. (1983) quoted a report where novobiocin production using Streptomyces niveus in airlift fermenters was found to be as good or better compared to results obtained in stirred tanks; they have also given several other examples of successful cultivation of mycelial fungi in airlift reactors. Schugerl (1990) has compared the performances of stirred tank and airlift reactors for the production of secondary metabolites: penicillin V by Penicillium chrysogenum, cephalosporin C by Cephalosporium acremonium, and tetracycline by Streptomyces aureofaciens. He stated that in stirred tank reactors, cell mass concentrations, volumetric productivities and specific power inputs are higher than in the airlift reactors; but in airlifts, efficiencies of oxygen transfer are higher, and specific productivities with regard to power input, substrate and oxygen consumption, and yield coefficients of product formation with regard to substrate and oxygen consumption are considerably higher than in stirred tank reactors. Bonnarme et al. (1993) cultivated the fungi Phanerochaete chrysosporium in both airlift and stirred tank reactors. They reported that in airlift reactors, the yields of lignin and manganese peroxidases as well as extracellular protein, were considerably increased as compared with stirred tank reactors.

Fermentations of Candida lipolytica (Seipenbusch et al., 1976) and Candida intermedia (Hatch, 1973) on n-paraffin in internal loop airlifts have been illustrated and continuous cultivation of another yeast (Candida utilis) in a concentric draft-tube airlift has been shown to be possible (Huang et al., 1976). A hydrocarbon fermentation in airlift of
external loop design has also been reported (Blakebrough et al., 1967). Sousa et al. (1994) studied ethanolic fermentation in bioreactors. Wood and Thompson (1986) have used airlift reactors for hybridoma culture for monoclonal antibody production. The relatively low shear tolerances of mammalian cells and their low oxygen demands mean that very different reactor operation regimes must be used for these fermentations than is conventional for microbial processes. Trager et al. (1989) have proved the superiority of airlift reactors over stirred tank reactors for the production of gluconic acid by Aspergillus niger. Shuler et al. (1984) studied the large scale production of a wide range of plant products from plant cell tissue cultures. They reported that one of the difficulties with large scale systems was the shear sensitivity of the cells. To overcome this problem, they suggested that airlift fermenters can be used. Ganzeveld et al. (1995) investigated the hydrodynamic behaviour of animal cell suspensions in internal loop airlift bioreactors. They found that airlift bioreactors displayed suitable hydrodynamic characteristics for supporting anchorage dependent animal cell cultures on suspended microcarriers.

Very deep airlifts (deep shaft) of the internal loop type (concentric draft tube or split cylinder reactors) has successfully been employed both for municipal and industrial aerobic wastewater treatment (Bouchard et al., 1977; Bolton and Ousby, 1977; CIL publication, 1978; Cox et al., 1980; Hemming, 1979; Hines et al., 1975; Kubota et al., 1978; Leistner et al., 1979; Otewell, 1994; Redman, 1987; Siegel et al., 1994). The deep shaft process was developed by the Agricultural Division of ICI, as a spin-off during research into single cell protein. The deep shaft is a modification of the activated-sludge process. The deep shaft is essentially a very efficient oxygen-transfer device comprising a shaft 50-150 m deep, vertically partitioned into a downflow section (downcomer) and an upflow section (riser) and surmounted by a gas disengagement tank. The shaft can be 0.5 m to 10 m diameter depending on the duty. The influent is fed to the top of the deep shaft. Initially compressed air is injected into the riser at a comparatively shallow depth to initiate liquid circulation around the entire system. Once this circulation is established, the air is gradually transferred to the downcomer, until all or most of it is injected into the downcomer, where it is drawn down by the liquid flow. Circulation continues to be supported by the net change of gas voidage between the downcomer and the riser. Gas disengagement is required at the head of the shaft because of the unfavourable effect on liquid circulation that would be caused by carrydown of spent gas. In any case the oxygen content of the spent gas is extremely low. The deep shaft process has a number of advantages over other various biological wastewater treatment processes: (1) transfer of
liquid aeration from shallow horizontal basins to a deep vertical shaft leads to a land requirement that is a tiny fraction of the conventional municipal usage. This has two major advantages. The conserving of land and the saving of its purchase price which, in regions of high population density, can be very significant. Secondly, it may be possible to extend an existing works or build a new one on existing sewers in the population centre, thus saving substantial interceptor sewer construction and, in some cases, pumping costs. (2) the device (apart from the air compressor building) is located below ground level and the gas disengagement basin is relatively small which makes covering easier and cheaper than for conventional aeration basins if this is necessary either for amenity reasons or for protection in a very cold climate (3) the elimination of primary settlement in most cases (4) the oxygen requirement of high strength industrial wastes is comparatively easy to supply (5) less sludge is produced for ultimate disposal (6) activated sludge and aerobic digestion duties may be combined in a single partitioned deep shaft (7) oxygen transfer characteristics that are reliable and beneficial in terms of intensity and economy for both activated sludge duty and aerobic sludge digestion and (8) the power economy for the deep shaft plants has been estimated at three times that of a conventional plant (Gallo and Sandford, 1979).

At present, airlift reactors are receiving increasing interest in wastewater treatment due to their high biomass concentration and efficient performance (Cai et al., 1992; Gjaltema et al., 1995; Heijnen et al., 1992, 1993; Jian-an and Nieuwstad, 1991; Kristiansen and Bu’lock, 1980; Tijhuis et al., 1992a, 1994). A number of investigators have studied the airlift loop reactor at different scales aimed at treatment of various wastewaters. For example, Tang et al. (1987) and Hirata et al. (1990) investigated treatment of phenolic wastewater, Qian and Nieuwstad (1989) municipal wastewater treatment, Maue et al. (1994) wastewater containing emulsified waste oil and Sofer et al. (1990) treatment of 2-chlorophenol wastewater. The biodegradation of naphthalene-sulphonic acid containing sewage was investigated by Krull and Hempel (1994) in two external loop airlift reactors. Tyagi et al. (1990) examined the digestion of municipal sludge in an airlift bioreactor. They concluded that the airlift bioreactor produced stabilized sludge in significantly shorter retention times than conventional digesters.

Jian-an and Nieuwstad (1991; 1992) developed a model to predict the liquid circulation rate in an internal loop airlift reactor for wastewater treatment. They claimed that the aforementioned reactor differs from other types of airlift reactors, due to a shrunk
top cover and low gas supplement. Dluhy et al. (1994) have also mathematically modelled an airlift bioreactor for wastewater treatment.

Hao and Nieuwstad (1993; 1994) and Tijhuis et al. (1992) have studied the possibility of nitrification and/or denitrification of wastewater in airlift reactors. Hao and Nieuwstad (1994) stated that, for pre-settled municipal wastewater, 70-80% nitrogen removal can be reached in an airlift reactor. They concluded that this finding makes airlift reactors competitive for nitrogen removal compared to conventional low-loaded activated sludge systems.

Bacterial process of metal solubilisation from sewage sludge has been examined in airlift reactors (Couillard and Mercier, 1991; Tyagi et al., 1991). Couillard and Mercier (1991) have demonstrated that sludge that could normally not have been spread on agricultural lands, because of copper and cadmium content, were made suitable for land spreading after 18 hours of bacterial leaching in an airlift reactor.

Moo-Young and Chisti (1994) have discussed the general possibility of treating gaseous, liquid and solid wastes in airlift reactors. They have stated that airlift bioreactor based processes will play an increasingly vital role in waste treatment and reuse in the future. Chisti and Moo-Young (1994) explained a process for the surfactant enhanced removal of hydrocarbon contaminants from soils and biotreatment of resulting wastewater in airlift contactors.

Biological treatment of polluted air has also been examined in airlift bioreactors. Ritchie and Hill (1995) investigated the biodegradation of phenol polluted air in an external loop airlift bioreactor using Pseudomonas putida. They reported that the saturated air was found to be cleansed of phenol below the detectable limit (over 99 % degradation) due to the high mass transfer rate of the pollutant from the air and the high growth rate of P. putida. Ritchie and Hill concluded, from mass transfer calculations, that the external loop airlift bioreactor had a very promising future for removing phenol from high volumes of contaminated air. Suzuki et al. (1995) have studied carbon dioxide fixation by microalgae (Dunaliella tertiolecta) in a bubble column and an airlift bioreactor.

In summary, we can conclude that airlift reactors are employed in many different applications. Since commercial organisations which manufacture or treat products are interested in employing technically and economically optimal reactors, airlifts are finding increasing use in industry; but a lack of reliable design basis for scale up is still restricting the wider use of airlifts commercially.
2.6 The Rheology of Fluids

The fluids normally employed in airlift reactors fall into two categories:

1) Newtonian fluids - examples include water, wastewater, salt solutions and most bacterial and yeast fermentation broths; and

2) non-Newtonian fluids - examples include oils, polysaccharide fermentations, polymer solutions and broths of Aspergilli, Penicillia and Streptomyces.

We now consider the definitions of Newtonian and non-Newtonian fluids and their properties.

Viscosity of Newtonian Fluids

Consider a thin layer of fluid between two parallel planes a distance \( dy \) apart as in Figure 2.3. One plate is fixed and a shearing force \( F \) is applied to other. When the conditions are steady the force \( F \) will be balanced by an internal force in the fluid due its viscosity. For a Newtonian fluid in laminar flow the shear stress is proportional to the velocity gradient, i.e.

\[
\frac{F}{A_w} = \tau \propto \frac{du}{dy}
\]

Eqn 2.17 may be written as

\[
\tau = \mu \frac{du}{dy} = \mu \gamma
\]

where the constant of proportionality \( (\mu) \) is called the Newtonian viscosity. It will be seen that \( \mu \) is the tangential force per unit area exerted on layers of fluid a unit distance apart and having a unit velocity difference between them.

The Newtonian viscosity \( (\mu) \) depends only on temperature and pressure and is independent of the rate of shear. The diagram relating shear stress and rate of shear for Newtonian fluids (flow curve) is therefore a straight line of slope \( \mu \) as in Figure 2.4 and the single constant \( (\mu) \) completely characterises the fluid.

Newtonian behaviour is exhibited by fluids in which the dissipation of viscous energy is due to the collision of comparatively small molecular species. All gases and liquids and solutions of low molecular weight come into this category. Notable exceptions are colloidal suspensions and polymeric solutions where the molecular species are large. These fluids show marked deviations from Newtonian behaviour.
Viscosity of Non-Newtonian Fluids

Non-Newtonian fluids are those for which the curve is not linear, that is the 'viscosity' of a non-Newtonian fluid is not constant at a given temperature and pressure but depends on other factors such as the rate of shear in the fluid, the apparatus in which the fluid is contained or even on the previous history of the fluid.

These real fluids for which the flow curve is not linear may be classified into three broad types:

1) fluids for which the rate of shear at any point is some function of the shearing stress at that point and depends on nothing else (time-independent non-Newtonian fluids);

2) more complex systems for which the relation between shear stress and shear rate depend on the time the fluid has been sheared or on its previous history (time-dependent non-Newtonian fluids);

3) systems which have characteristics of both solids and fluids and exhibit partial elastic recovery after deformation, the so-called viscoelastic fluids.

Since the majority of the non-Newtonian fluids which are encountered in the biochemical industry and in wastewater engineering are of the time-independent variety, we will only consider this category in detail. A brief description of the other two categories are also given.

Time-independent non-Newtonian fluids

Fluids whose properties are independent of time may be described by a rheological equation of the form

\[ \gamma = f(\tau) \]  

Eqn 2.19

This equation implies that the rate of shear at any point in the fluid is a simple function of the shear stress at that point. Such fluids may be termed non-Newtonian viscous fluids.

These fluids may conveniently be subdivided into three distinct types depending on the nature of the function in Eqn 2.19. These types are:

1) pseudoplastic fluids;

2) dilatant fluids; and

3) Bingham plastics.

The typical flow curves for these three fluids are shown in Fig 2.4 and compared with the linear relation typical of Newtonian fluids.
Pseudoplastic fluids

Pseudoplastic fluids (also known as shear thinning) show no yield value and the typical flow curve for these materials indicates that the ratio of shear stress to the rate of shear, which may be termed the apparent viscosity ($\mu_a$) falls progressively with shear rate and the flow curve becomes linear only at very high rates of shear. This limiting slope is known as viscosity at infinite shear ($\mu_\infty$).

The logarithmic plot of shear stress and rate of shear for these materials is often found to be linear with a slope between zero and unity. As a result, an empirical functional relation known as the power law (Ostwald de Waele relation) is widely employed to characterise fluids of this type. This relation, which was originally proposed by Ostwald and Auerbach (1926), can be written as

$$\tau = K \gamma^n$$  \hspace{1cm} (2.20)
where $K$ and $n$ are constants ($n<1$) for the particular fluid and $\gamma$ is the average shear rate. $K$ is a measure of the consistency of the fluid, the higher the $K$ the more viscous the fluid; $n$ is a measure of the degree of non-Newtonian behaviour and the greater the departure from unity the more pronounced are the non-Newtonian properties of the fluid. It is important to remember that although $n$ is nearly constant in many cases over wide ranges of shear rate it is not a true constant for real fluids over all possible ranges of shear. This is not a serious drawback in engineering applications because all that is needed is a rheological equation which describes the fluid over the particular range of shear rate encountered in the particular problem. Over such a range $n$ can often be regarded as constant.

The apparent viscosity ($\mu_a$) for a power law fluid may be expressed in terms of $n$ since

$$\mu_a = \frac{\tau}{\gamma}$$

i.e.

$$\mu_a = K \gamma^{n-1}$$

and since $n<1$ for pseudoplastics the apparent viscosity decreases as the rate of shear increases.

Other empirical equations which have been used to describe pseudoplastic behaviour are

1) Prandtl

$$\tau = A \sin^{-1} \left( \frac{\gamma}{C} \right)$$

2) Eyring

$$\tau = \frac{\gamma}{B} + C \sin \left( \frac{\tau}{A} \right)$$

3) Powell-Eyring

$$\tau = A \gamma + B \sinh^{-1} \left( C \gamma \right)$$

4) Williamson

$$\tau = \frac{A \gamma}{B + \gamma} + \mu_a \gamma$$

In these equations $A$, $B$ and $C$ are constants which are typical of the particular fluid. These equations are considerably more difficult to use than the power law and usually do not offer any compensating advantages (Wilkinson, 1960). For most biological fluids the power law model is satisfactory and the question as to whether which of the complex models above better fits a particular fluid is often academic (Chisti, 1989).

Dilatant fluids

Dilatant fluids (also known as shear thickening) are similar to pseudoplastics in that they show no yield stress but the apparent viscosity for these materials increases with
increasing rates of shear. The power law equation is again often applicable but in this case \( n > 1 \).

In the process industries, dilatant fluids are much less common than pseudoplastic fluids but when the power law is applicable the treatment of both types is much the same.

**Bingham plastics**

A Bingham plastic is characterized by a flow curve which is a straight line having an intercept \( \tau_y \) on the shear-stress axis. The yield stress \( (\tau_y) \) is the stress which must be exceeded before flow starts. The rheological equation for a Bingham plastic is

\[
\tau - \tau_y = \mu_v \gamma \quad ; \quad \tau > \tau_y
\]

where \( \mu_v \), the plastic viscosity or coefficient of rigidity, is the slope of the flow curve.

The concept of an idealized Bingham plastic is very convenient in practice because many real fluids closely approximate this type of behaviour. Common examples are slurries, drilling mud and sewage sludge.

**Time-dependent non-Newtonian fluids**

Many real fluids cannot be described by a simple rheological equation such as Eqn 2.19 which applies to fluids for which the relation between shear stress and shear rate is independent of time. The apparent viscosity of more complex fluids depends not only on the rate of shear but also on the time the shear has been applied. These fluids may be subdivided into two classes:

1) thixotropic fluids and
2) rheopectic fluids

according as the shear stress decreases or increases with time when the fluid is sheared at a constant rate.

**Viscoelastic fluids**

A viscoelastic material is one which possesses both elastic and viscous properties, i.e. although the material might be viscous, it exhibits a certain elasticity of shape. The rheological equation for this type of fluid is very complex.

Some biological fluids possess viscoelastic properties which can have important implications in bioreactor design (Chisti, 1989).
2.6.1 Measurement of Rheological Properties

Many types of rheometer have been manufactured to make manual, semi-automatic or automatic measurement of rheological properties (Whorlow, 1980). The choice of rheometer will mainly depend on the type of fluid employed and the range of shear rates that will be encountered in the process of interest. In this section, a brief description of the various classes of viscometers presently available are given. They fall under three categories:

(1) Capillary viscometers

The capillary viscometer (also known as tube or pipeline viscometer) consists of a true-bore capillary, with known length and diameter, attached to a pressurized reservoir containing the fluid under test. The viscosity of the fluid is gauged by measuring the volumetric flow rates for various pressure differences. Many investigators (Allen and Robinson, 1990; Bjorkman 1987 and 1987a; Blakebrough et al., 1978; Perley et al. 1979; Whorlow, 1980) have used the tube viscometer and Warren (1994) has given the advantages and the disadvantages of this type of viscometer.

(2) Rotational viscometers

In this type of viscometer the test fluid is sheared either between rotating cylinders (concentric cylinder viscometer) or cone and plate (cone and plate viscometer) or in a stirred tank type of configuration (impeller viscometer). Rotational viscometry is based on the principle that the movement of a solid surface in contact with a fluid causes the fluid to move in a distinctive pattern which produces the development of forces against the solid surface. The fluid motion is sustained by continuously counter balancing these forces. The nature of the flow pattern and the magnitude of the internal stresses and applied forces depend primarily on the rate of fluid motion, the geometry of the system and the rheological properties of the fluid.

Three main types of rotational viscometers are considered below.

Coaxial cylinder viscometer

The concentric cylinders or cup and bob systems are standard rotational viscometers, which are versatile and widely used (Chisti, 1989; Pollard, 1995; Warren, 1994).
Coaxial cylinder viscometers embody two concentric cylinders containing the fluid to be measured in the narrow gap between the cylinders. Either the inner or the outer cylinder rotates at known angular velocity ($\Theta_i$) and the resultant torque per unit length ($\Gamma$) produced on the stationary cylinder is measured. If there are no end effects, then the shear stress ($\tau$) on the surface of the inner cylinder is

$$\tau = \frac{\Gamma}{2 \pi R_i^2}$$

where $R_i$ is the radius of the inner cylinder. The shear rate ($\gamma$) on inner cylinder is

$$\gamma = \frac{2 \Theta_i}{\left(1 - \frac{R_i^2}{R_o^2}\right)}$$

where $R_o$ is the radius of the outer cylinder. Equation 2.29 applies to Newtonian fluids and for non-Newtonian fluids it needs to be modified (Whorlow, 1980). For power law fluids

$$\tau = K \gamma^n$$

where the parameters $K$ and $n$ are determined as the intercept and the slope, respectively, of log-log plots of $\tau$ versus $\gamma$.

For some fluids containing suspended solids, the narrow gap between the cylinders can prove too small (Kemblowski and Kristiansen, 1986). Impeller viscometers have been developed to deal with these fluids (see below). A number of researchers have successfully employed coaxial cylinder viscometers in their work (Allen and Robinson, 1990; Barker and Worgan, 1983; Ghildyal et al., 1987; Ledy et al., 1974; Pollard, 1995; Russell, 1989; Tangsathitkulchai and Austin, 1988; Warren, 1994; Wittler et al., 1983).

**Cone and plate viscometer**

A cone and plate viscometer shears a fluid sample in the small angle between a flat surface and a rotating cone whose apex just contacts the flat surface. The shear rate throughout the fluid under test is roughly constant providing the cone angle is less than 3 degrees (Warren, 1994). Although this type of viscometer has been employed to study non-Newtonian fluids, their use in the study of fermentation broths has been minimal (Charles, 1978).
Impeller viscometer

This group of viscometers make use of a stirred tank type of configuration. The torque on the stirrer rotating in the fluid at various speeds is established from which the stirring power is calculated. From the stirring power the Power number is obtained. For a known Power number the corresponding impeller Reynolds number is calculated and then the viscosity is obtained (Chisti, 1989).

Impeller viscometers can be used to study particularly mycelial fermentation broths (Kemblowski and Kristiansen, 1986); this is due to the fact that the broth is not a homogeneous liquid and measurements obtained for such liquids using other types of rheometers are of limited value because of gravity settling of the suspended particles and (or) other forms of phase separation. Additionally, large particles (in the fluid) of the same order of magnitude as the measuring gap of the instrument (concentric cylinder or cone and plate) cause destruction of the particles in the shear field and impairment of accurate measurements (Metz et al., 1979; Roels et al., 1974).

Impeller viscometers have been studied by various investigators (Bongenaar et al., 1973; Charles, 1978; Chisti, 1989; Warren, 1994).

(3) Other viscometers

Apart from capillary and rotational viscometers, three various types of viscometers have been developed for use with fermentation systems. These are the Convimeter (Rapp and Wagner, 1984), the vibrating rod viscometer (Picque and Corrieu, 1988) and the falling ball viscometer. Very little information exists on these viscometers and their applications (Warren, 1994).
2.6.2 Shear Rate and Apparent Viscosity

In contrast to Newtonian fluids, where the viscosity is evaluated independently of shear rate, the apparent viscosity for non-Newtonian media following a power law behaviour (Eqn 2.20) depends on the shear rate. This is stated by Eqn 2.22 ($\mu_a = K \gamma^{n-1}$). In order to calculate the apparent viscosity ($\mu_a$), the average shear rate ($\gamma$) in the airlift reactor should be known. However, the measurement of local shear rate in airlifts is complex and requires the fluid velocity profile throughout the reactor (Shi et al., 1990). Hence, usually, an empirical shear rate estimation method is employed.

Many previous investigators have based the average shear rate on the superficial gas velocity into the reactor (Allen and Robinson, 1991; Henzler, 1980; Henzler and Kauling, 1985; Nishikawa, 1991; Nishikawa et al., 1977; Shi et al., 1990; Shumpe and Deckwer, 1987). They have assumed that there exists an average shear rate ($\gamma$) in the reactor which is proportional to the superficial gas velocity supplied to the reactor:

$$\gamma = C U_{sg}$$  \hspace{1cm} 2.31

where C is a constant. However there is a wide variation, among researchers, on the exact value for C (see Table 2.2). Nishikawa et al. (1977) obtained a value of 5000 m$^{-1}$ on the basis of some heat transfer work in bubble columns. Schumpe and Deckwer (1987) did some work with non-Newtonian media in bubble columns and gave a value of 2800 m$^{-1}$. While Henzler (1980) came up with a value of 1500 m$^{-1}$ for C in bubble columns. All these correlations have been used by numerous other investigators for the estimation of shear rate in bubble columns (Deckwer et al., 1982; Godbole et al., 1984; Haque et al., 1986, 1987a; Kelkar and Shah, 1985; Merchuk and Ben-Zvi, 1992; Shah et al., 1982). Moreover, some researchers (Akita et al., 1994; Allen and Robinson, 1987; Popovic and Robinson, 1988) have extended these correlations (obtained in bubble columns) to airlift vessels. Chisti and Moo-Young (1989) have severely criticised this approach. They explained that a bubble column operating at the same superficial gas velocity as the riser of an airlift does not have the same specific power input to it as in the airlift. Hence the turbulence intensities in the two types of reactors must be different and so must be the average shear rates. Allen and Robinson (1991) have tried to justify the extension of Eqn 2.31 (with C as 5000 m$^{-1}$) to airlift vessels. However, they stated that it was unclear as to whether C was independent of column diameter, type of sparger or rheology of the media. Chisti and Moo-Young (1989) even questioned the applicability of the above correlations to bubble columns, since the
average shear rate could vary by more than three fold depending on which value of C was chosen.

As far as this author is aware, only Shi et al. (1990) have proposed an empirical correlation for the average shear rate in airlift reactors. They obtained this correlation in an external loop reactor with aqueous CMC and xanthan solutions. This relationship was also based only on the superficial gas velocity into the reactor. Shi et al. (1990) found that the average shear rate in airlift reactors was lower than in bubble columns.

<table>
<thead>
<tr>
<th>Reference</th>
<th>Proposed correlation</th>
<th>Reactor type</th>
</tr>
</thead>
<tbody>
<tr>
<td>El-Temtamy et al. (1984)</td>
<td>( \gamma = \frac{U_g}{d_c} )</td>
<td>Bubble column</td>
</tr>
<tr>
<td>Henzler (1980)</td>
<td>( \gamma = 1500 \ U_g )</td>
<td>Bubble column</td>
</tr>
<tr>
<td>Henzler and Kauling (1985)</td>
<td>( \gamma = \left( \frac{\rho_i g U_g}{\mu_i} \right) )</td>
<td>Bubble column</td>
</tr>
<tr>
<td>Kawase and Moo-Young (1986a)</td>
<td>( \gamma = \frac{U_g}{d_c} )</td>
<td>Bubble column</td>
</tr>
<tr>
<td>Nishikawa et al. (1977)</td>
<td>( \gamma = 5000 \ U_g )</td>
<td>Bubble column</td>
</tr>
<tr>
<td>Schumpe and Deckwer (1977)</td>
<td>( \gamma = 2800 \ U_g )</td>
<td>Bubble column</td>
</tr>
<tr>
<td>Shi et al. (1990)</td>
<td>( \gamma = 14800 \ U_g^2 - 351 \ U_g + 3.26 )</td>
<td>External loop airlift reactor</td>
</tr>
</tbody>
</table>

Table 2.2: Literature correlations for the estimation of average shear rate.

Chisti (1989) has expressed doubts on the validity of correlations which consider the average shear rate as just a function of the superficial gas velocity. He stated that the degree of turbulence in a bioreactor must depend not only on the power input into the reactor, but also on the momentum transport characteristics (viscosity and density) of the fluid itself.

Some investigators (Pollard, 1995; Russell; 1989; Wood and Thompson, 1986) have employed the Blasius correlation for estimating the shear stress at a vessel wall. However, this relationship is only applicable to Newtonian media. This correlation gives an estimate of the maximum shear stress and not an average value.
2.7 Two Immiscible Liquid Phases in Pneumatic Devices

So far, for no apparent reason, all of the studies performed in airlift reactors have been with only one liquid system. Chisti (1989) commented that work on two-liquid systems (e.g., oil/water), with a gas providing the necessary power input and oxygen, in airlift reactors was still lacking. In bubble columns, there are some studies of the hydrodynamics with two immiscible liquid phases (Arguelles et al., 1993; Hatzikiriakos et al., 1990; 1990a; Pino et al., 1990; 1990a; 1992; Siquier et al., 1991).

The physical properties of a single liquid phase are well known to have a strong effect on the hydrodynamics and liquid mixing in airlift reactors. Therefore, when the liquid phase is a mixture of two immiscible liquids the fundamental behaviour is even more complex on account of the fact that three different fluid-fluid interfaces are usually present (Arguelles et al., 1993). Arguelles et al. (1993) observed a decrease in the global gas holdup upon an addition of an immiscible liquid phase (water) to a gas-liquid (air/commercial kerosene, $\mu_L = 1.60$ mPa s) bubble column. In two-phase bubble columns, Pino et al. (1990) reported that pure kerosene yielded higher gas holdups than water. Hatzikiriakos et al. (1990) also observed that mixtures of two immiscible liquid phases (dibenzyl ether and water; TCE water and linseed, cottonseed and castor oils) exhibited lower gas holdups than water in a gas-liquid bubble column. This reduction in gas holdup with immiscible liquids has been ascribed to the increase in the apparent viscosity of the liquid due to the presence of dispersed drops (Arguelles et al., 1993). Pino et al. (1990) suggested that an increase in the liquid viscosity produced an increase in the bubble coalescence rate and hence larger bubbles which have a high rise velocity in the column. However, Arguelles et al. (1993) suggested that this explanation might be a bit too simplistic considering the variety of observations previously mentioned and the possible interactions at the different fluid-fluid interfaces.

Some fermentation media, in biochemical applications, have two immiscible liquid phases where usually one of the liquids acts as a food source for the microorganisms. In the chemical industry, gas-agitated liquid-liquid dispersions arise in applications such as coal liquefaction (Shaw et al., 1988) and pyrometallurgical processes for the production of steel and copper and processes for electro-organic synthesis (Hatzikiriakos et al., 1990). Hence, the study of the hydrodynamics and mixing of two immiscible liquid systems in airlift reactors would be of some interest.
3 THEORETICAL MODELLING

3.1 HYDRODYNAMIC MODEL

In view of the shortcomings of the existing hydrodynamic correlations (Section 2.4) on airlift reactors, and considering their lack of generality, Ayazi Shamlou et al. (1994) have presented a new hydrodynamic model. In this section a revised version of this model (Rajarajan and Ayazi Shamlou, 1995; Rajarajan et al., 1995) is given. This model, which is based on energy balances, is applicable to all types of airlift reactors: external and internal loop.

3.1.1 Model Development

If a gas is introduced in the form of discrete bubbles into the base of a stagnant column of liquid, the rising bubbles cause the liquid to circulate around the column by two separate mechanisms:

1) liquid motion is caused by liquid entrainment and transport in the wake of the ascending bubbles;

2) bulk liquid flow occurs by 'natural circulation', due to the differences of the density of the mixture at the centre and that near the vertical walls of the vessel.

In a bubble column, these density differences are always present because the rising bubbles tend to concentrate more towards the axis of the column. In a concentric tube internal loop airlift device, for example, differences in gas holdup between the core and the wall region are profoundly increased by the presence of the draft tube, which enhances the overall circulation of the liquid.

Under steady-state operation in the column, the fractional phase holdups are related by

\[ \varepsilon_g + \varepsilon_r + \varepsilon_w = 1 \]  \hspace{1cm} 3.1

Experimental measurements of the liquid-wake volume are reported in the literature for multiphase flows (Ostergaard, 1965; Philip et al., 1990). Darton and Harrison (1975) have given the following equation for the estimation of the ratio of the liquid-wake volume to bubble volume:

\[ k = 1.4 \left( \frac{U_{\text{fl}}}{U_g} \right)^{0.33} - 1 \]  \hspace{1cm} 3.2
Darton and Harrison (1975) obtained this equation from data referring to particulate fluidised beds using water as liquid. They used their data together with those of Efremov and Vakhrushev (1970) and Michelsen and Ostergaard (1970).

The linear liquid velocity (due to density difference) $U_{li}$ and the superficial liquid velocity ($U_{sl}$) are related as follows in Darton and Harrison's (1975) paper:

$$U_{li} = U_{sl} (1 - \varepsilon)$$  \hspace{1cm} (3.3)

Substituting Eqn 3.3 into 3.2, we obtain

$$k = 1.4 \left[ \frac{U_{li} (1 - \varepsilon)}{U_{sl}} \right]^{0.33} - 1$$ \hspace{1cm} (3.4)

By definition, $k$ is equal to $\varepsilon_w / \varepsilon_g$; therefore

$$k = \left( \frac{\varepsilon_w}{\varepsilon_g} \right) = 1.4 \left[ \frac{U_{li} (1 - \varepsilon)}{U_{sl}} \right]^{0.33} - 1$$ \hspace{1cm} (3.5)

The gas voidage ($\varepsilon_g$), in Eqns 3.1 and 3.5, is related to the superficial fluid velocities through the slip velocity ($v_r$) which is the mean relative velocity between the gas and the liquid phase. In gas-liquid systems in which there is a net flow of both phases, the slip velocity is

$$v_r = \left( \frac{U_{sl}}{\varepsilon_g} \right) - \left[ \frac{U_{sl}}{1 - \varepsilon} \right]$$ \hspace{1cm} (3.6)

Wallis (1969) has proposed a method of characterising the relative motion using the drift flux of gas ($J_{g0}$); defined as the volumetric flux of gas relative to a surface moving at the average velocity, i.e.

$$J_{g0} = v_r \varepsilon_g (1 - \varepsilon_g)$$ \hspace{1cm} (3.7)

Hence, substituting Eqn 3.6 into Eqn 3.7:

$$J_{g0} = U_{sl} (1 - \varepsilon_g) - U_{sl} \varepsilon_g$$ \hspace{1cm} (3.8)

In the uniform bubbling regime, in which the bubbles are all of a similar size, the gas drift flux is a function only of the gas holdup and the rising velocity of an isolated bubble, $U_B$. There are various equations known for the influence of holdup on drift flux; all these are dependent on how a particular bubble is influenced by its neighbours. These equations are:

$$J_{g0} = \varepsilon_g U_B$$ \hspace{1cm} (3.9)

$$J_{g0} = \varepsilon_g U_B (1 - \varepsilon_g)$$ \hspace{1cm} (3.10)
Theoretical Modelling

\[ J_y = \varepsilon_g U_B \left( 1 - \varepsilon_g \right)^m \]  
3.11

Equation 3.9 supposes the velocity of each bubble to be enhanced by its neighbours and has been verified by Nicklin et al. (1962) in the case of gas-liquid slug flow. Turner (1966) proposed Eqn 3.10 which assumes no interaction between bubbles. Equation 3.11 is the correlation of Richardson and Zaki (1954); the parameter \( m \) has a value between 0 and 3. If \( m > 1 \), this supposes the presence of other bubbles to increase the effective viscosity of the liquid and hence to retard the rate of rise.

Gas holdup in airlift contactors is normally sufficiently low enough to permit the use of Eqn 3.9. In the churn-turbulent regime Eqns 3.9 - 3.11 do not apply and the approach of Zuber and Findlay (1965) may be used.

The drift-flux model of Zuber and Findlay (1965) stated that

\[ \frac{U_g}{\varepsilon_g} = C_o \left( U_g + U_w \right) + \frac{(U_d \varepsilon_g)_{\text{average}}}{\varepsilon_g} \]  
3.12

Where \( C_o \) (termed the distribution parameter) is a constant accounting for the interaction of velocity and gas void distributions. The drift velocity term containing \( U_d \) accounts for the local slip between the phases. In churn-turbulent bubbly upflow, where bubble interaction and waking are strong, the drift velocity term is equated to the bubble rise velocity in an infinite medium, \( U_{bi} \).

Hence equation 3.12 becomes

\[ \frac{U_g}{\varepsilon_g} = C_o \left( U_g + U_w \right) + U_{bi} \]  
3.13

But the rising velocity of an isolated bubble in stagnant fluid (\( U_B \)) is equal to \( U_g / \varepsilon_g \), therefore we can rewrite Eqn 3.13 as

\[ U_B = C_o \left( U_g + U_w \right) + U_{bi} \]  
3.14

Substituting Eqn 3.9 into Eqn 3.8,

\[ \varepsilon_g U_B = U_g \left( 1 - \varepsilon_g \right) - U_w \varepsilon_g \]  
3.15

Using Eqn 3.15 into Eqn 3.14 and solving gives

\[ \varepsilon_g = \frac{U_g}{\left( 1 + C_o \left( U_g + U_w \right) + U_{bi} \right)} \]  
3.16
Theoretical Modelling

Determination of gas holdup from equation 3.16, requires the knowledge of the distribution parameter \( C_0 \), the superficial liquid velocity \( U_{si} \) and the bubble terminal rise velocity \( U_{bt} \).

**Analysis for the determination of the superficial liquid velocity**

In this model the total superficial liquid velocity along the vertical axis of the column \( U_{si} \) is assumed to be equal to the sum of the contributions of density differences between the riser and the downcomer and the liquid-wake associated with the rising bubbles.

The superficial liquid velocity can be obtained from the following mass balance for the liquid:

\[
U_{si} = U_{bi} \varepsilon_s + U_w (\varepsilon_s + \varepsilon_w) \quad 3.17
\]

where \( U_{bi} \) is the linear liquid velocity due to gas holdup differentials in the column and \( U_w \) is the linear velocity of the liquid-wake.

Ostergaard (1965) has in his paper given an equation relating bubble velocity to the liquid-wake; in the present analysis it is assumed that the linear velocity of the liquid-wake is equal to the rising velocity of an isolated bubble in stagnant fluid \( U_B \). i.e.

\[
U_w = U_B = \frac{U_w}{\varepsilon_s} \quad 3.18
\]

Substituting Eqn 3.18 into Eqn 3.17 and using Eqn 3.1 and Eqn 3.5 to eliminate \( \varepsilon_l \) and \( \varepsilon_w \) from the resulting term gives

\[
U_{si} = U_{bi} \left[ 1 - \varepsilon_s (1 + k) \right] + U_w \left( 1 + k \right) \quad 3.19
\]

**Evaluation of an expression for the linear liquid velocity along the column due to gas holdup differentials**

Several equations are available for the estimation of the linear liquid velocity due to density difference (Chisti *et al.*, 1988; Joshi and Sharma, 1979; Philip *et al.*, 1990). These equations are based either on an energy balance or a pressure balance carried out over the circulation path of the column. In this analysis, the energy balance method is used; models based on energy balances are considered to be more widely applicable (Chisti, 1989).
The Energy Balance

The energy balance over an airlift loop may be written as:

Rate of energy input into reactor = Rate of energy dissipation in the reactor

Energy input

The rate of energy input into the liquid by the gas (Joshi and Sharma, 1979):

\[ E_i = \frac{\pi T^2 U_{sg} \rho_L g H}{4} + \frac{A_n U_{sa} \rho_G}{2} \]

where the first term represents the potential energy and the second term the kinetic energy of the gas.

Under normal operating conditions, the kinetic energy of the gas is negligible (Joshi and Sharma, 1979). Furthermore, Lehrer (1968) has shown that about 94% of the kinetic energy is dissipated near the sparger exit. Therefore, we can modify Eqn 3.20 to:

\[ E_i = \frac{\pi T^2 U_{sg} \rho_L g H}{4} \]

Energy loss

The energy dissipation in the column is assumed to be mainly due to the following factors:

1) energy dissipation in the wakes behind the bubbles,
2) energy loss in forming the bubbles,
3) energy dissipation in the net bubble break-up,
4) energy wasted due to bulk liquid flow through the riser, and
5) energy loss due to friction and fluid turn-around at the bottom and the top of reactor.

In this investigation, the energy losses due to bubble formation and bubble break-up are ignored; Joshi and Sharma (1979), after performing an order of magnitude calculation, have declared that the energy loss due to bubble formation and bubble break-up is negligible. Hence the most important modes of energy dissipation are due to the bulk liquid flow through the riser \((E_b)\), the friction and fluid turn-around at the bottom and top of the reactor \((E_f)\) and the liquid-wake behind the bubbles \((E_w)\).
**Theoretical Modelling**

Energy losses due to the bulk liquid flow through the riser ($E_b$)

The energy dissipation rate in the liquid motion is (Rietema and Ottengraf, 1970):

$$E_b = - \int_0^R 2 \pi r H \tau \left( \frac{dU_{bl}}{dr} \right) dr$$ \hspace{1cm} 3.22

Where $U_{bl}$ is the axial liquid velocity in the bubble column as a function of the radial position ($r$).

Kawase and Moo-Young (1986) provide experimental evidence to show that the liquid velocity distribution in a bubble column and the riser section of an airlift contactor is:

$$\frac{U_{bl}}{U_c} = 1 - 2 \left( \frac{r}{R} \right)^3$$ \hspace{1cm} 3.23

where $U_c$ is the liquid velocity at the column axis.

Kawase and Moo-Young (1986) state that Eqn 3.23 is applicable to non-Newtonian fluids as well as to Newtonian fluids. They used Eqn 3.23 together with an expression for the average mixing length and the shear stress in the turbulent flow region to solve Eqn 3.22 and obtain the rate of energy dissipation in the liquid motion ($E_b$):

$$E_b = 0.512 n^2 \pi \rho_l H T U_c^3$$ \hspace{1cm} 3.24

Energy loss due to friction and fluid turn-around ($E_f$)

The energy dissipation ($E_f$) arising from changes in flow direction at the top and bottom of the reactor can be calculated by employing the concept of frictional loss coefficients which is well established for pipe flow. Garcia Calvo (1989) has given an expression for the energy loss due to the friction between fluids and the device:

$$E_f = 0.5 \left( \frac{A_d}{A_j} \right) K_f \rho_l v_l^2$$ \hspace{1cm} 3.25

where $v_l$ is the linear liquid velocity in the column and $K_f$ is the total frictional loss coefficient.

In the present analysis, we have taken the linear liquid velocity in Eqn 3.25 to be $U_{bl}$ (i.e. the liquid velocity due to density differences between the riser and the downcomer). Hence we can rewrite equation 3.25 as:

$$E_f = 0.5 \left( \frac{A_d}{A_j} \right) K_f \rho_l U_{bl}^3$$ \hspace{1cm} 3.26
Energy dissipation due to the liquid-wake behind the bubbles

The energy loss in the bubble wakes \( (E_w) \) is supposed to be the result of three terms (Whalley and Davidson, 1974), which are:

1) rate of loss of pressure energy of the liquid \[ A \rho L H U g(1- \varepsilon_g) \],

2) rate of loss of pressure energy of the gas \[ A \rho L H g U sg(1- \varepsilon_g) \], and

3) rate of gain of potential energy of the liquid \[ AH \rho L U sg \].

That is

\[
E_w = (\text{rate of loss of pressure energy of the liquid}) + (\text{rate of loss of pressure energy of the gas}) - (\text{rate of gain of potential energy of the liquid})
\]

\[
= A \rho L H U g(1- \varepsilon_g) + A \rho L H g U sg(1- \varepsilon_g) - [AH \rho L U sg]
\]

\[
= A \rho L g [U g(1- \varepsilon_g) + U sg(1- \varepsilon_g) - U sl]
\]

If \( A \) is taken as the cross-sectional area of the riser section, then we can conclude that

\[
E_w = \frac{\pi T^2 H \rho g}{4} \left[ U g \left( 1 - \varepsilon_g \right) - U sl \varepsilon sl \right] \quad 3.27
\]

Substituting the slip velocity term (Eqn 3.6) into Eqn 3.27 and saying that the slip velocity \( (V_r) \) is equal to \( U bl \) (Turner, 1966), we can rewrite equation 3.27 as:

\[
E_w = \frac{\pi T^2 H U bl \rho g \varepsilon g}{4} \left( 1 - \varepsilon_g \right) \quad 3.28
\]

Joshi and Sharma (1979) and Kemblowski et al. (1993) have in their paper quoted an expression similar to Eqn 3.28.

An expression for \( U bl \)

The energy balance is:

\[
E_i = E_b + E_f + E_w \quad 3.29
\]

Substituting equations 3.21, 3.24, 3.26 and 3.28 into Eqn 3.29 and taking the liquid velocity at the column axis \( (U_c) \) to be approximately equal to 1.325 \( U bl \) (Verlaan et al., 1986), we obtain

\[
U bl = \left( \frac{T g \left[ U g - U b l \varepsilon g \left( 1 - \varepsilon_g \right) \right]}{4 \left( 1.19 n^2 + \frac{A_s K_f}{2 \pi H T A_r} \right)} \right)^\frac{1}{3} \quad 3.30
\]
3.1.2 Application

The gas holdup and the superficial liquid velocity can be estimated for the riser section of an airlift using equations 3.5, 3.16, 3.19 and 3.30. The model applies to airlifts operating in bubbly flow, coalesced bubble flow and churn turbulent flow regimes. In order to obtain numerical solutions to these equations, values for certain parameters have to be obtained. For a given superficial gas velocity, the distribution parameter ($C_o$), the loss coefficient ($K_d$) and the terminal bubble rise velocity ($U_{bt}$) must be known, together with the physical properties of the liquid and the gas. We shall now consider the distribution parameter, loss coefficient and the terminal bubble rise velocity in more detail.

3.1.2.1 The distribution parameter

In their general expression, for the average volumetric concentration, Zuber and Findlay (1965) invented the distribution parameter term ($C_o$) to take into account the effect of non-uniform flow and concentration profiles in any two phase flow regime. They defined the distribution parameter as

$$C_o = \frac{\langle \alpha j \rangle}{\langle \alpha \rangle \langle j \rangle} = \frac{\left[ \frac{1}{A_o} \right] \left[ \alpha_j dA_o \right]}{\left[ \frac{1}{A_o} \right] \left[ j dA_o \right] \left[ \frac{1}{A_o} \right] \left[ \alpha_j dA_o \right]}$$

3.31

To investigate the effects of non-uniform flow and concentration distribution on the value of the distribution coefficient, we consider an axially symmetric flow through a circular pipe and assume that the flow and concentration distributions are given by (Zuber and Findlay, 1965)

$$\frac{j}{j_c} = 1 - \left( \frac{r}{R} \right)^m$$

3.32

and

$$\frac{\alpha - \alpha_c}{\alpha_c - \alpha_w} = 1 - \left( \frac{r}{R} \right)^p$$

3.33

where the subscripts c and w refer to the values evaluated at the centre line and at the wall of the circular pipe. The superscripts m and p are the exponents on velocity and void distribution, respectively. Inserting equations 3.32 and 3.33 into Eqn 3.31, we obtain the following expressions for the distribution parameter:
Theoretical Modelling

\[ C_o = 1 + \left[ \frac{2}{(m+p+2)} \right] \left[ \frac{1 - \alpha_w}{\langle \alpha \rangle} \right] \]

when expressed in terms of the volumetric concentration \((\alpha_w)\) at the wall or when

\[ C_o = \left[ \frac{m+2}{(m+p+2)} \right] \left[ \frac{1 + p\alpha_c}{(m+2)\langle \alpha \rangle} \right] \]

expressed in terms of the volumetric concentration \((\alpha_c)\) at the centre line.

Zuber and Findlay (1965) gave three situations where the distribution parameter changes with the concentration:

1) if the concentration is uniform across the pipe, then

\[ \alpha_w = \alpha_c = \langle \alpha \rangle \]

from equations 3.34 and 3.35

\[ C_o = 1 \]

2) if the concentration at the centre line is smaller than that close to the wall, then

\[ \alpha_c < \alpha_w \]

and

\[ C_o < 1 \]

3) if the concentration at the centre line is larger than that at the wall, then

\[ \alpha_c > \alpha_w \]

and

\[ C_o > 1 \]

In the hydrodynamic model, the distribution parameter \((C_o)\) characterises the shape of the liquid velocity profile in the column. The value of the distribution parameter \((C_o)\) also depends on the flow and concentration profiles (Clark and Flemmer, 1985; Zuber and Findlay, 1965). The flow and distribution of bubbles across the column are affected by the type and design of the sparger (Ayazi Shamlou et al., 1994). Therefore, the distribution parameter is also expected to be influenced by sparger design since the liquid velocity profile in the column is dependent on the flow and distribution of bubbles across the column.

Zuber and Findlay (1965) stated that for fully established parabolic profiles (similar to laminar profiles), the distribution parameter attained a value of 1.5 and for flat profiles, it tended to reach a value of 1.0 (when the concentration at the centre line is larger than that at the wall). For most two phase flow situations, including bubble columns and airlift devices, the gas holdup at the centre exceeds that at the wall (for fully established profiles).
and this would suggest that $C_o > 1$. However, Ayazi Shamlou et al. (1994) have used a $C_o$ value of 0.7 for a concentric tube airlift with a ring sparger because this was found to give a better agreement between model predictions and experimental data. They explained that since their ring sparger ejected air close to the draft tube, it was not surprising that the liquid flow profile in the riser had a velocity higher nearer the wall than at the centre (close to the sparger). Furthermore, they stated that there was a gradual change in the velocity profile as the liquid moved up the riser and hence the value of $C_o$ represented the average of the liquid velocity profiles in the riser section of an airlift.

Theoretical calculations suggested a maximum value of 1.6 for $C_o$ (Zuber and Findlay, 1965). Govier and Aziz (1972) proposed a value of 1.15 for continuous bubble swarm and bubble sizes ranging from very small bubbles up to an equivalent bubble diameter of 2 cm. In churn turbulent flow, where bubble interaction and waking are strong, $C_o$ has been found to assume values between 1.1 and 1.3 (Nassos and Bankoff, 1967; Wallis, 1969). Hatch (1973) determined the $C_o$ value experimentally in a draft tube of an internal loop airlift reactor and found it to be 1.065. Clark and Jones (1987) employed a $C_o$ value of 1.2 in their calculation; they also implied that higher values of the distribution parameter (between 2.0 and 5.0) may be required to explain certain two phase flow situations (e.g. circulation due to the maldistribution of air caused by a single-pipe sparger). Kelkar and Shah (1985) experimentally determined the value of $C_o$ for various CMC solutions in a bubble column. They found the distribution parameter to vary from 2.36 to 2.76 as the concentration of the CMC solution changed.

The effect of varying the distribution parameter, in the hydrodynamic model, on riser gas holdup and liquid velocity for increasing superficial gas velocity (keeping all other parameters constant) are shown in Figures 3.1 and 3.2. The external loop airlift reactor with water as the liquid was chosen. The loss coefficient was taken to be 0.1, while 0.25 m/s was the value of the terminal bubble rise velocity. $C_o$ was varied from 0.7 to 1.7 in steps of 0.2. It can seen that higher values of $C_o$ gave lower holdup values (Fig 3.1) and higher liquid velocity values (Fig 3.2) than lower values of $C_o$. 

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Kelkar and Shah (1985) experimentally determined the value of $C_o$ for various CMC solutions in a bubble column. They found the distribution parameter to vary from 2.36 to 2.76 as the concentration of the CMC solution changed.
Figure 3.1: The effect of the distribution parameter \(C_o\) on the riser gas holdup - model predictions.

Figure 3.2: The effect of the distribution parameter \(C_o\) on the riser liquid circulation velocity - model predictions.
3.1.2.2 The loss coefficient

The $K_f$ is the loss coefficient for the top and the bottom section of an airlift reactor. Naturally, it is expected to vary with different airlift reactor geometries and with the type of fluid employed in the airlift contactor. Aayazi Shamlou et al. (1994) employed a $K_f$ value of 5.0 for an air/yeast ($Saccharomyces cerevisiae$) system in a pilot-plant internal loop airlift reactor. In the same reactor for a more viscous system (air/$Saccharomyces erythraea$) they used a $K_f$ value of 30.0 to get good agreement between model predictions and experimental data. Garcia Calvo et al. (1991) have divided the friction loss coefficient ($K_f$) into four components: (1) friction coefficient in the riser, (2) friction coefficient in the downcomer, (3) coefficient in the bottom connecting zone and (4) coefficient in the gas separator. They went on to calculate the friction coefficients in the riser and the downcomer from the well-known Poiseuille equation for straight tubes and laminar regime as

$$K_f = \frac{64 \mu L}{d^2 \rho L} \quad 3.42$$

The friction losses in the top and bottom section can be introduced into Eqn 3.42 as an equivalent length. The total friction coefficient is then (Garcia Calvo et al., 1991):

$$K_f = 64 \mu \left[ \frac{(H_r + L_w) \, A_d^2}{V_r \, T^2 \rho \, A_r^2} + \frac{(H_w + L_w)}{V_w \, d^2 \rho} \right] \quad 3.43$$

Verlaan et al. (1986) separated the loss coefficient ($K_f$) into eight different parts. In their study the calculated $K_f$ was 1.8 and 4.75 in the smaller and the larger external loop airlifts respectively, for air/water system. For an internal loop airlift device, Chisti et al. (1988) suggested that the flow in the gas separator is similar to that of an open channel, hence energy losses due to flow reversals at the top of the column are expected to be relatively small compared with the bottom section. They employed a $K_b$ value of 5.0 in a draft tube internal loop airlift and a value of 7.38 in a split cylinder internal loop airlift for air/water system.

The significance of varying the loss coefficient ($K_f$), in the hydrodynamic model, on riser gas holdup and liquid velocity for increasing superficial gas velocity (keeping all other parameters constant) are shown in Figures 3.3 and 3.4. The external loop airlift reactor was selected with water as the liquid. The distribution parameter was taken to be 0.9 and the terminal bubble rise velocity was taken as 0.25 m/s. $K_f$ was varied from 0.1 to 100. As the loss coefficient increased the riser gas holdup also increased. But the riser liquid velocity decreased as the loss coefficient increased.
Figure 3.3: The effect of the loss coefficient ($K_f$) on the riser gas holdup - model predictions.

Figure 3.4: The effect of the loss coefficient ($K_f$) on the riser liquid circulation velocity - model predictions.
3.1.2.3 The terminal bubble rise velocity

The rising velocity of a single bubble in isolation \((U_{bt})\) has been studied comprehensively (Clift et al., 1987; Govier and Aziz, 1972). The velocity with which bubbles of gas rise in a stagnant liquid depends upon the fluid properties and rather remarkably upon the bubble volume or equivalent diameter. There are two approaches to the question of the rise velocity of bubbles in a stagnant liquid:

1) under essentially ideal conditions of spherical bubbles rising in laminar motion the problem is amenable to theoretical analysis based upon the continuity and momentum equations
2) under any reasonably steady state condition the problem may be approached, as in the case of the settling of solid particles, from the viewpoint of a balance between the buoyant and the drag force acting on the bubble; the drag force is commonly determined empirically through a drag coefficient.

Govier and Aziz (1972) have followed the second approach and derived the bubble rise velocity to be

\[
U_{bt} = 1.35 \left( \frac{2\sigma}{\rho_g d_b} \right)^{0.50}\] 3.44

The Table 3.1 below includes Eqn 3.44 and further equations for the calculation of \(U_{bt}\) (Ayazi Shamlou et al., 1994).

<table>
<thead>
<tr>
<th>Equivalent bubble radius, (R_b) (cm)</th>
<th>Single bubble terminal rise velocity, (U_{bt}) (m/s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>(R_b &lt; 0.035)</td>
<td>(\frac{1}{9} \left( \frac{g R_b^2}{\nu} \right))</td>
</tr>
<tr>
<td>(0.035 &lt; R_b &lt; 0.07)</td>
<td>(0.33 \left( \frac{g^{0.76} R_b^{1.28}}{\nu^{0.52}} \right))</td>
</tr>
<tr>
<td>(0.07 &lt; R_b &lt; 0.2)</td>
<td>(1.35 \left( \frac{\sigma}{(R_b \Delta \rho)} \right)^{0.5})</td>
</tr>
<tr>
<td>(0.2 &lt; R_b &lt; 0.6)</td>
<td>(1.53 \left( \frac{g \sigma \Delta \rho}{\rho_g^2} \right)^{0.25})</td>
</tr>
<tr>
<td>(R_b &gt; 0.6)</td>
<td>(1.0 \left( g R_b \right)^{0.5})</td>
</tr>
</tbody>
</table>

Table 3.1: Terminal rise velocity of a single bubble as a function of the diameter of the bubble.
Data for airlift columns suggest that, for orifice and perforated disc spargers, bubble radius in the column appears to fall in the range 0.2 - 0.3 cm. Hence according to Table 3.1, in this range, the terminal rise velocity of single bubbles is independent of bubble radius. For water, using Table 3.1 gives a $U_{bt}$ value of 25 cm/s (Ayazi Shamlou et al., 1994). Freedman and Davidson (1969) have employed a value of 23.5 cm/s for air/water system and Joshi and Sharma (1979) have taken a value of 23 cm/s for air/water system, when the bubble diameter varies from 0.72 cm to 0.6 cm.

The effect of changing the terminal bubble rise velocity, in the model, on riser gas holdup and liquid velocity for increasing superficial gas velocity (keeping all other parameters constant) is shown in Figures 3.5 and 3.6. The external loop airlift reactor with water was chosen. The distribution parameter and the loss coefficient were taken to be 0.9 and 0.1, respectively. $U_{bt}$ was varied from 0.20 to 0.30 m/s in steps of 0.225 m/s. It can seen that as $U_{bt}$ increased, gas holdup and liquid velocity values decreased (Figs 3.5 and 3.6).

Figure 3.5: The effect of the terminal bubble rise velocity ($U_{bt}$) on the riser gas holdup - model predictions.
Figure 3.6: The effect of the terminal bubble rise velocity ($U_{bt}$) on the riser liquid circulation velocity - model predictions.

### 3.1.3 Prediction of Gas Holdup and Liquid Circulation Velocity

For a given airlift configuration and a gas and liquid of known physical properties, numerical solutions to equations 3.5, 3.16, 3.19 and 3.30 can be acquired using an iterative procedure as outlined below for the case of the riser section of an airlift contactor:

1) Assume a value for the gas holdup ($\varepsilon_g$) and use it in equation 3.30 to obtain a value for the linear liquid velocity due to density difference ($U_{yi}$). In order to calculate $U_{bi}$, the value of the single bubble rise velocity ($U_{bt}$) and the loss coefficient ($K_f$) must be estimated. The terminal bubble rise velocity ($U_{bt}$) can be obtained from Table 3.1. The loss coefficient ($K_f$) usually depends on the type of liquid in the column and on the physical shape of the top and bottom section of the reactor (between 0.1 and 100).

2) The calculated value of $U_{bi}$ is used in Eqn 3.5 to obtain a value for $k$. The assumed value of $\varepsilon_g$, together with the calculated value of $U_{bi}$ and $k$ is substituted into Eqn 3.19 to acquire a value for the overall superficial liquid velocity ($U_{si}$).
3) The computed value of $U_{st}$ together with a suitable value for the distribution parameter \( C_0 \) is inserted into equation 3.16 to get a value for the gas holdup \( \varepsilon_g \). Initially, \( C_0 \) is assumed to have a value close to unity.

4) The value of the gas holdup obtained from (3) is compared with the assumed value from (1), and if the two values of \( \varepsilon_g \) are not the same, the above steps are repeated with a new estimate of \( \varepsilon_g \) until a fit is obtained, to the desired number of decimal places.

The above iterative procedure is ideally suited to solving by a micro-computer. Either a computer program can be specifically written to solve the problem or a commercially available 'mathematical' software can be utilised. In our case, we chose the latter option and wrote macros on the spreadsheet Excel™ (Version 4.0 for Windows™) from Microsoft Corporation. This software is powerful (usually compares the two \( \varepsilon_g \) values to ten decimal places) and is relatively fast.
Figure 3.7: Algorithm for the simultaneous calculation of the riser gas holdup and liquid circulation velocity, for a given gas flow and reactor geometry.
3.2 SHEAR RATE

In view of the inadequacy of the existing correlations to calculate the average shear rate in airlift reactors for non-Newtonian fluids (section 2.6.2), in this section we present an approximate equation for the calculation of the average shear rate in airlift reactors. As far as the author is aware, this equation is the only correlation that indicates the effect of gas holdup, liquid velocity and flow behaviour index on the average shear rate.

For low to intermediate gas flow rates \( (U_{sg} < 0.04 \text{ ms}^{-1}) \), which is the case of interest in this study, the shear rate at the wall of the riser is assumed to be due primarily to the flow of the liquid. The shear rate at the wall of a pipe can be defined as (Govier and Aziz, 1972)

\[
\gamma_w = \frac{8 V_{lr} \left(1 + 3 \frac{n}{n} \right)}{d_p \left(4 \frac{n}{n} \right)}
\]

In Eqn 3.45, \( V_{lr} \) is the linear liquid velocity which in the case of two phase flow is given by \( V_{lr} = \frac{U_s}{(1 - \varepsilon_g)} \). Additionally, for low gas flow rates, the radial liquid velocity profile in the column is reasonably flat (Hills, 1974; Yu and Kim, 1991). Hence, Eqn 3.45 can be rewritten as

\[
\gamma = \frac{8 U_s}{T \left(1 - \varepsilon_g \right)} \left(1 + 3 \frac{n}{n} \right)
\]

Eqn 3.46 shows that the average shear rate \( (\gamma) \) in the riser section of an airlift reactor is affected by the superficial liquid velocity \( (U_s) \), the gas holdup \( (\varepsilon_g) \), the flow behaviour index \( (n) \) of the liquid and the diameter of the riser section \( (T) \). For low values of riser gas holdup, the change in the term \( (1 - \varepsilon_g) \) is small and therefore the average shear rate \( (\gamma) \) seems to be unaffected by gas holdup. However, the shear rate is strongly affected by the superficial liquid velocity which is in turn closely related to the gas holdup. Thus, while the effect of gas holdup on the average shear rate can be ignored, its indirect effect on \( \gamma \) is important.

Chisti (1989) stated that correlations which express \( \gamma \) as functions of \( U_{sg} \) alone are incomplete. Since most researchers agree that the liquid flow properties profoundly influence the gas holdup and the liquid velocity in the column, it is illogical to expect the shear rate in the column to be independent of the viscosity (or the flow behaviour index) of the liquid.
Throughout this work Eqn 3.46 was employed to calculate the average shear rate in the airlift reactors with non-Newtonian liquids.

### 3.3 LIQUID MIXING

#### 3.3.1 Liquid Circulation Time Model

By definition, circulation time in an airlift reactor is the time necessary for a fluid element to complete a closed loop. Hence to calculate the circulation time in an airlift reactor, the loop length (L) of the reactor is required. For a given superficial gas velocity, reactor geometry and fluid properties the hydrodynamic model (Section 3.1) can be employed to calculate the riser gas holdup ($\varepsilon_g$) and the riser superficial liquid velocity ($U_{sl}$). Then the riser linear liquid velocity ($V_{ir}$) can be obtained by

$$V_{ir} = \frac{U_{sl}}{1 - \varepsilon_g}$$

Performing a mass balance on the reactor, we can get the downcomer linear liquid velocity ($V_{id}$) in terms of the riser interstitial liquid velocity

$$V_{id} = V_{ir} \left( \frac{A_r}{A_d} \right)$$

The average linear liquid velocity in the reactor is given by

$$V_l = \frac{1}{2} \left( V_{ir} + V_{id} \right)$$

From the loop length of the reactor and the linear liquid velocity, the circulation time in the reactor can be derived as

$$t_c = \frac{2L \left( 1 - \varepsilon_g \right)}{U_{sl} \left( 1 + \frac{A_r}{A_d} \right)}$$

For a given superficial gas velocity, reactor geometry and fluid properties, the theoretical circulation time can be calculated using Eqn 3.50 and the hydrodynamic model.
3.3.2 Axial Dispersion

The calculation of Bodenstein (Bo) number and the axial dispersion coefficient (D) in airlift reactors were also undertaken in this study. On all three configurations of the airlift contactors the overall, the riser and the downcomer Bodenstein numbers and dispersion coefficients were calculated. In this procedure, it was assumed that the bottom section of the airlift was insignificant in terms of liquid mixing (Chisti and Moo-Young, 1993; Fields and Slater, 1983; Merchuk and Yunger, 1990; Verlaan et al., 1989).

For a given superficial gas velocity, airlift reactor configuration and fluid properties, one way of obtaining the overall Bodenstein number for the reactor as a whole is as follows:

For various Bo numbers, the theoretical lines are plotted using the following equation given by Levenspiel (1972); see Section 2.3.1.3

\[ \frac{C}{C_*} = \left( \frac{Bo}{4 \pi \theta} \right)^{\frac{1}{2}} \sum_{\nu=1}^{\infty} \exp \left[ \frac{-Bo}{4 \theta} (x - \theta)^2 \right] \] 3.51

The experimental response curve of \( \frac{C}{C_*} \) versus \( t/t_0 \) is plotted on the same graph as the theoretical lines (Fig 3.8) and the best fitting theoretical line to the experimental data gives the overall Bo number for that particular experiment. Alternatively, Bodenstein numbers can also be obtained by a least squares fit of Eqn 3.51 to the experimental data.

From the overall Bodenstein number, the overall axial dispersion coefficient can be calculated from the following equation (see Section 2.3.1.3):

\[ D = \frac{U_{sl} L}{(1 - \varepsilon_o) Bo} \] 3.52

The change in variance of pulse of tracer due to passage through riser, gas-liquid separator and downcomer can be measured separately (Fig 3.9). The riser Bo number can be obtained, using the change in variance of pulse of tracer due to passage through the riser section, from the following correlation (Fields and Slater, 1983; Levenspiel, 1972):

\[ \Delta \sigma^2_r = \frac{H^2 (1 - \varepsilon_g)^2}{U_{sl}^2} \left[ \frac{2}{Bo} + \frac{8}{Bo^2} \right] \] 3.53

Then the axial dispersion coefficient in the riser section of the airlift can be obtained using the following equation:

\[ D = \frac{H U_{sl}}{(1 - \varepsilon_g) Bo} \] 3.54
Theoretical Modelling

The downcomer Bo number can be obtained from Eqn 3.55, using the change in variance of pulse of tracer due to passage through the downcomer section (Fields and Slater, 1983; Levenspiel, 1972).

\[ \Delta \sigma_d^2 = \frac{H_d^2 (1 - \varepsilon_d)^3}{U_{std}^2} \left[ \frac{2}{Bo} + \frac{8}{Bo^3} \right] \] 3.55

The axial dispersion coefficient in the downcomer is calculated, employing the downcomer Bo number, from the following:

\[ D = \frac{H_d U_{std}}{(1 - \varepsilon_d) Bo} \] 3.56

For a complete circulation of the airlift loop, the overall change in variance (\(\Delta \sigma_o^2\)) of the tracer pulse is (Aris, 1959; Bisschof, 1960):

\[ \Delta \sigma_o^2 = \Delta \sigma_r^2 + \Delta \sigma_d^2 + \Delta \sigma_i^2 \] 3.57

where \(\Delta \sigma_i^2\) is the change in the variance of the tracer pulse due to passage through the gas-liquid separator. Another way to obtain the overall Bodensten number is to obtain \(\Delta \sigma_o^2\) by using equation 3.57. Then the overall Bo number for the airlift reactor can be attained from the following equation (Levenspiel, 1972):

\[ \frac{\Delta \sigma_o^2}{t_c^2} = \frac{2}{Bo} + \frac{8}{Bo^3} \] 3.58

This method has been previously employed by a number of investigators (Fields and Slater, 1983; Lu et al., 1994a; Verlaan et al., 1989).

In this study, the overall Bo numbers were obtained in the airlift contactors by both methods (Eqn 3.51 and 3.58). Only the overall Bodensten numbers obtained using Eqn 3.58 are given in Section 5. A comparison between the overall Bodensten numbers obtained using Eqn 3.51 and that using Eqn 3.58 is made in Section 5.1.3. The riser and the downcomer Bo numbers were procured employing equations 3.53 and 3.55, respectively.
The change in the variance of a pulse of tracer through, for example, the riser section of an airlift vessel can be obtained by having two probes a known distance apart in the riser (Fig 3.9). The variance of the injected tracer pulse is measured at probes 1 and 2 (Fig 3.9); then subtracting the variance at probe 1 from that of probe 2 gives the change in the variance between probes 1 and 2.

Example: The calculation of a change in variance of a pulse of tracer through the riser section (External loop airlift; liquid is tap water; \( U_{sg} = 0.0169 \text{ m/s} \)). The details of the experimental tracer curve (at 5 sec time interval) at probes 1 and 2 (Fig. 3.9) are given in the table below.

<table>
<thead>
<tr>
<th>t (s)</th>
<th>( C (\text{kg/m}^3) )</th>
<th>( C_e (\text{kg/m}^3) )</th>
<th>( C/C_e )</th>
<th>( t^2 (\text{s}^2) )</th>
<th>( tC/C_e (\text{s}) )</th>
<th>( t^2C/C_e (\text{s}^2) )</th>
</tr>
</thead>
<tbody>
<tr>
<td>Probe 1</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>0</td>
<td>0</td>
<td>13</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>5</td>
<td>256</td>
<td>13</td>
<td>19.6923</td>
<td>25</td>
<td>98.4615</td>
<td>492.308</td>
</tr>
<tr>
<td>10</td>
<td>3</td>
<td>13</td>
<td>0.23077</td>
<td>100</td>
<td>2.30769</td>
<td>23.0769</td>
</tr>
<tr>
<td>15</td>
<td>0</td>
<td>13</td>
<td>0</td>
<td>225</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>Probe 2</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>0</td>
<td>0</td>
<td>16</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>5</td>
<td>55</td>
<td>16</td>
<td>3.4375</td>
<td>25</td>
<td>17.1875</td>
<td>85.9375</td>
</tr>
<tr>
<td>10</td>
<td>50</td>
<td>16</td>
<td>3.125</td>
<td>100</td>
<td>31.25</td>
<td>312.5</td>
</tr>
<tr>
<td>15</td>
<td>7</td>
<td>16</td>
<td>0.4375</td>
<td>225</td>
<td>6.5625</td>
<td>98.4375</td>
</tr>
</tbody>
</table>

The equation to calculate the variance of pulse of tracer at probe 1 is

\[
\text{Variance (1)} = \frac{\sum t^2 \frac{C}{C_e}}{\sum \frac{C}{C_e}} - \left( \frac{\sum t \frac{C}{C_e}}{\sum \frac{C}{C_e}} \right)^2 = 0.28622 \text{ s}^2 \quad \text{(using the data given in the table above)}
\]

Similarly, the variance of pulse of tracer at probe 2 is

\[
\text{Variance (2)} = \frac{\sum t^2 \frac{C}{C_e}}{\sum \frac{C}{C_e}} - \left( \frac{\sum t \frac{C}{C_e}}{\sum \frac{C}{C_e}} \right)^2 = 9.24745 \text{ s}^2
\]

Therefore, the change in variance of a pulse of tracer through the section is

\[
\text{Variance (2)} - \text{Variance (1)} = 9.24745 - 0.28622 = 8.96123 \text{ s}^2
\]

This value is then used in Eqn 3.53 together with the riser gas holdup, superficial liquid velocity and liquid height to obtain the riser Bo number for a particular gas flow rate.
Figure 3.8: The comparison between the experimental data and the curve generated by Equation 3.51 for the value of overall Bodenstein number which gives the closest fit between experimental data and theory.

Figure 3.9: The measurement of a change in variance of pulse of tracer in the riser, gas-liquid separator and downcomer by the four thermocouples in the internal or external loop airlift reactor.
4 EXPERIMENTAL WORK

4.1 The Airlift Reactors

In order for useful research, the laboratory scale airlift reactor should reproduce closely the hydrodynamic behaviour expected to be encountered on larger operational scale. Chisti (1989) warned that in very small vessels, entrance and exit effects, surface aeration and wall effects can strongly affect the reactor performance. In many of the previous studies with laboratory scale airlift contactors, the working volume of the vessels were between 0.01 and 0.09 m$^3$ (Fields and Slater, 1983; Jones, 1985; Kawase and Moo-Young, 1986a; Kawase et al., 1994; Popovic and Robinson, 1993; Wachi et al., 1991; Wu and Jong, 1994).

Another important consideration in laboratory scale airlift contactors is the need for visual observations. The vessel must be susceptible to visual inspection over as much of its volume as possible. For many hydrodynamic effects there is no better way than flow visualization of determining how well the contactor is performing (Chisti, 1989). Hence, usually airlift reactors have been constructed either from plexiglass or glass or transparent acrylic plastic (Choi et al., 1995; Glennon et al., 1993; Li et al., 1993; Lu et al., 1994a; Muller and Davidson, 1992; Philip et al., 1990; Popovic and Robinson, 1988; Vatai and Tekic, 1986; Young et al., 1991).

Both of the above points were taken into account in the construction of the airlift contactors used in this study. In this work, we have employed three different laboratory scale airlift devices. These include two concentric draft-tube internal loop airlift reactors and an external loop airlift contactor. Further details of the reactors are given below.

4.1.1 Internal Loop Airlift Devices

The concentric draught-tube internal loop airlift contactor was made of plexiglass and had a total height of 2.26 m. The entire vessel was made out of plexiglass to permit visual inspection of the flow regime, bubble size and the dispersion level of the liquid. The total contactor volume was 0.052 m$^3$. The column had an outer diameter of 0.15 m. Two concentric draft-tubes with different diameters were employed. The first draft-tube had an inner diameter of 0.103 m [which produced the internal loop airlift (1) contactor] and the other one had an inner diameter of 0.114 m [which gave the internal loop airlift (2) vessel]. This produced a downcomer-to-riser cross-sectional area ratios ($A_d/A_r$) of 0.92 and 0.49,
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respectively. Both draft-tubes had a height of 1.725 m and were located 0.095 m above the base of the vessel. The unaerated liquid height was kept at 1.94 m. The internal loop airlift (1) contactor is shown schematically in Figure 4.1.

The draft-tube (the riser) was sparged by a 0.05 m diameter ring sparger, which contained 12 equally spaced holes of 1 mm diameter. The sparger was placed just (5 mm) inside the base of the draft-tube [after Chisti and Moo-Young’s (1987) observations]. The bottom section of the airlift vessel was specifically designed to keep the frictional losses, due to a change in the flow direction (at the bottom), to a minimum. Further geometric details of the two contactors are given in Table 4.1.

Figure 4.1: The schematic diagram of the concentric draft-tube internal loop airlift (1) vessel. For clarity, only one set of the manometer probes and thermocouples are shown.
4.1.2 External Loop Airlift Device

The external loop airlift contactor consisted of a circular riser (dia. of 0.095 m) connected to a downcomer (dia. of 0.06 m) as shown schematically in Figure 4.2. Gas-liquid separation took place, at the top of the reactor, in a relatively large rectangular tank. The downcomer-to-riser cross-sectional area ratio ($A_d/A_r$) was 0.40.

The vessel was made of plexiglass and it had a total height of 3.16 m. The contactor volume was 0.035 m$^3$. In all the experiments a static liquid height (measured from the base of the riser) of 2.45 m was used. Other geometric details of the device is given in Table 4.1. A ring sparger (dia. of 4.5 cm) with twelve equally spaced holes of 1 mm diameter, located at the base of the riser, was used for aeration.

A photo of the internal loop airlift and the external loop airlift contactors is shown in Plate 4.1.

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Figure 4.2: The schematic drawing of the external loop airlift device. For clarity, only one set of the manometer probes and thermocouples are shown.
Plate 4.1: A photo of the internal and external lead lift contactors used in the study.
4.1.3 The Experimental Setup

A schematic diagram of the experimental setup is shown in Figure 4.3. For any experiment, the liquids were initially placed into the agitator. The agitator had a working volume of 0.1 m$^3$. After the desired concentration of the liquid had been achieved (see Section 4.3), the fluid was transferred to the by-pass tank. The by-pass tank behaved as the storage tank in this setup and had a working volume of 0.25 m$^3$. The liquid was then pumped up (using a Stuart Turner centrifugal pump) into either the internal or the external loop airlift contactor. The airlift reactors were emptied using the valves at the bottom of the reactors.

The gas (air) passed through a rotameter (Rotameter MFG Co. Ltd., UK) before entering the bottom of the reactor via a sparger. More details about the gas flow setup is given in Section 4.2.

Figure 4.3: Schematic drawing of the experimental setup.
<table>
<thead>
<tr>
<th>Reactor property</th>
<th>External loop airlift</th>
<th>Internal loop airlift (1)</th>
<th>Internal loop airlift (2)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Total height (m)</td>
<td>3.20</td>
<td>2.26</td>
<td>2.26</td>
</tr>
<tr>
<td>Working volume (m$^3$)</td>
<td>0.035</td>
<td>0.052</td>
<td>0.052</td>
</tr>
<tr>
<td>Unaerated liquid height (m)</td>
<td>2.45</td>
<td>1.94</td>
<td>1.94</td>
</tr>
<tr>
<td>Column diameter (m)</td>
<td>-</td>
<td>0.150</td>
<td>0.150</td>
</tr>
<tr>
<td>Riser diameter (m)</td>
<td>0.095</td>
<td>0.103</td>
<td>0.114</td>
</tr>
<tr>
<td>Downcomer diameter (m)</td>
<td>0.06</td>
<td>0.047</td>
<td>0.036</td>
</tr>
<tr>
<td>Height of draft-tube (m)</td>
<td>-</td>
<td>1.725</td>
<td>1.725</td>
</tr>
<tr>
<td>Gas-liquid separator; length, width &amp; height (m)</td>
<td>0.40, 0.15 &amp; 0.40</td>
<td>Top: 0.315, 0.315 &amp; 0.45</td>
<td>Top: 0.315, 0.315 &amp; 0.45</td>
</tr>
<tr>
<td>Aspect ratio, H/d&lt;sub&gt;c&lt;/sub&gt; (H/T)</td>
<td>-</td>
<td>12.9</td>
<td>12.9</td>
</tr>
<tr>
<td>A&lt;sub&gt;d&lt;/sub&gt;/A&lt;sub&gt;r&lt;/sub&gt; (A&lt;sub&gt;e&lt;/sub&gt;/A&lt;sub&gt;d&lt;/sub&gt;)</td>
<td>0.40 (2.50)</td>
<td>0.92 (1.09)</td>
<td>0.49 (2.04)</td>
</tr>
<tr>
<td>Riser-to-column diameter (T/d&lt;sub&gt;c&lt;/sub&gt;)</td>
<td>-</td>
<td>0.69</td>
<td>0.76</td>
</tr>
<tr>
<td>Draft-tube length-to-column diameter (L&lt;sub&gt;d&lt;/sub&gt;/d&lt;sub&gt;c&lt;/sub&gt;)</td>
<td>-</td>
<td>11.5</td>
<td>11.5</td>
</tr>
</tbody>
</table>

Table 4.1: Geometric details of the airlift reactors.
4.2 The Gas Flow Setup

Air from approximately 586 Pa mains routed via a pressure regulator, a filter, a flow control valve and a rotameter setup (Rotameter MFG Co. Ltd., UK), was supplied to the airlift contactors at ambient temperature (Fig 4.4). Throughout, the air entered the airlifts near its saturation humidity. There was negligible water loss from the airlift devices even at extended aeration (three days) at relatively high gas flow rates (up to 13 L/min.).

All experiments were conducted with continuous air flow and batchwise for the liquid. The airlift rigs were operated at room temperature and atmospheric pressure. The airlift vessels were aligned with the vertical line to within better than ±0.05° to avoid the possible detrimental effects (on mixing, especially on the axial dispersion coefficient) of small deviations from the vertical that have been stated by previous researchers (Ityokumbul, 1993; Rice et al., 1990; Rice and Littlefield, 1987; Tinge and Drinkenburg, 1986; Valdes-Krieg et al., 1975).

![Diagram of the air flow line to the airlift devices.](image)

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1. Air
2. Ball valve
3. Filter
4. Pressure regulator
5. Rotameter
6. Control valve
7. Airlift reactor

Not to scale
4.3 The Fluids Used

Most biofluids of interest can be simulated by model media with respect to their rheological properties and hydrodynamic behaviour. The fluid used for the simulation should be easy to formulate in a reproducible way, must be safe, should not be expensive and has to be sufficiently stable over the expected duration of experiments at the required conditions.

Newtonian (water-like) bacterial and yeast fermentations are satisfactorily duplicated by tap water and aqueous salt solutions (Snape et al., 1992; Verlaan et al., 1989; Young et al., 1991). Wastewater can be modelled using tap water and phenol (Fan, 1989). More viscous Newtonian media can be simulated with aqueous solutions of sugar, glycerol and glycol (Philip et al., 1990; Shi et al., 1990). Non-Newtonian fluids may be modelled with aqueous solutions of polymers such as carboxymethyl cellulose (CMC), polyacrylamide and xanthan gum (Kawase et al., 1994; Popovic and Robinson, 1993; Shi et al., 1990). Mycelial slurries can be duplicated with aqueous suspensions of fibrous solids (e.g., Solka-Floc cellulose fibre) with or without some dissolved polymer (Chisti, 1989).

4.3.1 Newtonian Liquids

The following liquids were employed in this study:

*Tap water* - relatively hard. London tap water is mostly drawn from the Thames and Lee rivers and partly from groundwater sources. The water undergoes long periods of reservoir storage and has quite steady properties. Usual concentrations are (Gregory, 1994):
- Total dissolved solids = 450 mg/L; total hardness = 264 mg/L; non-carbonate hardness = 59 mg/L; turbidity = 4 units; chlorides = 25 mg/L; pH = 8.1.

*Aqueous salt solution* - Sodium chloride (BDH Chemicals, UK) in tap water. Aqueous salt solutions were prepared by adding the required amount of salt (NaCl) to tap water to give either 0.1 M, 0.2 M or 0.5 M concentration of salt solution. Since for a given ionic strength solution, neither the type of inorganic ion nor the ionic charge influence bubble coalescence behaviour (Hikita et al., 1980), an inexpensive, readily available and safe salt (sodium chloride) was chosen.

The use of aqueous salt solutions as simulation for some bacterial and yeast fermentation systems is well known (Chisti, 1989) and it has been employed before (Hikita et al., 1980; Kelkar et al., 1983; Snape et al., 1992). Most bacterial and yeast fermentation broths are Newtonian, have viscosity and density close to that of water, and also have an ionic strength. Hence salt solutions were used to reproduce the ionic strength of
fermentation systems, thereby providing a close simulation of the salt associated coalescent behaviour of the real fluids and also of any possible salt effects on the gas holdup.

*Aqueous glycerol solution* - Glycerol \([\text{CH}_3(\text{CH}_2\text{OH})_2\text{CH}_3]\) in tap water. Aqueous glycerol solutions were prepared by adding tap water to undiluted glycerol to obtain the required concentration (i.e. viscosity).

To model relatively viscous Newtonian media aqueous solutions of glycerol were selected. Glycerol is relatively cheap, not toxic, easily available and is adequately stable. The glycerol employed in this work, obtained from BDH Chemicals, had an average initial viscosity of 1.300 Pa s.

*Food-grade oil* - Soyabean oil (Lupa vegetable oil, A. Donatantonio & Sons Ltd., UK). Pure soyabean oil has a density much lower than that of water. It was relatively inexpensive and was also easily available.

### 4.3.2 Non-Newtonian Liquids

*Aqueous carboxymethyl cellulose solution* - Carboxymethyl cellulose (CMC) powder in tap water. The CMC solutions were prepared by dissolving CMC powder (BDH Chemicals, UK) in tap water. Tap water was initially heated to about 50 °C in the and then the CMC powder was added to the hot water; at this point the heater was switched off. The mixture was continuously stirred for approximately 6 hours, while it cooled from 50 °C to room temperature.

To simulate non-Newtonian, time independent, pseudoplastic fluids, we chose aqueous solutions of CMC. Many investigators (Chisti, 1989; Haque *et al.*, 1986; Kawase and Moo-Young, 1987a; Philip *et al.*, 1990) have used CMC solutions in their work to model non-Newtonian media. The rheological properties were measured with a coaxial cylinder viscometer (Section 4.4.1).

*Oil-water emulsion* - Soyabean oil (Lupa vegetable oil, A. Donatantonio & Sons Ltd., UK) in tap water. Four different concentrations of oil-in-water mixtures were used ranging from 10 to 60 vol%.

Pal *et al.* (1986) have employed oil-in-water concentrations varying from 40 to 95 vol% and found that the oil-in-water emulsions behaved as time independent non-Newtonian fluids at higher dispersed phase (oil) concentrations. They stated that emulsion viscosity increased sharply with an increase in the volume concentration of the dispersed phase. Pal *et al.* (1986) stabilised their emulsions with a non-ionic emulsifier: Triton x-100.
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With this emulsifier they found that the highest concentration that could be obtained without the emulsion becoming unstable was 84 vol%.

In this study, the oil-in-water emulsions were prepared as follows:

An emulsifier (Sodium Dodecyl Sulphate, CH_{12}H_{25}OSO_{3}Na, 288.38 g/mol; BDH Chemicals, UK) was initially dissolved in water (continuous phase) and then the oil (dispersed phase) was slowly added and stirred (in the agitator) to obtain an emulsion of two immiscible liquids. A concentration of 0.3% SDS by volume of emulsifier based on the amount of water present in the emulsion was used. It was discovered that with 0.3% SDS vol., 60% was the highest oil-in-water fraction which can be obtained without the emulsion becoming unstable.

4.4 Measurement of Liquid Physical Properties

Before and after each experiment the physical properties of the liquid in the airlift reactor were measured. Temperature, density and viscosity of the liquid and the unaerated liquid height in the column were the properties that were ascertained.

The temperature and the density of the liquids were obtained using a thermometer and a hydrometer, respectively. It was discovered that the average temperature of the reactor increased by approximately 2 °C, before and after the experiments, due to the addition of hot liquid to the reactor for the purposes of liquid velocity and mixing measurements (see Section 4.6). An average of the temperature before and after the experiment was taken as the temperature of the reactor for that experiment. There was a slight decrease (0.5 - 2.0 kg/m$^3$) in the density of the liquid (before and after the experiment), when the liquid employed was glycerol solution. With the other liquids, no apparent change in the density occurred. It should be noted that glycerol solution, especially at high concentrations, is very sensitive to changes in temperature. An average of the density values before and after the experiment was obtained and used as the density of the liquid for that experiment. The measurement of viscosity is considered below (Section 4.4.1). The unaerated liquid height in the reactor was also measured before and after the experiment, since liquid was being added (typically 1400 to 1800 cm$^3$) to the reactor during the experiment. Then an average of the liquid heights was taken to be the unaerated liquid height for that experiment.
Each experiment in the airlift contactor was carried out three times. All the experiments yielded reproducible results (to within ± 20%). However, the results of the three individual experiments were averaged to give a representative value.

4.4.1 Measurement of Liquid Rheology
For all the liquids employed in this study, apart from glycerol-water solutions, a concentric cylinder viscometer (Section 2.6.1) was used for the measurement of liquid viscosity. The rheometer employed was a Rheomat 115 rotational viscometer (Contraves AG, Switzerland) with a module 77 operating system and a concentric cylinder (MS-0/115) measuring unit. Warren (1994) has given more details of the Rheomat 115 viscometer. Measurements were made at room temperature and atmospheric pressure.

The measuring cup was filled with about 20 mL of liquid (i.e. up to the fill line) and the measuring bob was put in place, ensuring that no gas remained in the fluid. The measuring bob was rotated from step 1 to 15 (increasing speed) in steps of 1. The measurement readings for each speed step were converted into shear stress values using the table supplied by Contraves AG. The shear rate step range was from 24.3 (step 1) to 3680 s⁻¹ (step 15). The values of shear stress and shear rate were then used in the power law relationship (Eqn 2.20) to determine the rheological behaviour of the sample.

Due to the relatively high viscosity (and density) of the glycerol-water solutions, the measurement of viscosity of these solutions could not be performed in the Rheometer 115 viscometer. Hence, tables relating the density of glycerol to viscosity at a certain temperature were employed (Bosart and Snoddy, 1928; Sheeley, 1932).

4.5 Gas Holdup Measuring Procedure
Gas holdup measuring methods have been already discussed in Section 2.3.1.1. Since the determination of the individual riser (ε_r) and downcomer (ε_d) gas holdups were required, the manometric probe technique was chosen. Inverted U-tube manometers were used and were filled with the suspending fluid used in the devices. Manometer probes were connected to pressure taps located at two different axial positions either in the riser or in the downcomer of the devices for ε_r and ε_d measurements, respectively. The pressure tappings consisted of 1 mm holes bored into the wall of the airlift device (flush with inner wall). Chisti (1989) reported that the contribution of any accelerational effects to the measured pressure was negligible. He showed a close agreement (± 10%) between the
holdups measured by the manometric method and by the volume expansion technique for a number of different reactor configurations. Other investigators have also observed similar results with the manometric method (Bello, 1981; Merchuk, 1986). The manometer probes were connected to the U-tube manometers via PVC tubing (3 mm dia.). Two manometer probes were attached to the riser section of the airlift contactors and another two probes were connected to the downcomer section. All the probes had a diameter of 1 mm and a length of 7.8 cm. The manometer probes were placed 1.50 m apart (in the riser and downcomer) in the external loop airlift contactor, while in the internal loop airlift there were situated 1.35 m (riser and downcomer) apart. In the external loop airlift, the protruding lengths of the probes into the riser and downcomer sections were 4.0 and 3.0 cm, respectively. In the internal loop airlift, the protruding lengths of the probes in the riser and downcomer were 5.0 and 1.5 cm, respectively.

It is essential that all the air bubbles are removed from the connecting tubes and the manometer before starting the gas flow rate. The presence of even one air bubble in the PVC tubing will give an erroneous reading. Hence it is important that, before and during an experiment, regular checks are made to see whether there are any air bubbles in the PVC tubing. The gas holdup can be calculated from Eqn 2.6. In this equation, dh is the difference between the liquid levels in the manometer and dz is the length between the two pressure tap points (see Fig 4.5).

A differential pressure transducer (Newport Ltd., UK) was also employed in the present study to measure gas holdup. It was connected to the airlift’s pressure tappings via PVC tubing (3 mm dia.). The transducer was only used to double check the pressure readings obtained by the manometers. There was good agreement (to within ± 10%) between the pressure transducer readings and the manometer values.
4.6 Liquid Velocity and Mixing Measurement Methods

From the various liquid velocity and mixing measurement systems (Section 2.3.1.2), we chose the thermal technique for determining the liquid circulation velocity, the mixing time, the circulation time and the axial dispersion coefficient in the airlift contactors. Numerous investigators (Douek and Livingston, 1994; Ford et al., 1972; Hoogendoorn and Hartog, 1967) have utilised this method. The main advantage of this method was that an unlimited number of experiments can be performed with the same liquid. Since in this study relatively large scale experiments were performed, the thermal technique proved to be a relatively inexpensive method.

A reservoir tank (volume of 0.03 m³) was filled with the liquid (same as in the airlift reactor) and the liquid in the tank was heated up to a preset temperature (usually 65.0 - 68.0 °C). The liquid pulse injections were made possible by electronically operated solenoid valves; the injection time could be varied (from 0.5 to 5.0 secs). Usually we used an injection time of 1.5 secs. This corresponded to a volume of 120-150 cm³ of pulse (hot liquid) injection. At any one time, the reservoir tank was connected to either the internal loop or the external loop airlift device (via a 0.025 m diameter pipe).

Figure 4.5: The inverted U-tube manometer arrangement. Redrawn from Chisti (1989).
Thermocouple probes were fixed to the airlift devices as shown in Figures 4.1 and 4.2 (for clarity, only the thermocouples in the riser section are shown). Two probes were placed in the riser section of the airlifts and another two probes were located in the downcomer section. All the probes had a diameter of 5 mm and a length of 8.3 cm. The diameter of the tip of the probe was 0.5 mm. In the external loop, the protruding lengths of the thermocouples into the riser and downcomer sections were 4.0 cm and 3.0 cm, respectively. The protruding lengths of the thermocouples in the riser and downcomer, in the internal loop, were 5.0 cm and 1.5 cm, respectively. On the internal loop airlift vessel the distance between the two thermocouple probes (both in the riser and downcomer) was 1.35 m. The thermocouple probes were placed 1.50 m apart (riser and downcomer) on the external loop airlift device. Each of the probes were connected to a liquid injection system (LIS). The LIS was attached to a chart recorder (two pen). A typical response curve, obtained with two thermocouples, is shown in Figure 4.6.

In the internal loop airlift vessel the pulse of hot liquid was injected into the riser section of the airlift (Fig 4.1), above the sparger. The hot liquid was injected at the bottom of the external loop airlift contactor, below the sparger, as shown in Fig 4.2.

The determination of the liquid circulation velocity involved injecting a pulse of hot liquid into the flowing liquid (in the airlift device) and measuring the time taken by the pulse to travel between a known distance. A knowledge of the time and the distance can lead to the calculation of the velocity of the flowing liquid in the airlift vessel. From the response curves the mixing and circulation times can also be obtained as shown in Fig 4.6. Furthermore, the experimental tracer response curve is also used to calculate the Bodenstein number and the axial dispersion coefficient, for that particular system, as outlined in Section 3.3.2.

When a pulse of hot liquid is injected into the column, we obtain on the chart recorder a trace of the circulation of the pulse (Fig 4.6). Knowing the chart speed \(U_{cr}\), the height between the two probes (either on the internal loop or the external loop), \(h_p\), and from the response curve obtaining the distance between two consecutive peaks on the chart \(X_{cr}\), the liquid velocity between the two thermocouple probes could be calculated:

\[
U_{vl} = \frac{U_{cr} h_p}{X_{cr}}
\]
Figure 4.6: A typical response curve from the two thermocouples in the riser section of the airlift device. How the liquid circulation velocity (the distance between two consecutive peaks on the chart, $X_c$), the mixing time ($t_m$) and the circulation time ($t_c$) were procured are shown. The 90% mixing time is given.
5 RESULTS AND DISCUSSION

This section presents and discusses the results of this study. It is divided into six parts. The first part is concerned with the effect of the superficial gas velocity on the hydrodynamics (namely gas holdup and liquid velocity) and liquid mixing of airlift contactors. The next two sections examine the influence of liquid physical properties on the airlift hydrodynamics and mixing. The fourth section presents the effect of liquid-liquid systems on the hydrodynamics and mixing. The fifth section compares the experimental gas holdup and liquid velocity data with model predictions. It also compares the experimental liquid circulation time values with predicted values. The final section investigates the shear rate in airlift reactors with non-Newtonian liquids.

5.1 The Effect of Superficial Gas Velocity on Airlift Hydrodynamics and Mixing

In airlift reactor operation, the principal way of controlling the hydrodynamics in the vessel is through the manipulation of the gas (air) flow rate into the reactor. Hence it is not surprising that, in previous literature, the main indicators of reactor performance (namely gas holdup, liquid velocity, mixing time and the liquid phase axial dispersion coefficient) have usually been correlated with the gas flowrate in airlift devices (Chisti and Moo-Young, 1987). An understanding of the relationship between gas throughput and the hydrodynamics is therefore essential for optimal reactor operation and performance.

Throughout this study, the superficial gas velocity \((U_{sg})\) in an airlift is defined as the volumetric air flow rate passing through the gas sparger divided by the riser cross-sectional area. Chisti and Moo-Young (1987) have warned that only the superficial gas velocity based on the riser cross-section has any real physical meaning in airlift contactors. In this work, Equation 2.4 was used to calculate the superficial gas velocity. In this equation, the temperature (average) of the fluid inside the airlift vessel was taken to be the temperature of the reactor and the cross-sectional area was taken to be the riser cross-sectional area. Atmospheric pressure was taken to be the headspace pressure. It must be remembered that the superficial gas velocity is not the real velocity of the gas, but is simply a convenient way of relating the gas throughput to the scale of the reactor.

Three different geometries of airlifts were employed in this study (Table 4.1). Throughout this work, air was used as the gas phase (Section 4.2). Twenty different Newtonian and non-Newtonian liquids were also employed in this study (Section 4.3).
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Tables A.1 to A.3 (Appendix 1) give the physical properties of the various liquids used in the three airlift contactors.

An air flow rate of 1.5 to 11.0 L/min. (25.1 - 183.7 cm³/s) was employed in this study with all three airlift contactors. This gave a superficial gas velocity in the range of 2.0 to 24.0 mm/s depending on the type of airlift vessel used and the properties of the fluids.

Three different types of flow regimes mentioned in Section 2.2.3 were observed in the airlifts with the various liquids. Although only low to intermediate superficial gas velocities were used in this study (below 5 cm/s), but churn turbulent flow regime was still observed in viscous Newtonian and non-Newtonian liquids. Many researchers (Chisti, 1989; Onken and Weiland, 1983; Pollard, 1995; Russell, 1989) have reported that the churn turbulent flow regime in airlift vessels only occurs over a superficial gas velocity of about 5 cm/s. However, Chisti and Moo-Young (1987) suggested that earlier transition to slug flow is more likely to occur in highly viscous fluids and mycelial media than in water-like systems. Joshi et al. (1990) also stated that the transition point from the homogeneous to heterogeneous regime is dependent on several factors such as superficial liquid velocity, diameter of the column, liquid viscosity, surface tension and even the sparger design.

With water (and with other liquids having a viscosity similar to that of water i.e. all three salt solutions and 48.5% and 57.5% glycerol solutions), only the homogeneous bubbly flow and coalesced bubble flow regimes were observed in the range of superficial gas velocities used in this study. Chisti and Moo-Young (1987) suggested that a transitional flow regime known as ‘coalesced bubble flow’ occurred between the homogeneous bubbly flow regime and churn turbulent flow regime. In this study, from visual observations, the transitional flow regime started above a superficial gas velocity of about 12.7 mm/s. This regime was characterised by increased bubble density and greater turbulence. The above observation was made in all three geometries of airlift devices.

With more viscous liquids (i.e. 71.5% to 86.0% glycerol solutions, the five CMC solutions and soyabean oil) the transition from bubbly flow to coalesced bubble flow regime was observed to occur above a superficial gas velocity of around 4.3 mm/s. The transition from coalesced bubble flow regime to churn turbulent flow regime occurred over a superficial gas velocity of about 12.8 mm/s. This heterogeneous flow regime was characterised by large bubbles having a high rise velocity in the presence of small bubbles. It must be pointed out that with the four oil-in-water emulsions no observations of the flow
Results and Discussion

regimes in the column were possible due to the opaque (milky white) nature of the emulsions.

It seems that for a given superficial gas velocity, changing the viscosity of the liquid in the column can also change the flow regime in the column. For example in the external loop airlift contactor, for a superficial gas velocity of 2.1 cm/s, coalesced bubble flow regime was observed with water. However, for the same gas velocity in the external loop airlift with 86.0% glycerol solution churn turbulent flow regime was observed.

From the above it is clear that the type of flow regime in the airlift columns was not just dependent on the superficial gas velocity but was also influenced by the properties of the liquids. This is in agreement with previous researchers’ (Chisti and Moo-Young, 1987; Joshi et al., 1990) observations that the type of flow regime in the reactor was dependent on the viscosity of the liquid in the reactor.

From visual observations, it was noticed that with water-like fluids at very low gas velocities small bubbles rose up in the riser and the downcomer section was usually free of any gas bubbles. As the gas flow rate increased, small bubbles were entrained by the liquid circulation into the downcomer. The penetration depth of these bubbles into the downcomer depended on their size and some recirculation of the small bubbles occurred. As the gas flow rate was further increased, larger bubbles were also pulled into the downcomer section. With viscous fluids (both Newtonian and non-Newtonian), even at very low gas flow rates bubble coalescence in the riser section led to the formation of large bubbles. A large number of small bubbles were also present in the downcomer and many of them recirculated. As the gas velocity increased, some of the large bubbles were dragged into the downcomer section.

5.1.1 The Effect of Superficial Gas Velocity on Gas Holdup

The gas void fraction varies throughout the airlift vessel and can be divided into the overall, riser and downcomer gas holdups (Section 2.3.1.1). The riser and downcomer gas holdups can be considered to be the mean gas holdups in the riser and the downcomer sections, respectively. In this study, the riser and downcomer gas holdups were measured (Section 4.5).

It has been reported (Philip et al., 1990; Schugerl, 1981) that, for a given gas velocity, the local and the overall gas holdups vary with time in airlift contactors containing viscous Newtonian and non-Newtonian liquids due to the accumulation of small
bubbles in viscous liquids. Hence, in this work, experiments were carried out to determine the effect of time on the local gas holdup. With the 20 different liquids, for a given gas velocity, the riser and the downcomer holdups were measured over a period of three hours. Figure 5.1 shows the results obtained with tap water in the external loop airlift device. With water, there was no noticeable change in the riser (or downcomer) gas holdup over the three hour period. The same effect was also observed with the three salt solutions and the four oil-in-water emulsions. However, with glycerol solution (71.5%) for a given gas flow rate the riser (or downcomer) gas holdup increased to a maximum (at about 30 mins) and then decreased to a stable value over the three hour period (Fig 5.2). Usually, depending on the aeration rate and liquid viscosity, the stable value of holdup was achieved after an hour of starting the gas flow. Similarly with CMC solutions and soyabean oil the gas holdup was observed to reach a stable value only after about an hour, depending on gas flow rate and liquid viscosity. The variation of riser gas holdup with time for 0.4% CMC solution in the external loop airlift device is given in Figure 5.3. It is interesting to note that, again the maximum value of holdup occurred at about 30 minutes after the aeration was started. In summary, it can be stated that with viscous Newtonian and non-Newtonian liquids a stable value of riser (or downcomer) gas holdup was only achieved after about an hour of aeration. In this study, only the stable values of riser and downcomer gas holdups were recorded.

Philip et al. (1990) reported that the overall gas holdup in airlift vessels with viscous Newtonian and non-Newtonian liquids increased gradually to a steady state value over a period of about an hour after the gas flow was turned on. They attributed this phenomenon to the two class of bubbles found in airlift columns containing viscous Newtonian and non-Newtonian liquids. These were large bubbles with ellipsoidal caps or slugs having an equivalent diameter of 20 mm and nearly spherical small bubbles with diameters in the range of 100 μm to 1 mm. They stated that when the gas flow was switched on, the immediate rise in the gas holdup was due to the large bubbles and that there was a negligible amount of small bubbles. Then with an increase in aeration time, there was a progressive increase in the concentration of small bubbles. Some of the small bubbles coalesce and leave the system. They further reported that the rate of coalescence and hence of disengagement being dependent on the concentration of small bubbles, eventually the system attains equilibrium with the rate of generation. The gas holdup is then constant and has attained a stable value.
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In the present study, visual observation (supported by photography) of the two
phase flow structure in the airlift contactors with viscous Newtonian and non-Newtonian
liquids confirmed Philip et al.’s (1990) observations. Plates 5.1 to 5.4 show the two phase
flow structure in the riser section of the external loop airlift contactor with 78.5% glycerol
solution.

Plate 5.1 shows the bubbles in 78.5% glycerol solution at a superficial gas velocity
of 2.6 mm/s. The photo was taken at a distance of 1 cm above the base of the riser section.
The scale (in centimeters) is shown on the photo. The photo was taken almost immediately
after the gas flow was turned on. It indicates that there was an insignificant amount of
small bubbles in the liquid. Generally, the diameters of the bubbles were in the range of 0.5
- 1.0 cm. Plate 5.2 illustrates the bubbles in the riser at a superficial gas velocity of 6.2
mm/s. The photo was taken almost immediately after the gas flow was switched on and at a
distance of 74 cm above the base of the riser section. It is evident that very small bubbles
are almost absent.

![Figure 5.1: Variation in riser gas holdup with time of aeration in water (external loop airlift
contactor).](image-url)
Results and Discussion

Figure 5.2: Variation in riser gas holdup with time of aeration in 71.5% glycerol solution (external loop airlift reactor).

Figure 5.3: Variation in riser gas holdup with time of aeration in 0.4% CMC solution (external loop airlift vessel).
Plate 5.3 depicts the two phase flow structure in 78.5% glycerol solution at a superficial gas velocity of 10.4 mm/s. The photo was taken at a distance of 74 cm above the base of the riser. Plate 5.4 portrays the two phase flow structure at a superficial gas velocity of 16.6 mm/s. The photo was again taken at a distance of 74 cm above the base of the riser. Plates 5.3 and 5.4 were taken about 10 mins after the gas flow was switched on. They illustrate two classes of bubbles: large bubbles (diameters from 1 mm to 30 mm) and very small bubbles (diameters up to 1 mm).

The variation in the measured riser and downcomer gas holdups in water with superficial gas velocity for internal loop airlift (1), internal loop airlift (2) and external loop airlift is shown in Figure 5.4. An average value of holdup from three separate runs is given. There was a rapid rise (almost linear) in riser holdup with increasing gas flow rate in all three airlift vessels. The downcomer holdup also increased with an increase in the gas velocity. This was also observed with the three salt solutions, 48.5% and 57.5% glycerol solutions, and the four oil-in-water emulsions (10% to 60%). This dependence of local gas holdup on the superficial gas velocity has been described by many previous investigators (Chisti, 1989; Russell, 1989; Siegel et al., 1988).

Similarly, with more viscous Newtonian and non-Newtonian liquids (namely 71.5% to 86.0% glycerol solutions, all five CMC solutions and soyabean oil) the riser and downcomer gas holdups increased with increasing gas velocity. However, there was a diminishing dependence of gas holdup on the superficial gas velocity as the gas velocity was increased (Figs 5.5 and 5.6). In all three airlifts, in this study, a reduction in the rate of increase of gas holdup occurred with increasing superficial gas velocities above about 12.8 mm/s. This was in agreement with the visual observations. Many researchers (Chisti, 1989; Joshi et al., 1990; Onken and Weiland, 1983; Pollard, 1995; Russell, 1989) have stated that this phenomenon was due to a change in the flow regime in the airlift (from coalesced bubble flow to churn turbulent flow). In churn turbulent flow regime, large rapidly rising spherical cap bubbles are formed in the riser. This would increase the mean bubble slip velocity causing a reduction in the gas residence time. Since the bubble rise velocity is inversely proportional to the riser gas holdup the formation of larger bubbles (above the superficial gas velocity of 12.8 mm/s) would reduce the rate of increase of riser gas holdup. Hence, this effect (increasing bubble slip velocity) would counteract the effect of increase in riser holdup with increasing gas velocity.
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Plate 5. Gas bubbles in 78.5% glycerol solution at a gas velocity of 6.2 mm/s.

Plate 6. Gas bubbles in 78.5% glycerol solution at a gas velocity of 6.2 mm/s.
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Figure 3: Air bubbles at a gas velocity of 1.04 m/s in 78.5% glycerol solution.

Figure 4: Air bubbles at a gas velocity of 16.6 m/s in 78.5% glycerol solution.
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The riser gas holdup in the internal loop airlift (2) vessel was considerably higher than in the internal loop airlift (1) contactor at a given gas velocity, while it was almost the same as the riser holdup in the external loop airlift reactor (Figs 5.4 - 5.6). This was mainly due to the considerably higher liquid velocity (almost twice) present in the internal loop airlift (1) contactor compared to the external loop airlift vessel and the internal loop airlift (2) contactor at a given gas velocity (see Section 5.1.2). Onken and Weiland (1983) reported that as the riser liquid velocity increased the riser holdup decreased due to the decrease in the residence time of the bubbles in the riser section. The same was true in the downcomer section.

![Figure 5.4: The effect of superficial gas velocity on the riser and downcomer gas holdups in the internal loop airlift contactor (1), internal loop airlift contactor (2) and the external loop airlift contactor. The liquid is water. The data points with lines depict the riser holdup, while the downcomer holdup is shown by just the symbols.](image)

Figure 5.4: The effect of superficial gas velocity on the riser and downcomer gas holdups in the internal loop airlift contactor (1), internal loop airlift contactor (2) and the external loop airlift contactor. The liquid is water. The data points with lines depict the riser holdup, while the downcomer holdup is shown by just the symbols.

As expected, the riser gas holdup was always higher than the downcomer holdup for a fixed gas flow rate. This difference in gas holdup between the riser and downcomer causes a density difference which is the primary driving force leading to liquid circulation.
in airlift reactors. Therefore, the gas holdup in the downcomer will always be less than that in the riser section for a liquid circulation to occur in airlift vessels (Siegel et al., 1988).

With increasing gas velocity, the absolute difference between the riser and downcomer gas holdups also increases in all three airlift devices (Figs 5.4 - 5.6). Pollard (1995) and Russell (1989) have also reported this phenomenon. Pollard (1995) suggested that this might be due to a change in the flow regime from coalesced bubble flow to churn turbulent flow (above a superficial gas velocity of 0.054 m/s). Since this caused a reduction in the rate of increase of the linear liquid velocity. However, Figure 5.4 shows the difference between the riser and downcomer holdups increases in water even when only bubbly flow and coalesced bubble flow regimes were present (below a superficial gas velocity of 0.054 m/s). This increase in the difference between the riser and downcomer gas holdups could also be due to a change in the flow regime from bubbly to coalesced bubble flow.

Figure 5.5: The effect of superficial gas velocity on the riser and downcomer gas holdups in the internal loop airlift contactor (1), internal loop airlift contactor (2) and the external loop airlift contactor. The liquid is 82.0% glycerol solution. The data points with lines depict the riser holdup, while the downcomer holdup is shown by just the symbols.
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It is interesting to note that the downcomer holdup varies with the geometry of the airlift vessel and also with the type of liquid employed in the column. Table 5.1 shows the amount of gas recirculated with water in the external loop airlift vessel for each of the eleven superficial gas velocities employed. The amount of gas recirculated from the riser section into the downcomer section for all the 20 different liquids with the three airlift contactors is given in Table 5.2. For each experiment, an average of the gas recirculation rates obtained with the eleven different gas velocities is given. For example, in the case of water in the external loop airlift contactor, the average amount of gas recirculation was calculated as

<table>
<thead>
<tr>
<th>( U_g ) (mm/s)</th>
<th>32</th>
<th>42</th>
<th>63</th>
<th>85</th>
<th>106</th>
<th>127</th>
<th>148</th>
<th>169</th>
<th>190</th>
<th>212</th>
<th>233</th>
</tr>
</thead>
<tbody>
<tr>
<td>Amount of gas recirculated (%)</td>
<td>51.4</td>
<td>52.7</td>
<td>51.5</td>
<td>50.8</td>
<td>43.2</td>
<td>42.5</td>
<td>37.1</td>
<td>35.3</td>
<td>33.5</td>
<td>36.5</td>
<td>38.0</td>
</tr>
</tbody>
</table>

Table 5.1: Amount (%) of gas recirculation for a given gas velocity. \( Q_{\text{recirc}} = \frac{Q_{\text{downcomer}}}{Q_{\text{riser}}} \times 100 \)

Therefore the average amount of gas recirculated was 42.95%.

In general, for a given fluid, the internal loop airlift (2) configuration gave the lowest gas recirculation rate while the internal loop airlift (1) geometry produced the highest gas recirculation rate. Usually, the average amount of gas recirculated in the external loop airlift vessel was lower than in the internal loop airlift (1) contactor but higher than in the internal loop airlift (2) device depending on the type of liquid.

The most significant geometrical difference between the internal loop airlifts (1) and (2) was the downcomer to riser cross-sectional area ratio \( (A_d/A_r) \). The internal loop airlift (1) contactor had nearly twice the \( A_d/A_r \) ratio of that of the internal loop airlift (2) device. This meant that the riser and downcomer liquid velocities in the internal loop airlift (1) vessel was considerably higher than in the internal loop airlift (2) contactor (Section 5.1.2).

Generally, in an airlift reactor, the amount of gas recirculation basically determines the gas holdup in the downcomer. The gas recirculation rate is fundamentally influenced by the gas-liquid separator size and configuration (Siegel et al., 1986). Siegel et al. (1986) reported that gas recirculation rate was the ratio between the downcomer gas flow rate and the true riser gas flow rate. They also showed that the gas recirculation rate was dependent on the fluid residence time in the gas-liquid separator. They further reported that the liquid level in the separator influenced the gas recirculation rate. Higher liquid levels led to
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Figure 5.6: The effect of superficial gas velocity on the riser and downcomer gas holdups in the internal loop airlift contactor (1), internal loop airlift contactor (2) and the external loop airlift contactor. The liquid is 0.3% CMC solution. The data points with lines depict the riser holdup, while the downcomer holdup is shown by just the symbols.

longer fluid residence times and decreased gas recirculation. For this reason, in this study, a constant liquid level was maintained in all three airlift reactors. Since we were only interested in the effect of liquid viscosity on gas recirculation in the airlifts, as Siegel et al. (1988) reported that the downcomer gas holdup was also dependent on the liquid velocity in the column, bubble size distribution and on liquid properties (Section 2.3.1.1). Some investigators (Bello et al., 1985; Siegel et al., 1986) declared that usually in order for any gas to be recirculated the liquid velocity in the airlift must exceed about 0.2 to 0.3 m/s.

In this study, in the internal loop airlift (1) vessel with water and the salt solutions the riser superficial liquid velocity was in the range of 0.2 - 0.48 m/s; in the internal loop airlift (2) contactor the riser liquid velocity was in the range of 0.11 - 0.24 m/s, while in the external loop airlift device the riser liquid velocity’s range was 0.10 - 0.22 m/s. With more viscous liquids (glycerol and CMC solutions) the range of riser superficial liquid velocities was 0.04 - 0.33 m/s in the internal loop airlift (1) contactor. In the internal loop airlift (2)
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and the external loop airlift vessels, with viscous liquids, the ranges were 0.02 - 0.23 and 0.04 - 0.17 m/s, respectively. Hence it is apparent that even if the liquid velocity is below 0.2 m/s, riser gas is being recirculated.

As the riser liquid velocity in the internal loop airlift (1) contactor was higher than that in the internal loop airlift (2) vessel, this produced a higher gas recirculation. Since in the internal loop airlift (1) reactor due to the higher liquid velocity more bubbles would be recirculated into the downcomer instead of disengaging in the separator. Although usually the riser liquid velocity in the external loop airlift was lower than in the internal loop airlift (2) for a given gas flow rate, a higher proportion of the riser gas was recirculated in the external loop airlift vessel. This can be attributed to the different gas-liquid separator geometries. Siegel et al. (1986) showed that the geometry of the gas-liquid separator will affect the gas recirculation rate. The internal loop airlift (2) vessel had a larger top section than the external loop airlift device (Table 4.1). Hence, there was potential for more gas disengagement in the internal loop airlift (2) vessel than in the external loop airlift contactor at a given gas velocity.

In all three airlift contactors, with water and the three salt solutions as the liquid phase, there was not much difference in the amount of gas recirculated between the different liquids. However, the amount of gas recirculated into the downcomer increases with increasing Newtonian and non-Newtonian liquid viscosity (when the liquids were glycerol or CMC solutions). For example, in the internal loop airlift (2), the gas recirculation rate more than doubled (from 29.3% to 69.1%) when the liquid was changed from 48.5% glycerol solution (0.0052 Pa s) to 86.0% glycerol solution (0.0759 Pa s). There was almost a two fold increase in the amount of riser gas circulating into the downcomer when the liquid was changed from 0.05% CMC solution (n=0.90) to 0.6% CMC solution (n=0.57) in the external loop airlift contactor. With increasing oil-in-water concentration (increasing viscosity), the gas recirculation rate was also observed to increase. From 10% oil-in-water emulsion to pure oil there was almost a 25% increase in gas recirculation rate in the internal loop airlift (2) device.

As the liquid viscosity (Newtonian or non-Newtonian) increases the riser and downcomer superficial liquid velocities decrease (Section 5.3). From the above discussion it might be expected that with increasing liquid viscosity the gas recirculation rate might decrease due to the decreasing riser liquid velocity. But the effect of liquid viscosity on the riser gas recirculation rate is more complicated than this. With increasing liquid viscosity
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not only are the riser and downcomer liquid velocities decreasing but the riser and downcomer gas holdups are increasing (Section 5.3). The terminal rise velocity of a single bubble is also dropping with increasing liquid viscosity. There is also a change in the flow regime from bubbly/coalesced bubble flow to churn turbulent flow with increasing liquid viscosity. From visual observations, it was noticed that with viscous liquids the downcomer holdup was mainly due to small bubbles rather than large ones. This might be due to the fact that the fast rising large bubbles disengage at the top and most of the small bubbles recirculate into the downcomer.

The modelling of riser gas holdup as a function of superficial gas velocity is presented and discussed in Section 5.5.

\[
\begin{array}{|c|c|c|c|}
\hline
\text{Liquid} & \text{Internal loop (1)} & \text{Internal loop (2)} & \text{External loop} \\
\hline
\text{Water} & 47.7 \% & 21.1 \% & 43.0 \% \\
0.1M \text{ NaCl} & 51.1 \% & 18.4 \% & 45.4 \% \\
0.2M \text{ NaCl} & 50.8 \% & 16.4 \% & 41.3 \% \\
0.5M \text{ NaCl} & 48.6 \% & 20.7 \% & 43.9 \% \\
\text{Glycerol} & & & \\
48.5 \% & 54.8 \% & 29.3 \% & 44.6 \% \\
57.5 \% & 65.8 \% & 30.0 \% & 47.8 \% \\
71.5 \% & 74.6 \% & 42.2 \% & 58.9 \% \\
78.5 \% & 80.9 \% & 48.1 \% & 61.3 \% \\
82.0 \% & 85.1 \% & 56.8 \% & 67.8 \% \\
86.0 \% & 88.7 \% & 69.1 \% & 71.5 \% \\
\text{CMC} & & & \\
0.05 \% & 47.4 \% & 29.9 \% & 41.4 \% \\
0.2 \% & 57.2 \% & 33.6 \% & 46.5 \% \\
0.3 \% & 68.8 \% & 43.6 \% & 51.7 \% \\
0.4 \% & 79.4 \% & 59.4 \% & 62.2 \% \\
0.6 \% & 89.8 \% & 72.1 \% & 73.4 \% \\
\text{Oil-in-water} & & & \\
10 \% & 39.6 \% & 22.6 \% & 35.2 \% \\
20 \% & 41.7 \% & 30.1 \% & 38.5 \% \\
40 \% & 43.2 \% & 35.4 \% & 40.5 \% \\
60 \% & 47.6 \% & 42.2 \% & 41.3 \% \\
100 \% (no water) & 64.4 \% & 53.0 \% & 61.0 \% \\
\hline
\end{array}
\]

Table 5.2: The amount (\%) of riser gas recirculating into the downcomer.
5.1.2 The Effect of Superficial Gas Velocity on Liquid Velocity

The liquid circulation velocity in an airlift reactor can be represented by either the linear liquid velocity or by the superficial liquid velocity (Section 2.3.1.2). Like the gas holdup, the liquid circulation rate can also be separated into the overall, riser and downcomer liquid velocities. The average riser and downcomer superficial liquid velocities were measured in this work in the riser and downcomer sections of the airlift, respectively (Section 4.6). With all the liquids, the liquid velocity measurements were made only after a stable value of gas holdup was achieved.

Figure 5.7 shows the effect of superficial gas velocity on the riser and downcomer superficial liquid velocities in the internal loop airlift (1), in the internal loop airlift (2), and in the external loop airlift contactors with water. The given liquid velocity values are an average from three separate runs. Both the riser and downcomer liquid velocities, like the gas holdups, increased with increasing gas flow rate. It was also observed that there was an increase in the absolute difference between the riser and downcomer liquid velocities with increasing gas velocity in all three airlifts. This was similar to the increase in the difference between riser and downcomer gas holdups with increasing gas velocity (Figs 5.4 - 5.6). This phenomenon was also observed with the other 19 liquids in the three airlift devices and might be due to a change in the two phase flow regime.

The effect of superficial gas velocity on the riser and downcomer superficial liquid velocities with 82.0% glycerol solution and 0.3% CMC solution are presented in Figures 5.8 and 5.9, respectively. A reduction in the rate of increase of riser and downcomer liquid velocities with increasing gas velocities roughly above 12.8 mm/s was noticed. This is likely to be due to the change in the flow regime from coalesced bubble to turbulent flow with viscous liquids. Merchuk and Stein (1981) explained that in the turbulent flow regime the formation and disengagement of large bubbles lead to greater dissipation of energy, and this would therefore reduce the amount of energy available for liquid circulation. Pollard (1995) also made similar observations in a draft-tube internal loop airlift reactor with baker’s yeast broth (a low viscosity fermentation liquid). But, he observed a decrease in the rate of increase of liquid velocity with increasing gas velocities above about 0.054 m/s. That is a change in the flow regime from coalesced bubble to turbulent flow took place at a higher gas velocity than was in this study. This was due to the low viscosity of the fermentation medium (viscosity \( \approx 0.001 \text{ Pa s} \)).
In all three airlift devices, the downcomer liquid velocity was higher than the riser liquid velocity (Figs 5.7 - 5.9). This was expected since the downcomer to riser cross sectional area ($A_d/A_r$) ratios, for all three airlifts, were below unity. A smaller downcomer cross sectional area (compared to the riser area) gives a higher downcomer liquid velocity. In the external loop airlift device the downcomer liquid velocity was more than twice the riser liquid velocity, depending on the gas velocity. This was on account of the $A_d/A_r$ ratio being 0.40. Since, the riser cross sectional area was more than twice the downcomer cross sectional area. For a given flow rate, a smaller cross sectional area gives a higher velocity. While in the internal loop airlift (1) vessel, the downcomer liquid velocity was only slightly higher than the riser liquid velocity. This was again mainly due to the $A_d/A_r$ ratio being 0.92. Which meant that the riser cross sectional area was nearly equal to the downcomer cross sectional area. The downcomer superficial liquid velocity was roughly twice that of the riser superficial liquid velocity in the internal loop airlift (2) contactor for a given gas velocity. The $A_r/A_d$ ratio of the internal loop airlift (2) reactor was 2.04.

Figure 5.7: The variation in liquid circulation velocity in water with gas flow rate in the internal loop airlift (1), internal loop airlift (2) and the external loop airlift. The liquid is water. The data points with lines represent the riser liquid velocity, while the downcomer liquid velocity is given by the symbols.
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Figure 5.8: The variation in liquid circulation velocity in water with gas flow rate in the internal loop airlift (1), internal loop airlift (2) and the external loop airlift. The liquid is 82.0% glycerol solution. The data points with lines symbolise the riser liquid velocity, while the downcomer liquid velocity is given by the symbols.

Numerous investigators (Bello et al. 1984; Chisti, 1989; Choi and Lee, 1993; Siegel et al., 1988) have reported the strong influence of the $A_d/A_r$ ratio on the superficial liquid velocity in airlift reactors. Choi and Lee (1993) showed that the riser superficial liquid velocity increased nearly three fold when the $A_d/A_r$ ratio of their external loop airlift vessel was changed from 0.11 to 0.53. In this study, for a given gas flow rate, the riser superficial liquid velocity in the internal loop airlift (1) was the highest (Figs 5.7 - 5.9). Generally, for a given gas flow rate, the riser liquid velocity in the internal loop airlift (1) was more than double the riser liquid velocity in the internal loop airlift (2). This was mainly due to the larger riser cross sectional area in the internal loop airlift (2) compared to the internal loop airlift (1). The riser superficial liquid velocity in the internal loop airlift (2) was slightly higher than in the external loop airlift for a specified gas flow rate. This was anticipated since the internal loop airlift (2) vessel had a slightly higher $A_d/A_r$ ratio (0.49) to that of the external loop airlift contactor (0.40). It must be remembered that both the internal loop
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Airlift contactors had a bottom section which was precisely designed to keep the frictional losses (at the bottom) to a minimum (Section 4.1.1). Chisti et al. (1988) claimed that in internal loop airlift reactors the frictional losses due to a change in the flow direction at the top is negligible. Compared to the internal loop airlift contactors, in this work, the external loop airlift vessel had four 90° bends which would result in higher frictional losses than in the internal loop airlift devices.

The main driving force for the liquid circulation in airlift contactors was the density difference between the riser and downcomer sections. However, the magnitude of the superficial liquid velocities in the three airlifts is higher (Figs 5.7 - 5.9) than would be expected if it was just due to the density difference. Since in some instances there was only a small density difference between the riser and downcomer. Ayazi Shamlou et al. (1994) have explained that the liquid circulation velocity was also due to the entrainment and transport of liquid in the wake associated with the rising bubbles.

![Diagram](image)

Figure 5.9: The variation in liquid circulation velocity in water with gas flow rate in the internal loop airlift (1), internal loop airlift (2) and the external loop airlift. The liquid is 0.3% CMC solution. The data points with lines denote the riser liquid velocity, while the downcomer liquid velocity is given by just the symbols.
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It has been reported that the liquid velocity is effected by the height of the reactor, due to the dependence of the liquid velocity on the hydrostatic pressure difference between the riser and the downcomer (Section 2.3.1.2). Thus, Siegel et al. (1988) suggested that experimental data obtained using taller airlift devices is more reliable for scale up purposes. When extrapolating experimental data from short airlifts, care must be taken to ensure that the end effects do not predominate (Chisti, 1989; Siegel et al., 1988). Siegel et al. (1988) further stated that it was good practice to design laboratory reactors with unaerated liquid heights sufficiently tall to minimise the influence of end effects. Hence, in this work, the laboratory scale airlift vessels were designed adequately tall enough to minimise the end effects. The unaerated liquid height in the external loop airlift contactor was maintained at 2.45 m, while the height was held at 1.94 m in the two internal loop airlift devices.

The modelling of riser superficial liquid velocity for a given superficial gas velocity in airlift reactors is undertaken in Section 5.5.

5.1.3 The Effect of Superficial Gas Velocity on Liquid Mixing

The liquid mixing in an airlift contactor can be characterised in terms of a mixing time and a liquid phase axial dispersion coefficient (Section 2.3.1.3). Although the axial dispersion coefficient and the mixing time are related to each other, in some situations one might prove to be more important than the other. For example, Kawase and Moo-Young (1989) suggested that in biotechnology the mixing time might be more useful than the axial dispersion coefficient. Since the mixing time is a more direct index of homogeneity of concentrations of components in broth, such as microorganisms, dissolved oxygen and substrates compared with the axial dispersion coefficient.

The overall mixing and circulation times were measured in this study as described in Section 4.6. Furthermore, the overall liquid phase axial dispersion coefficient together with the overall Bodenstein number were also obtained (Sections 3.3.2 and 4.6). Most of the previous studies in the literature have viewed the airlift reactor as a single unit. Hence a single global Bodenstein number would represent the dispersion in the reactor as a whole (Merchuk and Yunger, 1990). However, some researchers (Fields and Slater, 1983; Hallaile and Merchuk, 1988; Verlaan et al., 1986a; Verlaan and Tramper, 1987) found that the different sections of the airlift reactor have different mixing characteristics. Therefore, to understand the liquid mixing in the riser and the downcomer sections of the airlift
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reactor, the riser and downcomer Bo numbers and liquid phase axial dispersion coefficients were also obtained in this study. As with liquid velocity studies, all the mixing measurements were performed only after a stable value of gas holdup was achieved.

With increasing gas velocity, the liquid circulation time decreased gradually in all three geometries of airlift contactors with water (Fig 5.10). Generally, the circulation time tended to level off at higher gas velocities in the three airlifts irrespective of the liquid (Figs 5.10 - 5.12). The rate of decrease reduced as the gas flow rate was increased. A change in the flow regime from bubbly flow to coalesced bubble flow probably caused this effect. Also, with 82.0% glycerol solution and 0.3% CMC solution the change in the flow regime from bubbly/coalesced bubble to turbulent flow reduced the rate of decrease of liquid circulation time with increasing gas velocity (Figs 5.11 and 5.12). This was in agreement with the observations in Section 5.1.2, where a reduction in the rate of increase of liquid velocity was noticed as the gas velocity was increased. The liquid circulation time was directly influenced by the liquid velocity in the vessel. So if at lower gas velocities there was a more pronounced effect on the liquid velocity than at higher gas velocities then this would also be the case with the liquid circulation time.

Generally, for a specified gas velocity, a lower circulation time was observed in the internal loop airlift (1) vessel than in the internal loop airlift (2) contactor. This was because of the higher liquid velocities occurring in the internal loop airlift (1) device due to the higher $A_d/A_r$ ratio of the vessel. A higher liquid circulation time was observed in the external loop airlift device than in both the internal loop airlift vessels. The former had a higher closed loop length than the latter.

The relationship between 90% mixing time and superficial gas velocity (Figs 5.10 - 5.12) was similar to that observed for the circulation time with a gradual decrease in mixing time with increasing gas velocity. The mixing time, like the circulation time, tended to level off at higher superficial gas velocities. There was a more noticeable effect on the mixing time at lower gas flow rates than at higher gas velocities (similar to the liquid circulation time). As stated above, this was due to a change in the flow regime with increasing gas flow rates. The similarity between the liquid mixing and liquid circulation times implied that a close correlation existed between the two parameters in the three airlift vessels.

For a given gas velocity, the values of mixing time in the internal loop airlift (1) vessel was lower than in the internal loop airlift (2) contactor due to the higher liquid
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velocities occurring in the former. Therefore, the internal loop airlift (1) contactor had a faster liquid circulation and a better mixing performance than the internal loop airlift (2) vessel at the same gas velocities. The mixing time, for a specified gas flow rate, in the external loop airlift was higher than the mixing times in both the internal loop airlifts due to the higher liquid circulation time in the external loop airlift contactor.

Figure 5.10: The effect of superficial gas velocity on the mixing and circulation times in the internal loop airlift (1), internal loop airlift (2) and the external loop airlift with water. The data points with lines represent the mixing time, while the circulation time is shown by the symbols.

A number of researchers have reported that the mixing time decreased with increasing gas velocity (Margaritis and Sheppard, 1981; Onken and Weiland, 1980; Pandit and Joshi, 1983; Rousseau and Bu’lock, 1980). Margaritis and Sheppard (1981) found the mixing time sharply decreased with increasing gas velocity up to about 0.04 m/s and then for further increases in gas velocity the mixing time decreased at a slower rate. They inferred that this was due to the flow regime in the vessel changing from bubbly to churn turbulent flow. Velan and Ramanujam (1995) studied the effect of the superficial gas velocity on mixing time in a draft tube internal loop airlift reactor with Newtonian (water)
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Figure 5.11: The effect of superficial gas velocity on the mixing and circulation times in the internal loop airlift (1), internal loop airlift (2) and the external loop airlift with 82.0% glycerol solution. The data points with lines depict the mixing time, while the circulation time is given by the symbols.

and non-Newtonian (0.1 - 0.75% CMC solutions) fluids. They found the mixing time rapidly decreased with increasing gas velocity up to about 0.012 m/s and then for further increases in gas velocity (up to 0.02 m/s) the mixing time tended to reach a constant. This is in agreement with the results of this study. Velan and Ramanujam (1995) suggested that this could be due to a change in the flow regime from homogeneous bubbly flow to slug flow. Furthermore, they stated that the decrease in mixing time with increasing gas velocity was due to the increase in turbulence causing flow because of the introduction of the gas phase, which increases the dispersion effects and reduces the mixing time. They also observed the influence of gas velocity to be more pronounced at lower linear liquid velocities. Petrovic et al. (1990) observed the mixing time in a draft tube internal loop airlift reactor to decrease with increasing gas velocity. They stated that the gas flow regime has a secondary effect on the degree of mixing. When the liquid velocity is high enough, the entrainment of bubbles into the annulus starts. This stationary cloud of bubbles
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in the annulus, of a two phase system, promotes mixing. Hsiun and Wu (1995) also declared that increasing the superficial air velocity decreased the mixing time in a draft tube internal loop airlift reactor. This was due to increased eddy mixing in the column with increasing air velocity. The mixing time in an external loop airlift bioreactor was studied by Kawase et al. (1994) with Newtonian (water and glycerine) and non-Newtonian fluids (CMC and xanthan gum solutions). They found the mixing time to decrease with increasing superficial gas velocity. At higher gas flow rates (above about 0.02 m/s) the mixing time tended to level off and approach a constant value. Kawase et al. (1994) observed the liquid circulation time to also decrease with increasing gas velocity at low gas velocity region (below about 0.02 m/s). At higher gas flow rates, the circulation time was nearly independent of the superficial gas velocity. Lin et al. (1976) showed in an external loop airlift fermentor that the mixing and circulation times decreased with increasing superficial gas velocity (up to 0.05 m/s). Popovic and Robinson (1993) studied the liquid mixing in an external loop airlift reactor with non-Newtonian systems. Like Rousseau and

![Graph](image1.png)

Figure 5.12: The effect of superficial gas velocity on the mixing and circulation times in the internal loop airlift (1), internal loop airlift (2) and the external loop airlift with 0.3% CMC solution. The data points with lines portray the mixing time, while the circulation time is given by the symbols.
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Bu’Lock (1980), see Section 2.3.1.3, they have also employed the concept of the so-called specific mixing time - to compare the mixing times obtained for different fermentor volumes. Popovic and Robinson (1993) reported that the specific mixing times fell with increasing liquid circulation velocities. Fields et al. (1984) investigated the mixing in a concentric tube airlift reactor containing xanthan gum. They noticed a reduction in the circulation time with increasing gas flow rate and attributed this to the increasing head of voidage in the riser section.

The ratio of liquid mixing time to liquid circulation time was calculated in order to further understand the mixing process. This ratio is plotted as a function of superficial gas velocity in Figure 5.13 for the three geometries of airlift contactors with water. The ratio of mixing to circulation time is a measure of the number of circulations (N) required to achieve 90% homogeneity of an injected thermal pulse. A lower mixing to circulation time ratio indicates better mixing in the vessel than a higher ratio.

![Figure 5.13: The effect of superficial gas velocity on the ratio of mixing time to liquid circulation time in the internal loop airlift (1), internal loop airlift (2) and the external loop airlift with water.](image-url)
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In the internal loop airlifts (1) and (2), with water, the ratio of mixing to circulation time was in the region of 1.8 for all gas velocities. The ratio was slightly lower (1.6), for all gas velocities, in the external loop airlift with water. Hence, in the internal loop airlift contactors, about 1.8 circulations of the vessel were needed to disperse a hot liquid pulse to 90% homogeneity. While in the external loop airlift vessel, the hot liquid pulse was scattered to 90% homogeneity after approximately 1.6 circulations.

Figures 5.14 and 5.15 show the ratio of mixing to circulation time against gas velocity for 82.0% glycerol solution and 0.3% CMC solution, respectively. On the whole, about 1.1 circulations of the internal loop airlifts (1) and (2) were needed to disperse a hot liquid pulse to 90% homogeneity with both glycerol (82.0%) and CMC (0.3%) solutions, irrespective of the superficial gas velocity in the columns. In the external loop airlift, for all gas velocities, approximately 1.4 circulations of the vessel were required with the glycerol (82.0%) solution and 1.1 circulations with the 0.3% CMC solution to scatter the hot pulse to 90% homogeneity. The low ratios of mixing to circulation time indicate that the mixing in the three airlift vessels is relatively good.

![Figure 5.14: The effect of superficial gas velocity on the ratio of mixing time to liquid circulation time in the internal loop airlift (1), internal loop airlift (2) and the external loop airlift with 82.0% glycerol solution.](image-url)
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From the above, it seems that the ratio of mixing to circulation time was independent of the superficial gas velocity in all three geometries of airlift contactors. This implied that the mixing performance in the airlift was mainly determined by the frequency at which the hot liquid pulse travelled through the end section of the airlift vessel. Several previous researchers (Pollard, 1995; Russell, 1989; Weiland, 1984) also came to the same conclusion. Russell (1989) studied the mixing in a pilot scale concentric tube airlift fermenter. He reported that the ratio of mixing to circulation time seems fairly constant with increasing gas velocity and only showed a slight downward trend (Figure 4.15 of his thesis). This slight decline in the ratio was attributed to increased turbulence in the vessel as the gas velocity increased. From Figs 5.13 - 5.15, it is apparent that the rate of mixing in an airlift vessel was primarily controlled by the bulk circulation of the liquid rather than by the axial dispersion due to the ascending bubbles.

Figure 5.15: The effect of superficial gas velocity on the ratio of mixing time to liquid circulation time in the internal loop airlift (1), internal loop airlift (2) and the external loop airlift with 0.3% CMC solution.
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The relationship between the superficial gas velocity and Bodenstein (Bo) numbers in the internal loop airlift (1), internal loop airlift (2) and external loop airlift contactors with water are shown in Figures 5.16, 5.17 and 5.18, respectively. The overall, riser and downcomer Bodenstein (Bo) numbers are given. Throughout this study, the overall Bo numbers were obtained using Equations 3.57 and 3.58 (Section 3.3.2). A comparison of the overall Bo numbers obtained by this method and by the curve fitting method (Eqn 3.51) is made later in this section.

In all three geometries of airlifts, for a given gas velocity, the riser Bo numbers were higher than the overall Bo numbers which in turn were higher than the downcomer Bo numbers. This observation was also made with the other 19 liquids in the three airlift contactors. It indicated that the mixing in the downcomer section was better than the mixing in the riser section for all three geometries of airlifts. The overall, riser and downcomer Bo numbers were observed to decrease with increasing gas velocity for water (Figs 5.16 - 5.18), which implied that the mixing in both the riser and downcomer sections and for the reactor as a whole was improving with increasing gas flow rate. Similarly, with the other 19 liquids, all three Bo numbers decreased with increasing gas velocity.

There are only a few previous studies of the local Bodenstein numbers in airlift reactors. Chisti (1989) stated that the Peclet (Pe) number in the downcomer section is usually higher than the riser Pe numbers for a given gas velocity. But, for a given gas flow rate, Verlaan et al. (1989) observed the overall Bo numbers in their reactor to be higher than the downcomer Bo numbers which were in turn higher than the riser Bo numbers. Lu et al. (1994) also observed the overall Bo numbers to be higher than their riser or downcomer Bo numbers for a given gas velocity. Their downcomer Bo number, for a specified gas flow rate, was higher than their riser section Bo number.

In this work, with the liquid water, the overall Bo numbers in the internal loop airlift (1) vessel were between 43 and 54 for the gas velocities used. The riser Bo number range was 52 to 75 while the range of the downcomer was lower (3 - 10). In the internal loop airlift (2) contactor, the overall Bo number range was 37 to 51 while the riser Bo range was higher (50 - 84). The range of the downcomer Bo number was between 2 and 10. The overall Bo numbers in the external loop airlift reactor were between 20 and 31. The downcomer Bo number range was from 2 to 14, while the riser Bo number range was higher (28 - 45).
Verlaan et al. (1989) reported that in an external loop airlift reactor with water the riser Bo numbers were in the range of 30 to 40 while the downcomer Bo numbers were in the 40 to 50 range. Their overall Bo numbers were between 40 to 60, while Fields and Slater (1983) stated their overall Bo numbers to be between 40 to 70 (in an internal loop airlift with water). However, Obradovic et al. (1994) reported their overall Bo numbers to be between 0.3 and 0.9 in an internal loop airlift device (with air, water and calcium alginate particles) although they obtained riser Bo numbers between 5 and 50 and downcomer Bo numbers from 20 to 60. They explained that their overall Bo numbers were much lower than the other researchers due to better liquid mixing in their reactor. But, if this was the case then Obradovic et al.'s (1994) riser and downcomer Bo numbers would have been also much lower than previous researchers' numbers. Lu et al. (1994; 1994a) also obtained Bo numbers in the individual sections of an internal loop airlift contactor by time domain analysis (similar to Verlaan et al.'s method). They quoted riser Bo numbers to be between 20 to 30 and the downcomer Bo numbers to be in the range of 35 to 65 for the two phase system (air/water). Their overall Bo numbers for a two phase system were in the range of 42 - 81. Li et al. (1993) found their riser Pe numbers to be between approximately 40 and 55 in an internal loop airlift reactor with water. Glennon et al. (1988) obtained the overall Bo numbers in an external loop airlift reactor with water and xanthan gum solutions. For water, they reported the overall Bo numbers to be between 85 and 105 for superficial gas velocities from about 0.005 to 0.04 m/s. Bello (1981) studied the overall Bo numbers in four different geometries of external loop airlift vessels and three different concentric tube internal loop airlift devices with water. He reported that the Bo values show very little variations with increasing gas velocity (0.0137-0.0860 m/s in the external loops), although a slight decrease occurred. Bello (1981) also stated that the Bo numbers increased with increasing $A_d/A_r$ ratio. He explained that this was due to increasing liquid velocity with increasing $A_d/A_r$ ratio and this increases the convective component of the mass transport in relation to the dispersive and making the reactor more plug flow. However, if this was the case then it would have been reasonable to notice an increase in the Bo numbers with increasing gas flow rate (since the liquid velocity in the column is increasing with gas flow rate and making the flow in the vessel more plug).

The overall, riser and downcomer liquid phase axial dispersion coefficients (D) for all three airlift vessels, with water, increase with increasing gas velocity (Figs 5.19 - 5.21). Similarly, with the other 19 liquids in the three airlift contactors all three dispersion
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coefficients were observed to increase. This was in agreement with the observed effect of gas velocity on the Bodenstein numbers. Since if the riser, downcomer and overall Bodenstein numbers decrease with increasing gas velocity then the riser, downcomer and overall axial dispersion coefficients would be expected to increase - to indicate that mixing was improving with increasing gas flow rate.

Li et al. (1993) reported that with water and CMC solutions, the Peclet numbers decrease with increasing gas velocity while the liquid phase axial dispersion coefficients increase. This is in agreement with the present findings. Fields and Slater (1983) found their Bo numbers to decrease with increasing gas velocity. Obradovic et al. (1994) also found their Bo numbers to decrease with increasing gas flow rate, while the corresponding liquid phase axial dispersion coefficient increased. Nevertheless, Verlaan et al. (1989) reported that in their reactor, the riser and downcomer Bo numbers increased with increasing gas velocity as well as the riser and downcomer dispersion coefficients. Verlaan et al.’s (1989) results seem to contradict each other. If the Bodenstein numbers increase with increasing gas velocity then this implies that the flow in the reactor is becoming more plug and therefore the dispersion coefficient would be expected to decrease with increasing gas flow rate. In their external loop airlift reactor, Glennon et al. (1988) also found the overall Bodenstein numbers to increase with increasing superficial gas velocity. With increasing gas velocity, Lu et al. (1994) found their Bo numbers to decrease while their dispersion coefficients increase. Frohlich et al. (1991) also observed a decrease in the Bo numbers with increasing gas velocity.

The downcomer dispersion coefficient (D), at a given gas velocity, was much higher than the riser or the overall D in all three geometries of airlift vessels with water (since the downcomer Bo numbers were low this corresponded with high downcomer dispersion coefficients). Similar observations were also made with the other 19 liquids in the three airlift vessels. This was due to the relatively high liquid velocity in the downcomer sections of the airlift vessels and hence a high turbulence intensity.

Some previous researchers (Li et al., 1993; Verlaan et al., 1989) have also found the liquid phase axial dispersion coefficients in the downcomer to be higher than the riser dispersion coefficients for a given gas flow rate. Verlaan et al. (1989) even found their downcomer dispersion coefficients to be higher than their gas-liquid separator dispersion coefficients. This implied that the liquid mixing in their downcomer was even higher than the mixing in their separator. They attributed the high downcomer axial dispersion
coefficients to the relatively high liquid velocity in the downcomer section. Pollard (1995) found the dissolved oxygen tension and the volumetric mass transfer coefficient were higher in the downcomer than in the riser section of a draft tube internal loop airlift reactor. He stated that this was due to long gas residence times, high gas content and smaller bubble diameter in the downcomer of the airlift compared to the riser section.

In this study, the riser dispersion coefficient was lower than the downcomer liquid phase axial dispersion coefficient, for all gas flow rates. This was mainly due to the riser liquid velocity being much lower than the downcomer liquid velocity in all three geometries of airlift reactors. It seems the presence of air bubbles has only a minor influence on the axial dispersion coefficient and that the dispersion is primarily affected by the liquid velocity in the airlift vessel. The riser D was also lower than the overall D, for a given gas flow rate. But, for a fixed gas velocity, the downcomer D was higher than the overall D. This indicated that the overall liquid mixing in the three airlift columns was somewhere in between the liquid mixing in the riser and downcomer sections.

In the internal loop airlift (1) contactor, for the liquid water, the overall D range was from 0.0539 to 0.1364 m²/s for the gas velocities employed. The riser D range was between 0.0232 to 0.1700 m²/s, while the range of the downcomer D was from 0.1582 to 0.5031 m²/s. The range of the overall D was from 0.0145 to 0.0395 m²/s in the internal loop airlift (2) vessel. The riser D range was between 0.0079 and 0.0293, while the downcomer D range was higher (0.1342 - 0.4459 m²/s). In the external loop airlift contactor, the overall D range was from 0.0090 to 0.0989 m²/s, while the range of the riser D was between 0.0039 and 0.0101 m²/s. But, the downcomer D range was higher (0.0671 - 0.5844 m²/s).

Verlaan et al. (1989) reported that in their external loop airlift reactor with water (0.165 m³ reactor volume; gas velocity from 0.015 to 0.140 m/s) the riser dispersion coefficient was approximately between 0.0270 to 0.0480 m²/s. Their downcomer dispersion coefficient was from 0.0620 to 0.1200 m²/s. Li et al. (1993) found their riser dispersion coefficients, in an internal loop airlift vessel with water, to be between 0.0100 and 0.0260 m²/s and the range of the downcomer D to be from 0.0120 to 0.0330 m²/s. Wu and Jong (1994) studied the overall liquid phase axial dispersion in a draft tube internal loop airlift reactor (0.015 m³ vessel volume; Superficial air velocity from 0.005 to 0.06 m/s) with different draft tubes (various diameters). They reported their overall dispersion coefficients to be from about 0.0020 to approximately 0.0750 m²/s.
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Figure 5.16: The effect of superficial gas velocity on the Bodenstein number in the internal loop airlift (1) contactor with water.

Figure 5.17: The effect of superficial gas velocity on the Bodenstein number in the internal loop airlift (2) contactor with water.
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Figure 5.18: The effect of superficial gas velocity on the Bodenstein number in the external loop airlift contactor with water.

Figure 5.19: The effect of superficial gas velocity on the liquid phase axial dispersion coefficient in the internal loop airlift (1) contactor with water.
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Figure 5.20: The effect of superficial gas velocity on the liquid phase axial dispersion coefficient in the internal loop airlift (2) contactor with water.

Figure 5.21: The effect of superficial gas velocity on the liquid phase axial dispersion coefficient in the external loop airlift contactor with water.
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In terms of overall, riser and downcomer liquid phase axial dispersion coefficients, the internal loop airlift (1) vessel was better than the internal loop airlift (2) device for nearly all gas velocities primarily due to higher liquid velocities in the former. This was in agreement with the mixing time results, which indicated that the internal loop airlift (1) gave better mixing than the internal loop airlift (2) for a specified gas velocity. The external loop airlift vessel, for a given gas flow rate, gave slightly higher overall, riser and downcomer dispersion coefficients than the internal loop airlift (1) contactor.

As stated earlier in this section, the overall Bodenstein numbers were obtained in this study with Eqns 3.57 and 3.58. Another way to procure the overall Bo number in an airlift reactor is by fitting the best theoretical line (generated by Eqn 3.51 for a given system) to the experimental data (Section 3.3.2). To investigate the agreement between the two methods, the overall Bo numbers were also achieved using Eqn 3.51 (curve fitting method was employed). Figure 5.22 shows the overall Bo numbers obtained in the external loop airlift contactor, for one liquid, by the two methods against gas velocity.

![Figure 5.22: The overall Bodenstein numbers in the external loop airlift vessel. Comparison between the overall Bodenstein numbers obtained by Equation 3.51 and that attained by Equation 3.58. The liquid is 48.5 % glycerol solution.](image)

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It is apparent that there was a reasonable agreement between the overall Bo number procured by Eqn 3.58 and that obtained by the curve fitting method (Eqn 3.51). This was also observed with the other 19 liquids in the external loop airlift contactor.

5.1.4 Summary

The superficial gas velocity, within the range used in this study, significantly influenced the hydrodynamics and the liquid mixing in the internal and external loop airlift reactors. After the gas flow was turned on, the gas holdup in viscous fluids was observed to increase with the time of aeration. Both the riser and downcomer gas holdups were observed to increase as the gas flow rate was increased. Similarly, the riser and downcomer superficial liquid velocities increase with increasing gas flow rate. Generally, the amount of riser gas recirculating into the downcomer increases with increasing liquid viscosity.

The mixing and circulation times were observed to decrease with increasing gas flow rate. However, the ratio of the mixing to circulation time seemed unaffected by the superficial gas velocity. As the gas velocity increased, the overall, riser and downcomer Bodenstein numbers decreased. Conversely, the overall, riser and downcomer liquid phase axial dispersion coefficients increased as the gas flow rate was increased. For a given gas velocity, the Bodenstein numbers in the airlift contactors were found to be in the order of Bo (riser) > Bo (overall) > Bo (downcomer). This implied that the liquid mixing in the downcomer was better than in the riser section and is confirmed by the dispersion coefficient values. The axial dispersion coefficients in the airlift devices were found to be in the order of D (downcomer) > D (overall) > D (riser), for a specified gas flow rate.

The internal loop airlift (1) geometry gave lower gas void fraction than the internal loop airlift (2) vessel for a given gas flow rate. But, for a specified gas velocity, higher liquid circulation velocities were observed in the internal loop airlift (1) device than in the internal loop airlift (2) contactor. Consequently, the internal loop airlift (1) vessel exhibited a better mixing performance than the internal loop airlift (2) contactor for a given gas flow rate.
5.2 The Effect of Electrolyte Concentration on Airlift Hydrodynamics and Mixing

In this section, the influence of salt concentration on airlift hydrodynamics and liquid mixing will be investigated. Several different concentrations of sodium chloride salt solutions were used (i.e. concentrations typical of fermentation media). Tables A.1 to A.3 give the physical properties of the electrolyte solutions employed in the three geometries of airlift contactors.

5.2.1 The Effect of Electrolyte Concentration on Gas Holdup

The mean riser and downcomer gas holdups were measured in the three geometries of airlift contactors. Figures 5.23 to 5.25 show the local gas holdups as a function of electrolyte concentration in the three airlifts for various superficial gas velocities. For all gas velocities, the riser gas holdup increased as the salt concentration was increased. Similarly, the downcomer gas holdup was also observed to increase with increasing electrolyte concentration. In the internal loop airlift (1) vessel, there was a 55% increase in riser gas holdup from water (without salt) to 0.5 M salt solution at a gas velocity of 0.0032 m/s. However, at a higher gas velocity (0.0162 m/s) there was only a 32% increase in riser holdup from water to 0.5 M salt solution. This phenomenon was also observed with the other two geometries of airlift reactors. For example, in the internal loop airlift (2) contactor from water to 0.5 M salt solution, the riser gas holdup increased by 30% and 14% at a gas velocity of 0.003 m/s and 0.015 m/s, respectively. In the external loop airlift device, from water to 0.5 M electrolyte solution, the riser gas holdup increased by almost 87% at a gas velocity of 0.0042 m/s while it increased by only 24% at a gas velocity of 0.0212 m/s. Likewise the downcomer gas holdup, from water to 0.5 M salt solution, was also observed to show a larger increase at a lower gas velocity than at a higher gas velocity. For example, in the internal loop airlift (1) the downcomer holdup increased by almost 56% at a gas velocity of 0.0032 m/s and by just 30% at a gas velocity of 0.0162 m/s. There was a more pronounced effect on the riser and downcomer gas holdups at a lower gas velocity than at a higher gas flow rate. One of the causes of this observed phenomenon might be the change in the flow regime from bubbly to coalesced bubble flow as the gas velocity was increased.

A number of previous studies (Jamialahmadi and Muller-Steinhagen, 1992; Onken and Weiland, 1980; Snape et al., 1992) have reported, similar to this work, that the gas holdup in a coalescent liquid (pure water) was lower than in a non-coalescent liquid (salt
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solution). This was mainly due to the large bubbles occurring in the former which because of their higher rise velocity and shorter residence times tended to reduce gas holdup. Onken and Weiland (1980) found electrolyte solutions to have only a small effect on gas holdup in an airlift reactor compared to that in a bubble column. However, it is apparent from this study, that gas holdup was significantly influenced by the concentration of electrolyte in the liquid phase of airlift reactors (especially up to 0.2 M). Kelkar et al. (1983) varied the NaCl concentrations from 0.01 M to 1.00 M in a bubble column. They noted that the gas holdup increased with an increase in the concentration, but at higher concentrations (from 0.2 M to 1.00 M) the effect became insignificant. Snape et al. (1992) reported that at low air flow rates an increase in the concentration of the electrolyte solution (from 0.01 to 0.2 M) had little effect on the gas holdup. However, at higher gas flow rates the gas holdup increases.

Figure 5.23: The effect of electrolyte concentration on the riser and downcomer gas holdups in the internal loop airlift (1) contactor for several superficial gas velocities. The data points with lines depict the riser holdup, while the data points without the lines give the downcomer holdup.
with increasing electrolyte concentration. The range of the superficial gas velocity in their external loop airlift reactor was from 0.025 to 0.2 m/s. They employed five different salt solutions and reported that, at a given gas velocity, sodium sulphate salt gave the highest gas holdup. This was followed by calcium chloride and magnesium sulphate salts. Sodium chloride and potassium chloride gave the lowest holdup values. They suggested that a reduction in the surface tension might be an important factor in the gas holdup being increased. Jamialahmadi and Muller-Steinhagen (1992) studied the effect of alcohols and organic acids on gas holdup in bubble columns. They report that in pure water the rate of bubble coalescence is high and consequently the gas holdup is low. In alcohols and organic acids, the rate of bubble coalescence is low and therefore the gas holdup is high.

Some researchers (Posarac and Tekic, 1987; Wachi et al., 1991) have employed alcohol solutions as the non-coalescing liquids. Wachi et al. (1991) found the addition of
ethanol increased the riser gas holdup. This is due to ethanol reducing the bubble size in the column and thereby inhibiting bubble coalescence.

The riser and downcomer gas holdups increase with increasing superficial gas velocities in all three airlift contactors (Figs 5.23 - 5.25). In the external loop airlift vessel, there was roughly a four fold increase in riser and downcomer gas holdups for a five fold increase in gas velocity. However, in the internal loop airlift (1) and (2) contactors, there were approximately only a three fold increase in riser and downcomer holdups for a five fold increase in gas velocity. As stated in Section 5.1, only two different flow regimes were observed with the salt solutions in the three airlifts for superficial gas velocities employed in this work.

Figure 5.25: The effect of electrolyte concentration on the riser and downcomer gas holdups in the external loop airlift contactor for several superficial gas velocities. The data points with lines represent the riser holdup, while the data points without the lines give the downcomer holdup.
5.2.2 The Effect of Electrolyte Concentration on Liquid Velocity

For all gas flow rates, the riser and downcomer superficial liquid velocities gradually decrease with increasing electrolyte concentration in the three airlift contactors (Figures 5.26 - 5.28). In the internal loop airlift (1) device, from pure water to 0.5 M salt solution, the riser superficial liquid velocity was observed to decrease by almost 17% for a gas velocity of 0.0032 m/s and by about 19% for a gas velocity of 0.016 m/s. The inhibition of coalescence, in salt solutions, results in small bubbles in the riser section. These bubbles recirculate into the downcomer thereby reducing the driving force for liquid circulation. This was in agreement with the findings in Section 5.2.1 where a higher gas holdup was observed with a salt solution than in pure water. Unlike with gas holdup, there was only a slight change in the percentage decrease of the liquid velocity with changes in gas velocity when the fluids were changed. This was also observed in the other two geometries of the airlift vessels.

Figure 5.26: The effect of salt concentration on the riser and downcomer superficial liquid velocities in the internal loop airlift (1) vessel for various gas velocities. The data points with lines denote the riser liquid velocity, while the downcomer liquid velocity is given by just the symbols.
There are very few studies in the open literature on the effect of electrolyte concentration on liquid velocity in airlift reactors. Wachi *et al.* (1991) suggested that a non-coalescent liquid (ethanol in their case) tended to reduce the liquid circulation rate compared to water. Weiland (1984) also commented that a non-coalescent liquid will give a lower liquid velocity. This was thought to be due to the inhibition of coalescence in the column resulting in small bubbles in the riser section. These small bubbles recirculate into the downcomer section giving a higher holdup in the downcomer and this reduces the driving force for liquid circulation in the vessel. However, from Table 5.2 it is obvious that this is not the case. Table 5.2 indicates that the riser gas recirculation into the downcomer section did not change significantly with increasing electrolyte concentration. Another possible explanation might be that the liquid-wake behind the rising bubbles (in the riser section) is smaller in a non-coalescent liquid, since only small bubbles are present. This meant that the amount of liquid carried by the liquid-wake is reduced and therefore could result in a lower liquid velocity.

Figure 5.27: The effect of salt concentration on the riser and downcomer superficial liquid velocities in the internal loop airlift (2) vessel for various gas velocities. The data points with lines represent the riser liquid velocity, while the symbols without the lines depict the downcomer liquid velocity.
The riser and downcomer superficial liquid velocities increase with increasing superficial gas velocities. Generally, the downcomer superficial liquid velocity was higher than the riser liquid velocity for a given gas flow rate in all the airlift contactors. In all three airlift vessel geometries, the riser and downcomer superficial liquid velocities roughly doubled for a five fold increase in the superficial gas velocity.

![Graph showing the relationship between salt concentration and superficial liquid velocity](image)

**Figure 5.28**: The effect of salt concentration on the riser and downcomer superficial liquid velocities in the external loop airlift device for various gas velocities. The data points with lines represent the riser liquid velocity, while the symbols without the lines show the downcomer liquid velocity.

### 5.2.3 The Effect of Electrolyte Concentration on Liquid Mixing

The 90% mixing time increases with increasing electrolyte concentration in the three airlift contactors (Figs 5.29 - 5.31). The liquid circulation time was also observed to slightly increase with increasing salt concentration. On average from water to 0.5 M salt solution, in the internal loop airlift (1) vessel, the mixing time increased by almost 60% while the liquid circulation time increased by just 22%. In other words, there was a bigger increase in mixing time than in circulation time. This was also observed in the other two geometries of airlift contactors. For example, in the external loop airlift device from water...
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to 0.5 M salt solution there was an average increase of almost 38% in mixing time and only about 13% in liquid circulation time. It seems that changing the liquid from pure water to a salt solution, produces a larger impact on the mixing time than on the circulation time.

![Figure 5.29](image)

Figure 5.29: The effect of electrolyte concentration on the mixing and circulation times in the internal loop airlift (1) for various superficial gas velocities. The data points with lines represent the mixing time, while the symbols without the lines denote the circulation time.

It has already been shown that the presence of electrolytes in the liquid medium has an effect on the gas holdup and liquid velocity in the airlift column (Sections 5.2.1 and 5.2.2). Since the gas holdup and liquid velocity affect the mixing in the reactor, the presence of salts in the liquid medium can also be expected to influence the liquid mixing. As the salt concentration was increased, the gas holdup was observed to increase and the liquid velocity decreased (Figures 5.23 - 5.28). Consequently, a lower superficial liquid velocity leads to an increase in the liquid mixing and circulation times. There are very few previous studies of liquid mixing with electrolytes in an airlift reactor compared to bubble columns. Pandit and Joshi (1983) studied the liquid mixing in a bubble column and found that the mixing time increased in the presence of electrolytes (sodium sulphate and sodium chloride). But, Onken and Weiland (1980) studied the mixing in an external loop airlift
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Figure 5.30: The effect of electrolyte concentration on the mixing and circulation times in the internal loop airlift (2) for various superficial gas velocities. The data points with lines represent the mixing time, while the symbols without the lines give the circulation time.

Figure 5.31: The effect of electrolyte concentration on the mixing and circulation times in the external loop airlift for various superficial gas velocities. The data points with lines give the mixing time, while the symbols without the lines denote the circulation time.
fermentor. They reported that there was no clear difference in mixing time between water and salt (NaCl) solutions (0.1 - 1.0 M). In their work, the gas holdup and superficial liquid velocity was not influenced by the electrolyte concentration. This could be a reason why the mixing time was also unaffected by the concentration of the salt solution.

The electrolyte concentration had no noticeable effect on the ratio of mixing to liquid circulation time in the three airlift contactors (Figs 5.32 - 5.34). This suggested that the rate of mixing in the airlift vessel was not affected by the concentration of the salt solution in the liquid phase. On average, approximately 1.6 circulations of the internal loop airlift (1) and (2) vessels were required to disperse a hot liquid pulse to 90% homogeneity with all the salt solutions. In the external loop airlift contactor roughly 1.7 circulations were needed, irrespective of the concentration of the salt solution. The low ratio of mixing to circulation time indicated that mixing performance in the internal and the external loop airlift contactors, with water and electrolyte solutions, was relatively good.

![Figure 5.32: The effect of electrolyte concentration on the ratio of mixing to liquid circulation time in the internal loop airlift (1) for several superficial gas velocities.](image)

Figure 5.32: The effect of electrolyte concentration on the ratio of mixing to liquid circulation time in the internal loop airlift (1) for several superficial gas velocities.
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Figure 5.33: The effect of electrolyte concentration on the ratio of mixing to liquid circulation time in the internal loop airlift (2) for several superficial gas velocities.

Figure 5.34: The effect of electrolyte concentration on the ratio of mixing to liquid circulation time in the external loop airlift for several superficial gas velocities.
The effect of salt concentration on the overall, riser and downcomer Bodenstein numbers in the three geometries of airlift contactors are shown in Figures 5.35 to 5.37. Generally, all three Bodenstein (Bo) numbers increase with increasing electrolyte concentration indicating that the liquid mixing in the airlift vessel was deteriorating. This is confirmed by Figures 5.38 to 5.40, which showed the overall, riser and downcomer liquid phase axial dispersion coefficients decrease with increasing salt concentration in all three airlift vessels. The dispersion coefficients would be expected to decrease with increasing electrolyte concentration, since the superficial liquid velocities in the airlift columns were decreasing with increasing salt concentration. This finding was in agreement with the mixing time results which also indicate that the mixing was diminishing with increasing salt concentration.

A higher superficial gas velocity gave a lower Bo number than a lower gas flow rate (Figs 5.35 - 5.37). As discussed in Section 5.1.3, mixing improves with increasing gas flow rate. This was confirmed by the fact that the dispersion coefficient increased as the gas flow rate was increased for a given liquid (Figs 5.38 - 5.40).

For a given salt concentration, the Bo number in the riser section was higher than the overall and downcomer Bo numbers. The downcomer Bo number was the lowest for a given electrolyte concentration. This indicated that the mixing in the downcomer was higher than in the riser section. The mixing in the reactor as a whole (overall) was in-between the mixing in the riser and the downcomer sections. This was corroborated by the axial dispersion studies (Figs 5.38 - 5.40) which indicated that the dispersion in the downcomer was higher than that in the riser section for a given salt concentration. This was mainly due to the higher liquid velocity occurring in the downcomer compared to the riser section, which gave rise to a higher turbulence intensity.

Generally, in the internal loop airlift (1) vessel, the overall Bo numbers were in the range of 32 to 54 for all the gas velocities with water and salt solutions. The riser Bo numbers were between 52 and 86 while the downcomer Bo numbers were between 3 to 10, for water and salt solutions. For the same fluids, in the internal loop airlift (2) reactor, the overall Bo numbers were between 35 and 51 while the riser and downcomer Bo numbers were in the range of 44 to 84 and 2 to 8, respectively. In the external loop airlift device, for water and salt solutions, a range of 20 to 70 was observed for the overall Bo numbers. The riser Bo number range was from 28 to 95, while the downcomer Bo number range was lower (2 - 14).
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With water and salt solutions, for all gas velocities, the range of the overall dispersion coefficient \( D \) was from 0.0060 to 0.1364 m\(^2\)/s in the internal loop airlift (1) vessel. For the same liquids, the riser \( D \) was between 0.0083 and 0.1700 m\(^2\)/s while the downcomer \( D \) range was from 0.0158 to 0.6023 m\(^2\)/s. In the internal loop airlift (2) reactor, for the same liquids, the overall \( D \) was from 0.0145 to 0.0727 m\(^2\)/s and the riser \( D \) range was between 0.0079 and 0.0918 m\(^2\)/s. The range of the downcomer \( D \) was from 0.0635 to 0.6623 m\(^2\)/s. In the external airlift loop contactor the overall \( D \) range was from 0.0090 to 0.0989 m\(^2\)/s. The riser \( D \) range was between 0.0010 and 0.0101 m\(^2\)/s, while the downcomer \( D \) range was higher (0.0176 - 0.5844 m\(^2\)/s).

For a given salt concentration and gas flow rate, higher dispersion coefficients were obtained in the internal loop airlift (1) contactor compared to the internal loop airlift (2) vessel. This was primarily due to the higher liquid velocities occurring in the internal loop airlift (1) contactor, which gives rise to better liquid phase mixing.

Surprisingly, there seems to be no previous studies of Bodenstein numbers and axial dispersion coefficients in airlift reactors with electrolyte concentrations in the published literature.

5.2.4 Summary

The hydrodynamics and liquid mixing in airlift vessels were influenced by the salt concentration in the liquid phase. The local gas holdups were observed to increase with increasing electrolyte concentration. On the other hand, the local liquid velocities decreased with an increase in the salt concentration. The mixing deteriorated with increasing salt concentration due to the decreasing liquid velocity in the airlift column. The 90% mixing time and the liquid circulation time decrease with increasing concentration of the salt in the liquid phase. The ratio of mixing to circulation time was unaffected by the salt concentration. The overall, riser and downcomer Bodenstein numbers increase with increasing electrolyte concentration while the overall, riser and downcomer dispersion coefficients decrease.
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Figure 5.35: The effect of electrolyte concentration on the overall, riser and downcomer Bodenstein numbers in the internal loop airlift (1). The data points with lines represent a superficial gas velocity of 0.0160 m/s, while the symbols without the lines depict a gas velocity of 0.0080 m/s.

Figure 5.36: The effect of electrolyte concentration on the overall, riser and downcomer Bodenstein numbers in the internal loop airlift (2). The data points with lines represent a superficial gas velocity of 0.0150 m/s, while the symbols without the lines depict a gas velocity of 0.0075 m/s.
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Figure 5.37: The effect of electrolyte concentration on the overall, riser and downcomer Bodenstein numbers in the external loop airlift. The data points with lines represent a superficial gas velocity of 0.0212 m/s, while the symbols without the lines depict a gas velocity of 0.0106 m/s.

Figure 5.38: The effect of electrolyte concentration on the overall, riser and downcomer liquid phase axial dispersion coefficients in the internal loop airlift (1). The data points with lines represent a superficial gas velocity of 0.0160 m/s, while the symbols without the lines depict a gas velocity of 0.0080 m/s.
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Figure 5.39: The effect of electrolyte concentration on the overall, riser and downcomer liquid phase axial dispersion coefficients in the internal loop airlift (2). The data points with lines represent a superficial gas velocity of 0.0150 m/s, while the symbols without the lines depict a gas velocity of 0.0075 m/s.

Figure 5.40: The effect of electrolyte concentration on the overall, riser and downcomer liquid phase axial dispersion coefficients in the external loop airlift. The data points with lines represent a superficial gas velocity of 0.0212 m/s, while the symbols without the lines depict a gas velocity of 0.0106 m/s.
5.3  The Effect of Liquid Viscosity on Airlift Hydrodynamics and Mixing

In this section, the effect of both Newtonian and non-Newtonian liquid viscosity on the airlift hydrodynamics and liquid mixing will be examined. Various concentrations of glycerol solutions were used as the viscous Newtonian fluids. Solutions of CMC were employed as the non-Newtonian fluids. Tables A.1 to A.3 give the physical properties of the glycerol and CMC solutions utilised in the three geometries of airlift contactors.

5.3.1 Newtonian Fluids

5.3.1.1 The Effect of Liquid Viscosity on Gas Holdup

The riser and downcomer gas holdups exhibit a gradual increase with increasing liquid viscosity in the three geometries of airlift vessels for various superficial gas velocities (Figures 5.41 - 5.43). A higher superficial gas velocity gave a higher riser or downcomer gas holdup than a lower superficial gas velocity, irrespective of the viscosity of the liquid (Section 5.1.1). In the internal loop airlift (1) contactor, for a 25 fold increase in the liquid viscosity the riser gas holdup increased by about 200% at a superficial gas velocity of 0.0032 m/s. However, there was only a 29% increase in the riser gas holdup at a superficial gas velocity of 0.0162 m/s for the same amount of increase in liquid viscosity. Generally, for increasing liquid viscosity, a more marked effect on riser and downcomer gas holdups was noticed at lower gas velocities than at higher gas flow rates. For example, in the external loop airlift contactor for a 33 fold increase in liquid viscosity the downcomer gas holdup increased by about 800% at a superficial gas velocity of 0.0042 m/s while it increased by just 50% at a superficial gas velocity of 0.0212 m/s. This might be due to a number of reasons. It could be due to the changes in the flow regime occurring in the airlift column with increasing gas velocity. Both homogeneous and heterogeneous flow regimes were observed in the airlift devices with viscous glycerol solutions (Section 5.1). Also, at higher liquid viscosities a larger amount of gas recirculated into the downcomer than at lower liquid viscosities (Table 5.2).

From the above, it is clear that both the riser and downcomer gas holdups increase with increasing Newtonian liquid viscosity. This was primarily due to the fact that increasing the liquid viscosity leads to the formation of large and small bubbles. The large bubbles have a higher rise velocity than smaller ones (in the present study this effect was observed). Hence, large bubbles tend to reduce the gas holdup. However, the small bubbles enhance the gas holdup due to their low rise velocity and this effect seems to predominate.
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over the former. It was noticed that high viscosity glycerol solutions contained more small bubbles in a specified volume of liquid, at a given gas flow rate and time of aeration, than solutions with a lower viscosity. This could also contribute to the effect of increasing gas holdup with liquid viscosity.

![Graph showing the relationship between superficial gas velocity, liquid viscosity, and riser and downcomer gas holdups.](image)

**Figure 5.41:** Effect of Newtonian liquid viscosity on riser and downcomer gas holdups in the internal loop airlift (1) for several superficial gas velocities. The data points with lines represent the riser holdup, while the downcomer holdup is given by the symbols.

Philip et al. (1990) reported the overall gas holdup in an internal loop airlift reactor, for a given superficial gas velocity, increased as the Newtonian liquid viscosity was increased (from 0.115 Pa s to 2.85 Pa s; they used four different liquids: olive oil, SAE, castor oil and sugar syrup). They suggested the reason for this might be that, at a given gas flow rate, the velocity of each large bubble is higher in the lower viscosity liquid due to higher liquid circulation velocities. This might be true in the riser, but cannot be valid in the downcomer section. Since in this work, both the riser and downcomer holdups were observed to increase as the liquid viscosity was increased for a specified gas velocity. The reason for this might be the increase in the concentration of small bubbles with increasing liquid viscosity. Philip et al. (1990) also observed this effect. They divided the gas holdup
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into two parts: gas holdup due to large bubbles and holdup due to small bubbles. Both the gas holdups due to large bubbles and small bubbles increase with liquid viscosity. However, there was a more pronounced effect, with increasing liquid viscosity, on the gas holdup due to the small bubbles than on the other at a fixed gas velocity. They suggested that the overall gas holdup also exhibited the same trend as the gas holdup due to the small bubbles.

Some studies (Deckwer, 1992; Heijnen and Van’t Riet, 1984; Shah et al., 1982) reported that the gas holdup decreased as the viscosity of the liquid phase was increased. They suggested that an increase in the liquid viscosity can lead to a reduction in the bubble rise velocity and this would increase gas holdup. But, the formation of large bubbles with increasing liquid viscosity has an opposite effect. They found the latter to be predominant over the former effect. However, it is unclear as to whether they measured the stable or the unstable value of gas holdup.

![Graph](image.png)

Figure 5.42: Effect of Newtonian liquid viscosity on riser and downcomer gas holdups in the internal loop airlift (2) for several superficial gas velocities. The data points with lines represent the riser holdup, while the downcomer holdup is given by the symbols.
In the current work, the rise velocity of an isolated bubble was measured in different fluids (this is presented and discussed in Section 5.5; see also Appendix 3). With increasing liquid viscosity, the terminal bubble rise velocity was observed to decrease slightly (Table 5.6). This might also have contributed to the increased riser gas holdup as the liquid viscosity was increased. Onken and Weiland (1980) noted the liquid viscosity to have little effect on the gas holdup in airlift contactors. However, Onken and Weiland have not stated whether they measured the stable or the unstable value of gas holdup in their airlift vessel.

Figure 5.43: Effect of Newtonian liquid viscosity on riser and downcomer gas holdups in the external loop airlift for several superficial gas velocities. The data points with lines symbolise the riser holdup, while the downcomer holdup is given by the symbols.

Snape et al. (1992) studied the gas holdup with aqueous sugar solutions (0.5 to 8% w/v) in an external loop airlift contactor. They reported that there was little effect of sugar concentration on gas holdup at the lowest gas flow rate, but an increasing effect at higher air flow rates. This contradicts the findings of this study. Low concentrations of sugar solutions gave an increase in gas holdup, yet higher concentrations led to a lower value. They further reported that at the highest sugar concentration (8%) the gas holdup was less
than that obtained with tap water. However, they did not state whether they measured the stable value of holdup.

5.3.1.2 The Effect of Liquid Viscosity on Liquid Velocity

The effect of Newtonian liquid viscosity on the riser and downcomer superficial liquid velocities for different superficial gas velocities are depicted in Figures 5.44 to 5.46. In all three airlift reactors, higher gas velocities gave higher superficial riser and downcomer liquid velocities than lower gas velocities irrespective of the liquid viscosity. For any given superficial gas velocity, the downcomer superficial liquid velocity was higher than the riser superficial liquid velocity as stated in Section 5.1.2.

![Graph showing the effect of liquid viscosity on superficial liquid velocities](image)

Figure 5.44: Effect of Newtonian liquid viscosity on riser and downcomer superficial liquid velocities in the internal loop airlift (1) for various superficial gas velocities. The data points with lines depict the riser liquid velocity, while the symbols without the lines denote the downcomer liquid velocity.

It is apparent that there is a decrease in both the riser and downcomer liquid velocities with increasing liquid viscosity in all three geometries of airlift contactors. In the internal loop airlift (1) contactor, for a 25 fold increase in liquid viscosity there was a 71%
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and a 60% decrease in the riser superficial liquid velocity at superficial gas velocities of 0.0032 m/s and 0.0162 m/s, respectively (Fig. 5.44). There was a 87% decrease in the downcomer superficial liquid velocity in the internal loop airlift (2) reactor at a superficial gas velocity of 0.0030 m/s while there was a 76% decrease in downcomer superficial liquid velocity at a gas velocity of 0.0150 m/s (Fig. 5.45). There is a slightly more prominent effect on the liquid velocity at a lower gas flow rate than at a higher gas velocity (similar to the effect on gas holdup).

It is clear from the above that Newtonian liquid viscosity influences the riser and downcomer liquid velocities. This was primarily due to the fact that with increasing viscosity of the liquid phase, there was an increase in the frictional resistance to flow around the airlift loop. Therefore, with increasing liquid viscosity, the energy losses occurring in the airlift column were also increasing and this directly impedes the liquid velocity in the vessel. Also, the amount of gas recirculating into the downcomer section is

![Figure 5.45: Effect of Newtonian liquid viscosity on riser and downcomer superficial liquid velocities in the internal loop airlift (2) for various superficial gas velocities. The data points with lines depict the riser liquid velocity, while the symbols without the lines denote the downcomer liquid velocity.](image)
Increasing with an increase in the liquid viscosity. This results in a higher downcomer gas holdup and consequently a lower driving force for liquid circulation.

Onken and Weiland (1980) reported that increasing the liquid viscosity by a factor of 10 (saccharose solution) decreased the liquid velocity by 10%. Several other studies (Onken and Trager, 1990; Onken and Weiland, 1983; Philip et al., 1990) also reported the liquid circulation velocity decreased as the liquid viscosity was increased. Onken and Trager (1990) stated that, with glycerol solutions, higher liquid viscosity resulted in lower liquid velocity. However, they commented that liquid viscosity did not seem to have a major effect on the liquid velocity. This might be because they employed only a narrow range of liquid viscosities (from 4.87 to 12.00 mPa s). Philip et al. (1990) reported the liquid circulation velocity to decrease with an increase in the liquid viscosity (viscosity range: from 0.115 to 2.850 Pa s). They suggested that this was due to higher viscous drags at the walls of the riser and downcomer at a higher liquid viscosity.

![Figure 5.46: Effect of Newtonian liquid viscosity on riser and downcomer superficial liquid velocities in the external loop airlift for various superficial gas velocities. The data points with lines depict the riser liquid velocity, while the symbols without the lines denote the downcomer liquid velocity.](image)

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5.3.1.3 The Effect of Liquid Viscosity on Liquid Mixing

The mixing and circulation times increase with increasing liquid viscosity for all gas flow rates (Figs 5.47 -5.49). There was a 356% increase in the mixing time for a 25 fold increase in the liquid viscosity in the internal loop airlift (1) contactor for a gas velocity of 0.0032 m/s. But, there was only a 121% increase in the mixing time at a gas velocity of 0.0162 m/s for the same increase in liquid viscosity. A more pronounced effect on mixing time, as the liquid viscosity was increased, occurred at lower gas flow rates than at higher gas velocities. This phenomenon was also observed in the other two airlift vessels. Similarly, in the internal loop airlift (2) reactor, the liquid circulation time showed a 401% increase for a 15 fold increase in liquid viscosity at a gas velocity of 0.0030 m/s. However, at a higher gas velocity (0.0150 m/s) there was only a 229% increase in circulation time for the same increase in liquid viscosity. At a lower gas velocity, there was a more noticeable effect on the mixing and circulation times than at a higher gas flow rate. This was mainly due to the fact that at a lower gas velocity there was a larger decrease in the liquid velocity than at a higher gas flow rate. Since the mixing and circulation times are

Figure 5.47: Effect of Newtonian liquid viscosity on the mixing and circulation times in the internal loop airlift (1) for various superficial gas velocities. The data points with lines symbolise the mixing time, while the symbols without the lines signify the circulation time.
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primarily affected by the liquid velocity in the vessel, a more pronounced effect on the liquid velocity at a lower gas flow rate will consequently result in a more noticeable effect on the mixing and circulation times at a lower gas velocity.

As already stated, the mixing and circulation times in an airlift reactor are primarily dependent on the liquid velocity in the column. A decrease in the liquid velocity (Section 5.3.1.2), with increasing liquid viscosity, will bring about a corresponding increase in the mixing and circulation times (Figs 5.47 - 5.49). Guy et al. (1986) studied the liquid mixing in a bubble column with viscous liquids. They found the mixing efficiency decreased as the liquid viscosity was increased. This is in agreement with the present findings. Pandit and Joshi (1983) observed, in bubble columns, the mixing time to increase with an increase in the liquid viscosity. They also noticed, similar to the present results, that at lower gas superficial gas velocities the effect of viscosity was much more pronounced than at higher gas flow rates.

![Figure 5.48: Effect of Newtonian liquid viscosity on the mixing and circulation times in the internal loop airlift (2) for various superficial gas velocities. The data points with lines symbolise the mixing time, while the symbols without the lines signify the circulation time.](image)

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Figure 5.49: Effect of Newtonian liquid viscosity on the mixing and circulation times in the external loop airlift for various superficial gas velocities. The data points with lines symbolise the mixing time, while the symbols without the lines signify the circulation time.

In Figures 5.50 to 5.52, the ratio of the mixing to circulation time at a number of air flow rates is plotted against the liquid viscosity for the three airlift contactors. It is apparent that the ratio of the mixing to circulation time is not influenced by the viscosity of the liquid. This implied that the rate of mixing in the airlift reactor was independent of the liquid phase viscosity.

In the internal loop airlift (1) and (2) contactors, on average, roughly 1.3 and 1.1 circulations were required to disperse a hot liquid pulse to 90% homogeneity, respectively. The number of circulations required were slightly higher (1.5) in the external loop airlift vessel. The low ratio of mixing to circulation time indicated relatively good mixing in the internal and the external loop airlift reactors, with glycerol solutions.
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Figure 5.50: Effect of Newtonian liquid viscosity on the ratio of liquid mixing to liquid circulation time in the internal loop airlift (1) for different superficial gas velocities.

Figure 5.51: Effect of Newtonian liquid viscosity on the ratio of liquid mixing to liquid circulation time in the internal loop airlift (2) for different superficial gas velocities.
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Figure 5.52: Effect of Newtonian liquid viscosity on the ratio of liquid mixing to liquid circulation time in the external loop airlift for different superficial gas velocities.

The overall, riser and downcomer Bodenstein (Bo) numbers were observed to increase with an increase in the liquid viscosity for a fixed gas flow rate (Figs 5.53 - 5.55). This indicated that the mixing efficiency was decreasing with an increase in liquid viscosity. At a given liquid viscosity, usually the riser Bodenstein number was higher than the overall Bodenstein number which in turn was higher than the downcomer Bodenstein number, irrespective of the gas velocity. The mixing in the downcomer seems to be better than in the riser section, for a given liquid viscosity. This is due to the high liquid velocities occurring in the downcomer section, which increases the turbulence in the downcomer.

Figures 5.56 - 5.58 show the overall, riser and downcomer liquid phase axial dispersion coefficients (D) decrease as the liquid viscosity was increased. Especially, the downcomer D shows a sharp decrease as the liquid viscosity was increased. This confirmed the findings of the Bo number studies. Since, if the Bo numbers increased with an increase in liquid viscosity then the dispersion coefficients would be expected to decrease. The mixing in riser and downcomer sections of the airlift vessel were deteriorating with an increase in liquid viscosity. The overall mixing in the airlift was also declining with
increasing liquid viscosity. This was anticipated since the mixing time, for the reactor as a whole, increased with an increase in liquid viscosity. The liquid velocities in the airlift contactors were decreasing with increasing viscosity of the liquid phase (Section 5.3.1.2), which contributed to the lower dispersion coefficients at higher liquid viscosity.

Mixing efficiency improves with an increase in gas velocity irrespective of liquid viscosity. A higher gas flow rate gave a lower Bo number than a lower gas velocity at a fixed liquid viscosity (Figs 5.53 - 5.55). Conversely, a higher dispersion coefficient was obtained with a higher gas velocity than at a lower gas flow rate at a given liquid viscosity (Figs 5.56 - 5.58).

With glycerol solutions, in the internal loop airlift (1) vessel, the overall Bo numbers were in the range of 24 to 45. The riser Bo numbers were between 36 and 87 and 5 to 14 was the range for the downcomer Bo numbers. In the internal loop airlift (2) device with glycerol solutions, the overall Bo numbers were in the range of 24 to 53. While the riser Bo numbers were between 32 to 96 and the downcomer Bo numbers were from 3 to 13. For the same liquids, in the external loop airlift vessel, the overall Bo number range was from 20 to 43. The riser Bo number range was between 26 and 106 and the range of the downcomer Bo number was from 2 to 9.

The values of the overall dispersion coefficients with glycerol solutions in the internal loop airlift (1) contactor were between 0.0064 and 0.2693 m²/s. The range of the riser D was from 0.0021 to 0.0204 m²/s, while 0.0115 to 0.6350 m²/s was the range of the downcomer D. In the internal loop airlift (2) contactor, with the same liquids, the overall D range was between 0.0031 and 0.0592 m²/s. The riser D range was from 0.0017 to 0.0481 m²/s, while the downcomer D range was higher (0.0277 - 0.5778 m²/s). In the external loop airlift reactor, with glycerol solutions, the overall D was between 0.0102 and 0.0675 m²/s. The range of the riser and downcomer dispersion coefficients were from 0.0012 to 0.0103 m²/s and 0.0324 to 0.3712 m²/s, respectively.

For a given liquid viscosity and gas velocity, higher overall, riser and downcomer liquid phase dispersion coefficients were obtained in the internal loop airlift (1) reactor compared to the internal loop airlift (2) vessel. This was primarily due to the higher liquid velocities occurring in the former, which gives rise to better liquid phase mixing.

Apparently, as far as we are aware, there are no previous studies of Bodenstein numbers and axial dispersion coefficients in airlift reactors with viscous Newtonian fluids.

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Figure 5.53: Effect of Newtonian liquid viscosity on the overall, riser and downcomer Bodenstein numbers in the internal loop airlift (1). The data points with lines represent a superficial gas velocity of 0.0160 m/s, while the symbols without the lines indicate a gas velocity of 0.0080 m/s.

Figure 5.54: Effect of Newtonian liquid viscosity on the overall, riser and downcomer Bodenstein numbers in the internal loop airlift (2). The data points with lines represent a superficial gas velocity of 0.0150 m/s, while the symbols without the lines indicate a gas velocity of 0.0075 m/s.
Figure 5.55: Effect of Newtonian liquid viscosity on the overall, riser and downcomer Bodenstein numbers in the external loop airlift. The data points with lines represent a superficial gas velocity of 0.0212 m/s, while the symbols without the lines indicate a gas velocity of 0.0106 m/s.

Figure 5.56: Effect of Newtonian liquid viscosity on the overall, riser and downcomer liquid phase axial dispersion coefficients in the internal loop airlift (1). The data points with lines represent a superficial gas velocity of 0.0160 m/s, while the symbols without the lines indicate a gas velocity of 0.0080 m/s.
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Figure 5.57: Effect of Newtonian liquid viscosity on the overall, riser and downcomer liquid phase axial dispersion coefficients in the internal loop airlift (2). The data points with lines represent a superficial gas velocity of 0.0150 m/s, while the symbols without the lines indicate a gas velocity of 0.0075 m/s.

Figure 5.58: Effect of Newtonian liquid viscosity on the overall, riser and downcomer liquid phase axial dispersion coefficients in the external loop airlift. The data points with lines represent a superficial gas velocity of 0.0212 m/s, while the symbols without the lines indicate a gas velocity of 0.0160 m/s.

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5.3.2 Non-Newtonian Fluids

The flow behaviour index and the consistency index of the CMC solutions employed in the present work are given in Tables A.1 to A.3. For a given CMC solution and airlift reactor geometry, the apparent viscosity was calculated using Eqn 3.46 for different gas velocities. Hence, for a CMC solution, a range of the apparent viscosity is also given. Low gas velocities gave low apparent viscosities, while higher gas flow rates resulted in higher apparent viscosities.

5.3.2.1 The Effect of Flow Behaviour Index on Gas Holdup

As the flow behaviour index of the CMC solution decreased, the riser and downcomer gas holdups were observed to increase for various superficial gas velocities (Figures 5.59 - 5.61). At a given flow behaviour index, a higher gas velocity gave a higher riser or downcomer gas holdup than a lower gas velocity in the three airlift vessels.

For a 30 fold increase in the consistency index of the non-Newtonian liquid in the internal loop airlift (1) contactor (Table A.1), there was a 367% increase in the riser gas holdup at a superficial gas velocity of 0.0032 m/s and only a 50% increase at a gas velocity of 0.0162 m/s. In the external loop airlift vessel, there was a 125% increase in the downcomer gas holdup for a 49 fold increase in the consistency index at a gas velocity of 0.0042 m/s while there was only a 14% increase in downcomer holdup at a higher gas velocity (0.0212 m/s). Usually, with decreasing flow behaviour index of the liquid, there was a larger impact on riser and downcomer gas holdups at lower gas velocities than at higher gas flow rates. This was similar to that observed with the Newtonian fluids (Section 5.3.1.1). It could be due to the changes in the flow regime occurring in the column with increasing gas velocity. Also, at higher apparent viscosities there was a larger amount of gas recirculating into the downcomer than at lower apparent viscosities.

The increase in the riser and downcomer gas holdups with increasing apparent liquid viscosity were mainly due to the formation of small bubbles in viscous liquids. In viscous pseudoplastic liquids two different classes of bubbles were observed: large and small bubbles. The large bubbles had a high bubble rise velocity and therefore tended to reduce the gas holdup. However, in viscous liquids over a period of time after the gas flow had been turned on a progressive increase in the concentration of small bubbles was observed. This effect seemed to be predominant over the effect of large bubbles on holdup, for the stable value of gas holdup. Since in this study only a stable value of gas holdup was
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measured, there was an increase in the gas holdup with increasing liquid viscosity. This might be surprising, since most previous studies (Deckwer, 1992; Godbole et al., 1984; Haque et al., 1986; Heijnen and Van’t Riet, 1984; Li et al., 1995; Pollard, 1995; Popovic and Robinson, 1984; Shah et al., 1982; Vatai and Tekic, 1989) reported that the gas holdup decreased as the non-Newtonian liquid viscosity was increased in either bubble columns or airlift reactors. However, these researchers have not stated whether they measured the stable or the unstable value of gas holdup at a given gas velocity and liquid viscosity. For example, Li et al. (1995) studied the local gas holdup in a draft tube internal loop airlift reactor (55 L working volume) with CMC solutions (1 to 4% weight giving $n$ from 0.617 to 0.441). They observed the riser and downcomer gas holdups to decrease with an increase in the CMC concentration. But, they did not mention whether they measured the stable or the unstable value of gas holdup and also did not offer an explanation as to why their gas holdups decrease with increasing apparent viscosity. Heijnen and Van’t Riet (1984) suggested that an increase in apparent viscosity led to the formation of large bubbles which

![Figure 5.59: The effect of flow behaviour index on riser and downcomer gas holdups in the internal loop airlift (1) for different superficial gas velocities. The data points with lines gave the riser holdup while the downcomer holdup is denoted by just the symbols.](image)

(As $n$ decreases, $k$ increases; Appendices 1 & 2)
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Figure 5.60: The effect of flow behaviour index on riser and downcomer gas holdups in the internal loop airlift (2) for different superficial gas velocities. The data points with lines gave the riser holdup while the downcomer holdup is denoted by just the symbols.

(reduced gas holdup. Vatai and Tekic (1989) reported that an increase in the apparent viscosity in a 20 cm bubble column in churn turbulent flow regime led to a decrease in the gas holdup. However, in a smaller diameter (5 cm) column the influence of the apparent viscosity was the opposite. They suggested that this might be because the transition to a slug flow regime does not appear simultaneously in the whole column. Some of the bottom part of the column was necessary for the coalescence of bubbles in order to form the slugs. This length of column decreased as the apparent liquid viscosity was increased. Pollard (1995) also reported that the overall and downcomer gas holdups in an internal loop airlift bioreactor decreased with an increase in the xanthan gum concentration (n from 0.70 to 0.30).

Some investigators (Kelkar and Shah, 1985; Schumpe and Deckwer, 1987) stated that, with increasing apparent viscosity, the gas holdup increased (to a maximum) and then decreased. Kelkar and Shah (1985) explained this on the basis of hindered gas bubble motion in viscous fluids: at relatively low gas velocities drag forces are not large enough to
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cause bubble coalescence. These moderate forces contribute to more uniform distribution of bubbles and therefore higher gas holdup. However, Kelkar and Shah (1985) and Schumpe and Deckwer (1987) have not mentioned whether they took into account the effect of small bubbles on gas holdup at high apparent viscosities (i.e. stable value of gas holdup). Guy et al. (1986) studied the overall gas holdup in a bubble column with CMC solutions. They reported that the apparent viscosity had little influence on gas holdup. They also revealed that their reported gas holdup values were obtained just after turning on the gas flow. Hence, they must have measured the unstable value of holdup.

Philip et al. (1990) did not find much difference between the overall gas holdup (stable value) obtained in an airlift vessel with 1% xanthan broth \( n = 0.16 \) and 1.25% \( n = 0.64 \) and 1.5% \( n = 0.53 \) CMC solutions. However, they reported that the gas holdup due to small bubbles was higher with the higher viscosity liquid (xanthan broth) than with the lower viscosity liquid (1.5% CMC solution).

![Figure 5.61: The effect of flow behaviour index on riser and downcomer gas holdups in the external loop airlift for different superficial gas velocities. The data points with lines gave the riser holdup while the downcomer holdup is denoted by just the symbols. (As \( n \) decreases, \( n \) increases; Appendices 1 & 2)]](image)
5.3.2.2 The Effect of Flow Behaviour Index on Liquid Velocity

Figures 5.62 to 5.64 exhibit the effect of the flow behaviour index on the riser and downcomer superficial liquid velocities in the three airlift vessels for different gas velocities. For a given liquid flow behaviour index, higher gas velocities gave higher riser and downcomer superficial liquid velocities than lower gas velocities. The downcomer superficial liquid velocity was higher than the riser superficial liquid velocity, for a given superficial gas velocity.

From Figs 5.62 to 5.64, it is evident that both the riser and downcomer superficial liquid velocities are decreasing with decreasing flow behaviour index. In the internal loop airlift (1) contactor, for a 30 fold increase in the consistency index of the CMC solution there was a 68% and a 73% decrease in the riser superficial liquid velocities at gas velocities of 0.0032 m/s and 0.0162 m/s, respectively. For a 49 fold increase in the consistency index of the CMC solution in the external loop airlift reactor, there was a 26% decrease in the riser superficial liquid velocity.

Figure 5.62: The effect of flow behaviour index on riser and downcomer superficial liquid velocities in the internal loop airlift (1) for various superficial gas velocities. The data points with lines portray the riser liquid velocity while the symbols without the lines represent the downcomer liquid velocity.

(*As n decreases, κ increases; Appendix 1+2*)
and a 33% decrease in the downcomer superficial liquid velocities at gas velocities of 0.0042 m/s and 0.0212 m/s, respectively. At a higher gas velocity, there was a slightly larger increase in the liquid velocity than at a lower gas velocity for increasing apparent viscosity. This is opposite to what was observed with glycerol solutions.

Both the riser and downcomer superficial liquid velocities were decreasing with increasing apparent viscosity at a fixed superficial gas velocity. This was mainly due to the increasing frictional resistance to flow in the airlift contactor with an increase in the apparent viscosity of the liquid phase and therefore increasing energy losses occurring in the column. It also could be because of increased gas recirculation into the downcomer with an increase in the apparent viscosity (Table 5.2). This might lead to a lower driving force for liquid circulation with increasing apparent viscosity. Most previous studies (Allen and Robinson, 1989; Glennon et al., 1988; Li et al., 1995; Philip et al., 1990; Pollard, 1995; Popovic and Robinson, 1984; 1988; Weiland, 1984) have also observed the liquid velocity to decrease with an increase in the apparent viscosity. Li et al. (1995) found the

![Graph](image)

Figure 5.63: The effect of flow behaviour index on riser and downcomer superficial liquid velocities in the internal loop airlift (2) for various superficial gas velocities. The data points with lines portray the riser liquid velocity while the symbols without the lines represent the downcomer liquid velocity.

(R, n decreases, k increases; Appendixes 14-2)
apparent viscosity to have a pronounced effect on the riser liquid velocity in their airlift column. With increasing concentrations of CMC solutions, the riser liquid velocity was observed to substantially decrease. Pollard (1995) also observed the riser liquid velocity to sharply decrease with increasing concentration of xanthan gum.

There are some earlier works which have reported results contrary to the present findings. For example, Wachi et al. (1991) reported that CMC solutions (0.15% and 0.3% wt.) apparently enhanced the liquid circulation velocity in their airlift vessel. While, Fields et al. (1984) also observed increased liquid velocities with xanthan gum. Russell (1989) studied the liquid velocity in an airlift bioreactor with a mycelial fermentation broth (*P. chrysogenum*). He noticed that at low apparent viscosities the liquid velocity was enhanced by an increase in viscosity, whereas at higher viscosities the liquid velocity was impeded as the apparent viscosity increased. However, in the same airlift reactor Pollard (1995) observed the liquid velocity to be independent of the apparent viscosity of a fermentation broth.

Figure 5.64: The effect of flow behaviour index on riser and downcomer superficial liquid velocities in the external loop airlift for various superficial gas velocities. The data points with lines portray the riser liquid velocity while the symbols without the lines represent the downcomer liquid velocity.
medium (*S. erythraea*). He stated that the liquid velocity can be improved by the enhancement of the energy dissipation rate with viscous liquids.

### 5.3.2.3 The Effect of Flow Behaviour Index on Liquid Mixing

The effect of flow behaviour index on the mixing and circulation times in the three geometries of airlift reactors are given in Figures 5.65 to 5.67. It is apparent that the liquid mixing and circulation times increase with decreasing flow behaviour index. In the external loop airlift contactor, for a 49 fold increase in the consistency index of the non-Newtonian fluid there was a 97% and a 129% increase in the mixing time at gas velocities of 0.0042 m/s and 0.0212 m/s, respectively. The liquid circulation time increased, respectively, by 116% and 221% at gas velocities of 0.0030 m/s and 0.0150 m/s in the internal loop airlift (2) vessel for a 32 fold increase in the consistency index of the liquid. It is apparent that a higher gas flow rate had a larger impact on the rate of increase of mixing and circulation times with increasing apparent viscosity than a lower gas flow rate. This was due to the fact

![Graph](image_url)

**Figure 5.65:** The effect of flow behaviour index on the mixing and circulation times in the internal loop airlift (1) for several superficial gas velocities. The data points with lines give the mixing time, while the symbols without the lines represent the circulation time. *(As n decreases, k increases; Appendices 1&2)*
that, with increasing apparent viscosity, a lower gas velocity had a smaller impact on the liquid velocity than a higher gas flow rate.

The primary factor behind the cause of increasing mixing and circulation times with decreasing flow behaviour index was the liquid velocity. The liquid velocity was observed to decrease with decreasing flow behaviour index (Section 5.3.1.2) and this had a direct effect on the mixing and circulation times.

Most studies (Guy et al., 1986; Haque et al., 1986; Kawase et al., 1994; Li et al., 1995; Pandit and Joshi, 1983; Popovic and Robinson, 1993) have reported results similar to the present findings. Kawase et al. (1994) found the mixing time to increase with an increase in the shear thinning of the CMC solutions. They also reported that the viscous non-Newtonian flow behaviour had an insignificant effect on the liquid circulation time. This does not agree with the present results. Since mixing time and circulation time are closely related, it is unclear as to how Kawase et al. found the apparent viscosity to influence one and not the other. Li et al. (1995) found the apparent viscosity to have a
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major impact on the liquid circulation time. They observed an increase in the circulation
time with an increase in the CMC concentration in an internal loop airlift reactor. Popovic
and Robinson (1993) showed the specific mixing time to be highly dependent on the
apparent viscosity of the liquid phase. They reported that the specific mixing time
increased with an increase in the apparent viscosity.

Velan and Ramanujam (1995) studied the mixing time in an internal loop airlift
reactor with CMC solutions (0.1 to 0.75% wt. giving $n$ from 0.86 to 0.74). The mixing time
was observed to decrease with an increase in the concentration of the CMC solution. This
is contrary to the present results. They reported that with a decrease in the flow behaviour
index, large bubble formation took place and this enhanced the liquid circulation velocity.
Hence, the liquid circulation time decreased. Since the circulation time is usually
proportional to the mixing time, the mixing time also decreased. Glennon et al. (1989)
measured mixing times in external loop airlift reactors with xanthan gum solutions ($n =$
0.66 and 0.53). They also observed the mixing time to be lower in viscous non-Newtonian

![Figure 5.67: The effect of flow behaviour index on the mixing and circulation times in the external loop airlift for several superficial gas velocities. The data points with lines give the mixing time, while the symbols without the lines represent the circulation time. (As $n$ decreases, $n$ increases; Appendices 10-2)](image-url)
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fluids. Fields et al. (1984) investigated the circulation time in an internal loop airlift reactor with xanthan gum (0 - 0.5% wt). Apparently, the circulation time varied in a complex pattern with increasing gum concentration. They observed at low concentrations the circulation time was faster than for water (due to drag reduction) whilst at the highest concentrations liquid circulation times were increased.

The ratio of the mixing to circulation time versus the flow behaviour index in the three airlift contactors are shown in Figures 5.68 to 5.70. It is clear from the figures that the flow behaviour index of the liquid does not have an effect on the ratio of the liquid mixing to circulation time. This implied that the rate of mixing in the airlift contactor was independent of the apparent viscosity.

On average, 1.1 circulations of the vessel were required in all three airlift contactors to disperse a hot liquid pulse to 90% homogeneity. The low ratio of mixing to circulation time implied that there was relatively good mixing in the three airlift contactors, with the CMC solutions.

![Graph](image)

Figure 5.68: The effect of flow behaviour index on the ratio of liquid mixing to liquid circulation time in the internal loop airlift (1) for several superficial gas velocities.

(as n decreases, k increases; Appendices W2)
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Figure 5.69: The effect of flow behaviour index on the ratio of liquid mixing to liquid circulation time in the internal loop airlift (2) for several superficial gas velocities.
(as n decreases, k increases; Appendices 1&2)

Figure 5.70: The effect of flow behaviour index on the ratio of liquid mixing to liquid circulation time in the external loop airlift for several superficial gas velocities.
(as n decreases, k increases; Appendices 1&2)
All three Bodenstein numbers (overall, riser and downcomer) increased with an increase in the apparent liquid viscosity (Figs 5.71 - 5.73). Generally, at a specified gas flow rate, the riser Bo number was the highest and the downcomer Bo number was the lowest. This implied that the mixing in the downcomer section was better than the mixing in the riser section for a given gas flow rate. This is verified by Figures 5.74 to 5.76, which show the effect of the flow behaviour index on the overall, riser and downcomer dispersion coefficients. It is apparent that for a given gas velocity, the dispersion in the downcomer section was the highest while the dispersion in the riser section was the lowest.

The overall, riser and downcomer dispersion coefficients were observed to decrease with decreasing flow behaviour index of the liquid (Figs 5.74 - 5.76). This was mainly due to the fact that the liquid velocity in the airlift was also decreasing with increasing apparent liquid viscosity (Section 5.3.1.2). A higher superficial gas velocity in the column gave a higher axial dispersion coefficient (overall, riser and downcomer) and hence a lower Bodenstein number than a lower gas velocity.

Generally, for all gas velocities used in this study, in the internal loop airlift (1) contactor with CMC solutions the overall Bodenstein numbers were between 17 to 32 while the riser Bo numbers were from 24 to 97. However, the downcomer Bo numbers were much lower (from 4 to 7). In the internal loop airlift (2) vessel with CMC solutions, the overall Bo numbers were from 24 to 55 for all gas velocities, while the riser Bo numbers were between 30 to 134. The range of downcomer Bo numbers was from 2 to 13. The range of overall Bo numbers was from 17 to 45 in the external loop airlift contactor for all gas velocities with CMC solutions. The range of riser Bo numbers was slightly higher (24 - 61) while the downcomer Bo number range was much lower (2 - 14).

In the internal loop airlift (1) reactor with solutions of CMC, for the gas velocities employed in this work, the overall liquid phase axial dispersion coefficients (D) were from 0.0093 to 0.0311 m²/s. The range of the riser D was from 0.0012 to 0.0124 m²/s, while the downcomer D range was higher (0.0227 - 0.0851 m²/s). The overall liquid phase axial dispersion coefficients in the internal loop airlift (2) contactor with CMC solutions were between 0.0033 and 0.0410 m²/s, while the riser D range was 0.002 to 0.0142 m²/s. The range of the downcomer D was from 0.0098 to 0.0854 m²/s. The range of overall D, in the external loop airlift vessel with CMC solutions, was from 0.0067 to 0.0541 m²/s. The range of the riser D was between 0.001 and 0.0148 m²/s, while the downcomer D range was from 0.0011 to 0.2813 m²/s.
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In the literature, there is only one experimental data set of local axial dispersion coefficients in airlift reactors with non-Newtonian fluids. Li et al. (1993) studied the liquid phase axial dispersion in a draft tube internal loop airlift bioreactor with CMC solutions (1 - 4% wt giving n from 0.441 to 0.617). They reported that the riser Peclet (Pe) number decreased with an increase in the apparent viscosity for a given gas flow rate. Conversely, for a fixed gas velocity, the riser and downcomer dispersion coefficients increased with an increase in the concentration of the CMC solution. Li et al. stated that in highly viscous non-Newtonian liquids, for a given gas velocity, the liquid circulation velocity is low but the large bubble rise velocity is high. Apparently, this gives a high D or low Pe number with highly viscous non-Newtonian fluids. This is contrary to the present findings. A possible explanation might be that Li et al. measured the axial dispersion coefficient immediately after the gas flow was changed. In this work, with viscous liquids, the dispersion coefficient was only measured after a stable value of gas holdup was achieved. Another reason could be the method used by Li et al. to obtain the dispersion coefficient (the Kolmogoroff's theory of isotropic turbulence was employed). However, Li et al. observed the axial dispersion coefficient in the downcomer to be higher than in the riser section which is in agreement with the present results. Their downcomer dispersion coefficients were between 0.010 and 0.090 m²/s, while the riser D was from 0.010 to 0.065 m²/s for CMC solutions. Li et al. employed a superficial gas velocity range of 0.01 to 0.10 m/s. With CMC solutions, Li et al. obtained riser Pe numbers between about 5 and 43. Glennon et al. (1988) obtained Bo numbers from about 48 to 101 in a 300 L airlift vessel with xanthan gum solutions. Their gas velocity was from about 0.01 to 0.05 m/s.

There are many studies of axial dispersion in bubble columns with viscous non-Newtonian liquids (see Kelkar and Shah, 1985 and Shah et al., 1982 for a list of the literature). Kelkar and Shah (1985) studied the backmixing in bubble columns with CMC solutions (n from 0.66 to 0.99). They reported that the axial dispersion coefficients increased with an increase in gas velocity in the bubble column. They also found the apparent viscosity to have a considerable effect on the dispersion coefficient. The dispersion coefficients were observed to decrease with an increase in the apparent viscosity. This is in agreement with the present results. They explained the decrease in the dispersion coefficient with respect to viscosity on the basis of an increase in the bubble
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Figure 5.71: The effect of flow behaviour index on the overall, riser and downcomer Bodenstein numbers in the internal loop airlift (1). The data points with lines represent a superficial gas velocity of 0.0160 m/s, while the symbols without the lines indicate a gas velocity of 0.0080 m/s.

(As \( n \) decreases, \( k \) increases; Appendices 1b2)

Figure 5.72: The effect of flow behaviour index on the overall, riser and downcomer Bodenstein numbers in the internal loop airlift (2). The data points with lines represent a superficial gas velocity of 0.0150 m/s, while the symbols without the lines indicate a gas velocity of 0.0075 m/s.

(As \( n \) decreases, \( k \) increases; Appendices 1b2)
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Figure 5.73: The effect of flow behaviour index on the overall, riser and downcomer Bodenstein numbers in the external loop airlift. The data points with lines represent a superficial gas velocity of 0.0212 m/s, while the symbols without the lines indicate a gas velocity of 0.0106 m/s.

(As $n$ decreases, $K$ increases; Appendices 1+2)

Figure 5.74: The effect of flow behaviour index on the overall, riser and downcomer liquid phase axial dispersion coefficients in the internal loop airlift (1). The data points with lines denote a superficial gas velocity of 0.0160 m/s, while the symbols without the lines indicate a gas velocity of 0.0080 m/s.

(As $n$ decreases, $K$ increases; Appendices 1+2)
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Figure 5.75: The effect of flow behaviour index on the overall, riser and downcomer liquid phase axial dispersion coefficients in the internal loop airlift (2). The data points with lines denote a superficial gas velocity of 0.0150 m/s, while the symbols without the lines indicate a gas velocity of 0.0075 m/s.

Figure 5.76: The effect of flow behaviour index on the overall, riser and downcomer liquid phase axial dispersion coefficients in the external loop airlift. The data points with lines denote a superficial gas velocity of 0.0212 m/s, while the symbols without the lines indicate a gas velocity of 0.0106 m/s.

(As n decreases, k increases; Appendices 1-2)
size. At higher liquid viscosity, the bubbles carry less amount of liquid in their wake which led to a decrease in the backmixing.

5.3.3 Summary

The stable value of riser and downcomer gas holdups monotonically increased with an increase in the Newtonian and apparent liquid phase viscosity. With increasing liquid viscosity (Newtonian and non-Newtonian) the local liquid velocities were observed to decrease sharply in the airlift reactors. This had a profound effect on the mixing and circulation times, which increased with an increase in the liquid phase viscosity. However, the ratio of the mixing to circulation time was not influenced by the liquid phase viscosity (Newtonian and non-Newtonian). The overall, riser and downcomer Bodenstein numbers were found to increase with increasing Newtonian and non-Newtonian liquid phase viscosity. Consequently, the overall, riser and downcomer dispersion coefficients decreased with an increase in the liquid viscosity. For a given liquid viscosity (Newtonian or non-Newtonian) and superficial gas velocity, the liquid phase axial dispersion coefficient in the downcomer was higher than that in the riser section. While the value of the overall dispersion coefficient was in-between the riser and the downcomer values, for a fixed liquid viscosity and gas velocity.
5.4 The Effect of Two-Liquid Systems on Airlift Hydrodynamics and Mixing

The effect of two-liquid phases on airlift hydrodynamics and liquid mixing is investigated in this section. Different concentrations of oil-in-water emulsions were employed as the two immiscible liquids. Tables A.1 to A.3 give the physical properties of the different oil-in-water emulsions employed in the three airlift reactors. The physical properties of the pure oil are also given in Tables A.1 to A.3.

5.4.1 The Effect of Oil-in-water Emulsion on Gas Holdup

The riser and downcomer gas holdups are plotted against the volume content of oil in the liquid-liquid suspension in Figures 5.77 to 5.79. A zero percent oil content in the emulsion indicates that the fluid is pure water (i.e. one liquid phase only). On the other hand, a hundred percent oil content in the emulsion means that the liquid is pure oil. On addition of a relatively small amount of oil (10% by volume) to the pure water system (continuous phase) results in a sharp increase in the riser and downcomer gas holdups for a fixed gas velocity. As the concentration of oil in water increases (from 10 vol. % to 60 vol. %), the riser and downcomer gas holdups drop as shown in Figures 5.77 to 5.79. Then both the riser and downcomer gas holdups increase as the pure oil limit is reached. It must be mentioned that pure oil yielded higher riser and downcomer gas holdups than pure water, for a given gas velocity, in two phase airlift reactors. This was mainly due to the high viscosity of the oil, which gave rise to large and small bubbles. Hence, only a stable value of the gas holdup was measured.

As the dispersed phase (oil) was increased in the continuous phase (water), from 10 to 60 vol. %, the emulsion of two immiscible liquids changed from Newtonian behaviour to non-Newtonian behaviour. That is the flow behaviour index of the emulsions decreased with an increase in the dispersed phase concentrations. Conversely, the consistency index increased as the oil-in-water emulsion concentration increased (i.e. the viscosity of the oil-in-water emulsions increase with increasing dispersed phase concentration). Pal et al. (1986) also noted that oil-in-water emulsions changed from Newtonian fluids to non-Newtonian fluids as the concentration of the dispersed phase (oil) increased in the continuous phase (water).

The riser and downcomer gas holdups in a three phase system, might be expected to increase with increasing liquid viscosity (as in a gas-liquid system). However, it is clear from Figures 5.77 to 5.79 that this is not the case. It seems that the effect of two immiscible
On addition of oil (10%) to the air-water system the gas holdup showed a sharp increase (Figs 5.77 - 5.79). This could be due to the presence of another interface (air-oil) which significantly increased the gas content in the three fluid system.

As the oil in water emulsion concentration increased from 10 to 60% (i.e. the apparent viscosity of the emulsion increased), the riser and downcomer gas holdups decreased. Siquier et al. (1991) have reported that the reduction in gas holdup with immiscible liquids could be attributed to the increase in the apparent viscosity of the liquids due to the presence of dispersed drops. While Pino et al. (1990) stated that increasing the liquid viscosity yields faster bubble coalescence rates and larger stable bubbles which results in a lower gas residence time. However, Arguelles et al. (1993) have commented that the variety of the observations previously mentioned together with the possible interactions at the different fluid-fluid interfaces indicates that this might be too simple an explanation for the phenomenon.

It must be remembered that in this system at 50% oil/water emulsion a phase inversion takes place. That is initially the oil is the dispersed phase and water is the continuous phase; but at 50% oil/water emulsion the later (water) becomes the dispersed phase and the former (oil) turns into the continuous phase.

From 60% oil/water emulsion to pure oil (100%) the gas holdups increased sharply mainly due to the increasing viscosity of the liquid phase. This was in agreement with the observations made with glycerol and CMC solutions in this study (Section 5.3).
liquids on gas holdup in airlift reactors is complicated due to the interaction of three different fluid-fluid interfaces.

In the literature, as stated in Section 2.7, there is no experimental data of gas holdup in airlift reactors with two immiscible fluids. The results in bubble columns suggest that the overall gas holdup decreases upon adding an immiscible liquid to a gas/liquid system (Arguelles et al., 1993; Hatzikiriakos et al., 1990). Arguelles et al. (1993) observed, for water-kerosene system, a minimum in the gas holdup at approximately when the inversion in the continuity of the liquid phases occurred. They stated that when water was the dispersed phase, relatively large drops of water resulted in a variation of water content along the length of the column. Conversely, when kerosene was the dispersed phase the drop size was very small and therefore a uniform distribution of kerosene drops was obtained.

![Figure 5.77](image_url)

Figure 5.77: The effect of oil-in-water concentration on riser and downcomer gas holdups in the internal loop airlift (1) for different superficial gas velocities. The data points with lines symbolise the riser holdup, while the downcomer holdup is expressed by just the symbols.
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Figure 5.78: The effect of oil-in-water concentration on riser and downcomer gas holdups in the internal loop airlift (2) for different superficial gas velocities. The data points with lines represent the riser holdup, while the downcomer holdup is expressed by just the symbols.

For a given oil-in-water system, a higher superficial gas velocity gave a higher riser or downcomer holdup than a lower gas velocity. There was a 102% increase in the riser gas holdup at a superficial gas velocity of 0.0030 m/s upon addition of 10% oil to the air/water system in the internal loop airlift (2) contactor. While, at a higher gas velocity (0.0150 m/s) there was only a 42% increase. This phenomenon was also observed in the other two geometries of airlifts. It seems that the effect is more pronounced at a lower gas velocity than at a higher gas flow rate. In the internal loop airlift (1) contactor, from 10 to 60% oil-in-water emulsion, there was a 56% decrease in the riser gas holdup at a superficial gas velocity of 0.0037 m/s while only a 30% drop was observed at a higher gas velocity (0.0148 m/s). In the external loop airlift vessel, at a gas velocity of 0.0042 m/s there was a 68% drop in the downcomer holdup from 10 to 60% oil-in-water emulsion while a decrease of 43% occurred at a higher gas velocity (0.0212 m/s). It is apparent that at lower gas velocities there is a larger impact on the riser and downcomer gas holdups than at
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higher gas flow rates. However, from 60% oil-in-water emulsion to pure oil, in the internal loop airlift (2) vessel at a gas velocity of 0.0030 m/s, there was a 25% increase in the riser gas holdup. For the same system, at a higher gas velocity (0.0150 m/s) there was a 55% increase in the riser gas holdup. In the external loop airlift reactor, from oil-in-water (60%) emulsion to pure oil, there was a 85% increase in the downcomer holdup at a gas velocity of 0.0042 m/s while at a higher gas velocity (0.0212 m/s) an increase of 120% occurred. It is apparent that, for the change in liquids from 60% oil-in-water emulsion to pure oil, there was a more pronounced effect on the riser and downcomer gas holdups at a higher gas velocity than at a lower air flow rate.

Figure 5.79: The effect of oil-in-water concentration on riser and downcomer gas holdups in the external loop airlift for different superficial gas velocities. The data points with lines represent the riser holdup, while the downcomer holdup is signified by the symbols.
5.4.2 The Effect of Oil-in-water Emulsion on Liquid Velocity

Both the riser and downcomer superficial liquid velocities are decreasing with increasing oil-in-water concentration in the three airlift contactors (Figs 5.80 - 5.82). On addition of an immiscible liquid (10% oil) to an air/water system in the internal loop airlift (1) vessel, there was a 2% drop in the riser superficial liquid velocity at a gas velocity of 0.0037 m/s. While, at a higher gas velocity (0.0184 m/s) the riser liquid velocity decreased by about 13%. Similar trend was also observed in the other two airlift contactors. In the external loop airlift reactor, from pure water to 10% oil-in-water emulsion, there was a 11% reduction in the downcomer superficial liquid velocity at a gas velocity of 0.0042 m/s while at a higher gas velocity (0.0212 m/s) a lower reduction (4%) occurred.

In the internal loop airlift (2) contactor, there was a 74% decline in the riser superficial liquid velocity from 10% to 60% oil-in-water emulsion at a gas velocity of 0.0030 m/s and a 43% reduction at a higher gas velocity (0.0150 m/s). In the external loop airlift vessel, from 10% to 60% oil-in-water emulsion, there was a 27% drop in the

Figure 5.80: The effect of oil-in-water concentration on riser and downcomer superficial liquid velocities in the internal loop airlift (1) for various superficial gas velocities. The data points with lines portray the riser liquid velocity, while the symbols without the lines represent the downcomer liquid velocity.
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downcomer superficial liquid velocity at a gas velocity of 0.0042 m/s while a lower reduction (20%) was observed at a higher gas velocity (0.0212 m/s).

From 60% oil-in-water emulsion to pure oil, in the internal loop airlift (1) vessel at a gas velocity 0.0037 m/s, the riser superficial liquid velocity decreased by 70%. While, at a higher gas velocity (0.0184 m/s) there was a slightly lower reduction (63%). In the internal loop airlift (2) reactor, from 60% oil-in-water emulsion to pure oil at a gas velocity of 0.0030 m/s, there was a reduction of 71% in the downcomer liquid velocity. And at a higher gas velocity (0.0150 m/s) the downcomer liquid velocity decreased by almost the same amount (70%). It is apparent that there is a slightly more pronounced effect on the liquid velocity at a lower gas flow rate than at a higher gas velocity. This could be due to the change in the flow regime occurring in the airlift vessel with increasing gas velocity.

Figure 5.81: The effect of oil-in-water concentration on riser and downcomer superficial liquid velocities in the internal loop airlift (2) for various superficial gas velocities. The data points with lines portray the riser liquid velocity, while the symbols without the lines represent the downcomer liquid velocity.
Figure 5.82: The effect of oil-in-water concentration on riser and downcomer superficial liquid velocities in the external loop airlift for various superficial gas velocities. The data points with lines portray the riser liquid velocity, while the symbols without the lines represent the downcomer liquid velocity.

Liquid velocity is decreasing with increasing oil-in-water emulsions mainly because of the increasing viscosity of the liquids. With increasing liquid viscosity, the frictional resistance to flow in the airlift vessel is also increasing and this leads to a reduction in the liquid velocity. Also, the amount of riser gas recirculating into the downcomer is increasing with increasing liquid phase viscosity (Table 5.2). This leads to a reduction in the driving force for liquid circulation in the airlift vessel and hence a lower liquid velocity.

For a given oil-in-water emulsion, higher superficial gas velocities gave higher riser and downcomer superficial liquid velocities than lower gas velocities. The downcomer superficial liquid velocity was higher than the riser superficial liquid velocity for a given gas velocity (Figs 5.80 - 5.82).
5.4.3 The Effect of Oil-in-water Emulsion on Liquid Mixing

Figures 5.83 to 5.85 show the effect of oil-in-water emulsion on the mixing and circulation times for several superficial gas velocities. It is evident that with increasing oil in water concentration, the mixing and circulation times are also increasing. In the internal loop airlift (1) reactor, from pure water to 10% liquid-liquid phase, there was a 4% increase in the mixing time at a gas velocity of 0.0037 m/s. While at a higher gas velocity (0.0184 m/s) a larger growth (19%) occurred. A similar trend was also observed in the other two geometries of airlift contactors. However, the circulation time showed only a very slight increase (1%), from pure water to 10% oil-in-water emulsion in the internal loop airlift (2) vessel, at gas velocities of 0.0030 m/s and 0.0150 m/s.

There was a 35% increase in the mixing time from 10% to 60% oil-in-water emulsion at a gas velocity of 0.0037 m/s in the internal loop airlift (1) vessel, while a 33% increase at a higher gas velocity (0.0184 m/s) occurred. Similarly, with the other two airlift geometries.

![Diagram](image_url)

Figure 5.83: The effect of oil-in-water concentration on the mixing and circulation times in the internal loop airlift (1) for several superficial gas velocities. The data points with lines give the mixing time, while the symbols without the lines represent the circulation time.
vessels there were a slightly higher increase in the mixing time at a lower gas velocity than at a higher gas velocity. This phenomenon was also observed with the circulation time. For example, in the external loop airlift contactor, there was a 31% increase in the circulation time at a gas velocity of 0.0212 m/s, while a 51% increase occurred at a lower gas velocity (0.0042 m/s). This was due to the more obvious effect on the liquid velocity at a lower gas velocity than at a higher gas flow rate.

A much steeper increase occurred in the mixing and circulation times at all gas flow rates from 60% oil-in-water emulsion to pure oil. In the internal loop airlift (1) contactor, from 60% oil-in-water emulsion to pure oil at a gas velocity of 0.0037 m/s, the mixing time increased by about 246%. At a higher gas velocity (0.0184 m/s), there was a reduction in the rate of increase of the mixing time (196%). The circulation time was observed to increase by about 390% at a gas velocity of 0.0030 m/s in the internal loop airlift (2) vessel for a change in the liquids from 60% liquid-liquid phase to pure oil. While at a higher gas velocity (0.0150 m/s), there was a 321% increase.

Figure 5.84: The effect of oil-in-water concentration on the mixing and circulation times in the internal loop airlift (2) for several superficial gas velocities. The data points with lines give the mixing time, while the symbols without the lines represent the circulation time.
Results and Discussion

Figure 5.85: The effect of oil-in-water concentration on the mixing and circulation times in the external loop airlift for several superficial gas velocities. The data points with lines give the mixing time, while the symbols without the lines represent the circulation time.

Both the mixing and circulation times in the airlift vessels were mainly dependent on the liquid velocity in the column. Since, with increasing oil-in-water concentration the superficial liquid velocity was decreasing (Section 5.4.2), thus the mixing and circulation times will be expected to increase. There seems to be no previous studies in the open literature of mixing time or circulation time with two immiscible liquids in either bubble columns or airlift reactors.

It is clear from Figures 5.86 to 5.88 that the concentration of oil-in-water emulsions do not have an effect on the ratio of the mixing to circulation time in the airlift contactors. This indicated that the rate of mixing in the airlift contactor was independent of the oil-in-water concentration. On average, in the internal loop airlift vessels (1) and (2), about 1.2 circulations of the vessel were required to disperse a hot liquid pulse to 90% homogeneity. While in the external loop airlift contactor, a higher average number of circulations (2.2) were needed to scatter the pulse to 90% homogeneity.
Results and Discussion

Figure 5.86: The effect of oil-in-water concentration on the ratio of liquid mixing to liquid circulation time in the internal loop airlift (1) for several superficial gas velocities.

Figure 5.87: The effect of oil-in-water concentration on the ratio of liquid mixing to liquid circulation time in the internal loop airlift (2) for several superficial gas velocities.
Results and Discussion

Superficial gas velocity (m/s)

<table>
<thead>
<tr>
<th>Velocity (m/s)</th>
<th>0.0042</th>
<th>0.0106</th>
<th>0.0169</th>
<th>0.0212</th>
</tr>
</thead>
<tbody>
<tr>
<td>Volume of oil content in emulsion (%)</td>
<td>20</td>
<td>40</td>
<td>60</td>
<td>80</td>
</tr>
</tbody>
</table>

Figure 5.88: The effect of oil-in-water concentration on the ratio of liquid mixing to liquid circulation time in the external loop airlift for several superficial gas velocities.

The Bodenstein numbers (overall, riser and downcomer) increased with an increase in the concentration of oil-in-water emulsions as shown in Figs 5.89 to 5.91. A higher gas velocity gave lower Bodenstein numbers (overall, riser or downcomer) than a lower gas velocity. At a given gas velocity, the riser Bo numbers were the highest and the downcomer Bo numbers were the lowest. This implied that the mixing in the downcomer section was higher than the mixing in the riser section. This was confirmed by the axial dispersion coefficient studies (Figs 5.92 - 5.94) which gave the dispersion in the downcomer to be higher than that in the riser section for a given gas velocity. The overall dispersion coefficient value, for a specified gas flow rate, was between the riser and the downcomer value. This indicated that the overall mixing in the airlift column was somewhere in-between the mixing in the riser section and the mixing in the downcomer section.

In the internal loop airlift (1) reactor, ordinarily for all gas velocities with pure water and oil and oil-in-water emulsions, the overall Bodenstein numbers were between 20 to 48 while the riser Bo numbers were from 36 to 136. But, the downcomer Bo numbers were substantially lower (from 3 - 17). The range of overall Bo numbers in the internal
loop airlift (2) vessel for all gas velocities was from 25 to 73 while the downcomer Bo
range was from 3 to 17. The riser Bo number range was higher (46 - 167) than the overall
and the downcomer ranges. In the external loop airlift contactor, the overall Bo number
range was from 20 to 74 for all gas velocities. The riser Bo number range, for all gas
velocities, was from 28 to 102 while the downcomer Bo range was considerably lower
(from 2 - 17).

The overall, riser and downcomer dispersion coefficients decrease with increasing
oil-in-water emulsion concentration (Figs 5.92 - 5.94) indicating that the mixing was
deteriorating. This is primarily due to the decreasing liquid velocity with increasing oil-in-
water emulsion concentration. Since the dispersion coefficients were mainly influenced by
the velocity of the liquid, hence turbulence, in the airlift column.

For all gas velocities used in this study, with oil-in-water emulsions and pure water
and oil, the overall axial dispersion coefficient (D) in the internal loop airlift (1) reactor
was within the range of 0.0040 to 0.1364 m²/s. The riser D range was from 0.0008 to
0.1700 m²/s, while the downcomer D range was higher (0.0170 - 0.5031 m²/s). In the
internal loop airlift (2) contactor, for the same liquids, the range of the overall D was from
0.0029 to 0.0395 m²/s. The riser D range was within 0.0003 to 0.0293 m²/s and the range
of the downcomer was from 0.0140 to 0.4459 m²/s. The range of overall D with oil-in-
water emulsions and pure oil and water in the external loop airlift contactor was from 0.002
to 0.0989 m²/s. While the riser D range was lower (0.0010 - 0.0134 m²/s) and the range of
the downcomer D was higher (0.0524 - 0.5844 m²/s) than the overall D range.

There are only a small number of studies of Bodenstein numbers in bubble columns
with two immiscible liquids (Arguelles et al., 1993; Pino et al., 1990). Arguelles et al.
(1993) found that the overall Bo number was not significantly affected by the water content
in the liquid-liquid suspension. This seems to contradict the findings of this study. However, Arguelles et al.’s used a different approach, from that used in this study, to
obtain their Bo numbers (they employed a mass balance on the droplets in a bubble
column). Also, their Bo number range was very small (from 0.9 to 3.7) and they have
pointed out that their Bo numbers were subjected to appreciable experimental error (since
they represented a slope in a semi-logarithmic plot of experimental data). Arguelles et al.
(1993) also found the Bo numbers to slightly increase with increasing gas velocity. They
stated that this would lead to a higher dispersion coefficient with a higher gas velocity.
However, they did not calculate the dispersion coefficients in their column.
Figure 5.89: The effect of oil-in-water concentration on the overall, riser and downcomer Bodenstein numbers in the internal loop airlift (1). The data points with lines represent a superficial gas velocity of 0.0187 m/s, while the symbols without the lines indicate a gas velocity of 0.0093 m/s.

Figure 5.90: The effect of oil-in-water concentration on the overall, riser and downcomer Bodenstein numbers in the internal loop airlift (2). The data points with lines represent a superficial gas velocity of 0.0151 m/s, while the symbols without the lines indicate a gas velocity of 0.0076 m/s.
Results and Discussion

Figure 5.91: The effect of oil-in-water concentration on the overall, riser and downcomer Bodenstein numbers in the external loop airlift. The data points with lines represent a superficial gas velocity of 0.0212 m/s, while the symbols without the lines indicate a gas velocity of 0.0106 m/s.

Figure 5.92: The effect of oil-in-water concentration on the overall, riser and downcomer liquid phase axial dispersion coefficients in the internal loop airlift (1). The data points with lines denote a superficial gas velocity of 0.0187 m/s, while the symbols without the lines indicate a gas velocity of 0.0093 m/s.
Figure 5.93: The effect of oil-in-water concentration on the overall, riser and downcomer liquid phase axial dispersion coefficients in the internal loop airlift (2). The data points with lines denote a superficial gas velocity of 0.0151 m/s, while the symbols without the lines indicate a gas velocity of 0.0076 m/s.

Figure 5.94: The effect of oil-in-water concentration on the overall, riser and downcomer liquid phase axial dispersion coefficients in the external loop airlift. The data points with lines denote a superficial gas velocity of 0.0212 m/s, while the symbols without the lines indicate a gas velocity of 0.0106 m/s.
5.4.4 Summary

The effect of two immiscible liquid phases on the local gas holdups was complex. On addition of an immiscible liquid to an air/water system, the riser and downcomer gas holdups increase. However as the concentration of oil increases, the gas holdups decrease. Then as the pure oil limit was reached the riser and downcomer gas holdups were observed to increase. As the oil in water concentration increases, the riser and downcomer superficial liquid velocities decrease monotonically. Hence, the mixing and circulation times increased with an increase in the oil-in-water concentration. But, the ratio of the mixing to circulation time was not influenced by the two immiscible liquid phases. The overall, riser and downcomer Bodenstein numbers were observed to increase with increasing concentration of oil-in-water. While the dispersion coefficients (overall, riser and downcomer) decreased with an increase in the oil-in-water concentration. For a given oil-in-water concentration and gas velocity, the dispersion coefficient in the downcomer was higher than that in the riser section. The overall dispersion coefficient was in-between the riser and the downcomer coefficient for a given gas flow rate and oil-in-water concentration.
5.5 Comparison of Experimental Data with Model Predictions

In this section, the experimental riser gas holdup and superficial liquid velocity values in the three airlift contactors are compared with predictions from the hydrodynamic model (Section 3.1). Suitable experimental data of riser gas holdup and liquid velocity from the literature are also compared with the hydrodynamic model predictions. The experimental liquid circulation time data are also compared with the predictions from the liquid circulation time correlation (Section 3.3.1).

5.5.1 Gas Holdup and Liquid Velocity

For a given gas velocity, airlift vessel geometry and fluid properties, the riser gas holdup and superficial liquid velocity can be estimated following the procedure outlined in Figure 3.7. Three parameters must be inserted into the hydrodynamic model (Section 3.1.2) for the model to predict the gas holdup and the liquid velocity.

For the internal loop airlifts (1) and (2), in this study, the distribution parameter (Section 3.1.2.1) was taken to be 0.7. The value of the distribution parameter for the external loop airlift contactor was slightly higher at 0.9. Zuber and Findlay (1965) reported the distribution parameter was below unity if the concentration at the centre line was smaller than that close to the wall. In their draft-tube internal loop airlift reactor with a ring sparger, Ayazi Shamlou et al. (1994) employed a value of 0.7 for the distribution parameter. They stated that a C_o value of 0.7 gave a better agreement between model predictions and experimental data. Similarly, in this work, with the internal loop airlifts it was discovered that a C_o value of 0.7 gave a better agreement between experimental results and model predictions. With the external loop airlift contactor a slightly higher value of C_o (0.9) gave a good agreement. This was not surprising since in this work also, a ring sparger was used in the airlift contactors. This meant that the air was ejected close to the riser wall and nearer the sparger this would lead to a velocity profile where the velocity would be higher near the wall than at the centre of the column.

The second parameter which have to be specified in the hydrodynamic model is the loss coefficient (Section 3.1.2.2). This naturally varies with the geometry of the airlift contactor and with the type of fluid employed. The loss coefficient (K_f) was used as an adjustable parameter in this study. For a particular system, the K_f value giving the best agreement between model predictions and experimental data was chosen. Tables 5.3 to 5.5 give the K_f values used in this study with the twenty different liquids in the three airlift
Results and Discussion

contactors. Generally, the loss coefficient is increasing with increasing liquid viscosity. This was due to higher energy losses occurring in the airlift vessel with more viscous liquids, hence a higher value of $K_f$ was needed to give a good agreement between experimental data and model predictions.

The loss coefficient values used in the three airlift reactors with the 16 liquids (the four oil-in-water emulsions are excluded) are plotted against the liquid phase viscosity in Figure 5.95. For a given CMC solution, two apparent viscosity values were used (i.e. the lowest and the highest values). It is apparent from Figure 5.95 that, usually, a low viscosity liquid gave a lower value of loss coefficient than a high viscosity liquid. For a specified

<table>
<thead>
<tr>
<th>Liquid</th>
<th>Distribution parameter (-)</th>
<th>Loss coefficient (-)</th>
<th>Terminal bubble rise velocity (m/s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Water</td>
<td>0.7</td>
<td>0.1</td>
<td>0.250</td>
</tr>
<tr>
<td>0.1 M NaCl</td>
<td>0.7</td>
<td>0.1</td>
<td>0.257</td>
</tr>
<tr>
<td>0.2 M NaCl</td>
<td>0.7</td>
<td>0.1</td>
<td>0.256</td>
</tr>
<tr>
<td>0.5 M NaCl</td>
<td>0.7</td>
<td>0.1</td>
<td>0.259</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Glycerol Concentration (%)</th>
<th>Distribution parameter (-)</th>
<th>Loss coefficient (-)</th>
<th>Terminal bubble rise velocity (m/s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>48.5</td>
<td>0.7</td>
<td>0.1</td>
<td>0.242</td>
</tr>
<tr>
<td>57.5</td>
<td>0.7</td>
<td>0.1</td>
<td>0.240</td>
</tr>
<tr>
<td>71.5</td>
<td>0.7</td>
<td>0.1</td>
<td>0.238</td>
</tr>
<tr>
<td>78.5</td>
<td>0.7</td>
<td>0.1</td>
<td>0.235</td>
</tr>
<tr>
<td>82.0</td>
<td>0.7</td>
<td>0.1</td>
<td>0.233</td>
</tr>
<tr>
<td>86.0</td>
<td>0.7</td>
<td>0.1</td>
<td>0.228</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>CMC Concentration (%)</th>
<th>Distribution parameter (-)</th>
<th>Loss coefficient (-)</th>
<th>Terminal bubble rise velocity (m/s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.05</td>
<td>0.7</td>
<td>0.1</td>
<td>0.250</td>
</tr>
<tr>
<td>0.2</td>
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<td>0.1</td>
<td>0.248</td>
</tr>
<tr>
<td>0.3</td>
<td>0.7</td>
<td>1.0</td>
<td>0.243</td>
</tr>
<tr>
<td>0.4</td>
<td>0.7</td>
<td>2.0</td>
<td>0.239</td>
</tr>
<tr>
<td>0.6</td>
<td>0.7</td>
<td>10.0</td>
<td>0.228</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Oil Concentration in water (%)</th>
<th>Distribution parameter (-)</th>
<th>Loss coefficient (-)</th>
<th>Terminal bubble rise velocity (m/s)</th>
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<tbody>
<tr>
<td>10</td>
<td>0.7</td>
<td>0.1</td>
<td>0.250</td>
</tr>
<tr>
<td>20</td>
<td>0.7</td>
<td>0.1</td>
<td>0.250</td>
</tr>
<tr>
<td>40</td>
<td>0.7</td>
<td>0.1</td>
<td>0.250</td>
</tr>
<tr>
<td>60</td>
<td>0.7</td>
<td>0.1</td>
<td>0.250</td>
</tr>
<tr>
<td>100 (no water)</td>
<td>0.7</td>
<td>2.0</td>
<td>0.250</td>
</tr>
</tbody>
</table>

Table 5.3: Model parameters for the internal loop airlift (1) device.
Results and Discussion

liquid viscosity, generally the internal loop airlift (1) geometry gave the lowest loss coefficient value.

a) Water and electrolytes

<table>
<thead>
<tr>
<th>Liquid</th>
<th>Distribution parameter (-)</th>
<th>Loss coefficient (-)</th>
<th>Terminal bubble rise velocity (m/s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Water</td>
<td>0.7</td>
<td>0.1</td>
<td>0.250</td>
</tr>
<tr>
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<td>0.7</td>
<td>0.1</td>
<td>0.257</td>
</tr>
<tr>
<td>0.2 M NaCl</td>
<td>0.7</td>
<td>0.1</td>
<td>0.256</td>
</tr>
<tr>
<td>0.5 M NaCl</td>
<td>0.7</td>
<td>0.1</td>
<td>0.259</td>
</tr>
</tbody>
</table>

b) Glycerol solutions

<table>
<thead>
<tr>
<th>Glycerol Concentration (%)</th>
<th>Distribution parameter (-)</th>
<th>Loss coefficient (-)</th>
<th>Terminal bubble rise velocity (m/s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>48.5</td>
<td>0.7</td>
<td>0.1</td>
<td>0.242</td>
</tr>
<tr>
<td>57.5</td>
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<td>0.1</td>
<td>0.240</td>
</tr>
<tr>
<td>71.5</td>
<td>0.7</td>
<td>5.0</td>
<td>0.238</td>
</tr>
<tr>
<td>78.5</td>
<td>0.7</td>
<td>9.0</td>
<td>0.235</td>
</tr>
<tr>
<td>82.0</td>
<td>0.7</td>
<td>26.0</td>
<td>0.233</td>
</tr>
<tr>
<td>86.0</td>
<td>0.7</td>
<td>64.0</td>
<td>0.228</td>
</tr>
</tbody>
</table>

c) CMC solutions

<table>
<thead>
<tr>
<th>CMC Concentration (%)</th>
<th>Distribution parameter (-)</th>
<th>Loss coefficient (-)</th>
<th>Terminal bubble rise velocity (m/s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.05</td>
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<td>1.0</td>
<td>0.250</td>
</tr>
<tr>
<td>0.2</td>
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<td>6.0</td>
<td>0.248</td>
</tr>
<tr>
<td>0.3</td>
<td>0.7</td>
<td>24.0</td>
<td>0.243</td>
</tr>
<tr>
<td>0.4</td>
<td>0.7</td>
<td>92.0</td>
<td>0.239</td>
</tr>
<tr>
<td>0.6</td>
<td>0.7</td>
<td>100.0</td>
<td>0.228</td>
</tr>
</tbody>
</table>

d) Oil-in-water emulsions

<table>
<thead>
<tr>
<th>Oil Concentration in water (%)</th>
<th>Distribution parameter (-)</th>
<th>Loss coefficient (-)</th>
<th>Terminal bubble rise velocity (m/s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>10</td>
<td>0.7</td>
<td>0.1</td>
<td>0.250</td>
</tr>
<tr>
<td>20</td>
<td>0.7</td>
<td>0.1</td>
<td>0.250</td>
</tr>
<tr>
<td>40</td>
<td>0.7</td>
<td>0.1</td>
<td>0.250</td>
</tr>
<tr>
<td>60</td>
<td>0.7</td>
<td>8.0</td>
<td>0.250</td>
</tr>
<tr>
<td>100 (no water)</td>
<td>0.7</td>
<td>80.0</td>
<td>0.250</td>
</tr>
</tbody>
</table>

Table 5.4: Model parameters for the internal loop airlift (2) contactor.

The final parameter which has to be specified in the model is the terminal bubble rise velocity (Section 3.1.2.3). The evaluation of the terminal bubble rise velocity with the different liquids was undertaken in this study. The riser section of the external loop airlift contactor was used for the measurement. The time taken for a single bubble (approximately 5 - 10 mm diameter) to travel a distance of 1.50 m in the chosen liquid was recorded. The distance was then divided by an average time (from 20 different readings) to obtain the rise
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velocity of a single bubble in isolation. Table 5.6 gives the bubble rise velocity with 15 different liquids. Due to the opaque nature of the oil-in-water emulsions, the above experiment could not be performed with the oil/water system.

It is apparent from Table 5.6 that, with water and the salt solutions, the terminal rise velocity of a bubble was within the range of 25.0 to 25.9 cm/s. With glycerol and CMC solutions, the range of the bubble rise velocity was from 22.8 to 25.0 cm/s and as expected the bubble rise velocity slightly decreased with increasing liquid viscosity.

a) Water and electrolytes

<table>
<thead>
<tr>
<th>Liquid</th>
<th>Distribution parameter (-)</th>
<th>Loss coefficient (-)</th>
<th>Terminal bubble rise velocity (m/s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Water</td>
<td>0.9</td>
<td>0.1</td>
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</tr>
<tr>
<td>0.1 M NaCl</td>
<td>0.9</td>
<td>0.1</td>
<td>0.257</td>
</tr>
<tr>
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<td>0.9</td>
<td>0.1</td>
<td>0.256</td>
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<tr>
<td>0.5 M NaCl</td>
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b) Glycerol solutions

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<tbody>
<tr>
<td>48.5</td>
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<td>2.0</td>
<td>0.242</td>
</tr>
<tr>
<td>57.5</td>
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<td>5.0</td>
<td>0.240</td>
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<tr>
<td>71.5</td>
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<td>10.0</td>
<td>0.238</td>
</tr>
<tr>
<td>78.5</td>
<td>0.9</td>
<td>30.0</td>
<td>0.235</td>
</tr>
<tr>
<td>82.0</td>
<td>0.9</td>
<td>45.0</td>
<td>0.233</td>
</tr>
<tr>
<td>86.0</td>
<td>0.9</td>
<td>70.0</td>
<td>0.228</td>
</tr>
</tbody>
</table>

c) CMC solutions

<table>
<thead>
<tr>
<th>CMC Concentration (%)</th>
<th>Distribution parameter (-)</th>
<th>Loss coefficient (-)</th>
<th>Terminal bubble rise velocity (m/s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.05</td>
<td>0.9</td>
<td>12.0</td>
<td>0.250</td>
</tr>
<tr>
<td>0.2</td>
<td>0.9</td>
<td>16.0</td>
<td>0.248</td>
</tr>
<tr>
<td>0.3</td>
<td>0.9</td>
<td>21.0</td>
<td>0.243</td>
</tr>
<tr>
<td>0.4</td>
<td>0.9</td>
<td>60.0</td>
<td>0.239</td>
</tr>
<tr>
<td>0.6</td>
<td>0.9</td>
<td>75.0</td>
<td>0.228</td>
</tr>
</tbody>
</table>

d) Oil-in-water emulsions

<table>
<thead>
<tr>
<th>Oil Concentration in water (%)</th>
<th>Distribution parameter (-)</th>
<th>Loss coefficient (-)</th>
<th>Terminal bubble rise velocity (m/s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>10</td>
<td>0.9</td>
<td>1.0</td>
<td>0.250</td>
</tr>
<tr>
<td>20</td>
<td>0.9</td>
<td>1.0</td>
<td>0.250</td>
</tr>
<tr>
<td>40</td>
<td>0.9</td>
<td>1.0</td>
<td>0.250</td>
</tr>
<tr>
<td>60</td>
<td>0.9</td>
<td>1.0</td>
<td>0.250</td>
</tr>
<tr>
<td>100 (no water)</td>
<td>0.9</td>
<td>65.0</td>
<td>0.250</td>
</tr>
</tbody>
</table>

Table 5.5: Model parameters for the external loop airlift reactor.
Results and Discussion

The values of the terminal bubble rise velocities given in Table 5.6 was used in the hydrodynamic model with the first 15 liquids. However, for the last 5 liquids (oil-in-water emulsions and pure oil) a value of 0.25 m/s was taken to be the rise velocity of a single bubble in isolation.

![Graph](image)

Figure 5.95: The loss coefficient versus the liquid phase viscosity in the three airlift contactors.

Figures 5.96 to 5.101 show the experimental riser gas holdup and superficial liquid velocity data against the model predicted values for water and the three salt solutions. The thick solid lines in the figures indicate exact prediction, while the two lines on either side of the thick line denote the ± 20% values.

Generally, the agreement between the experimental and predicted gas holdup and liquid velocity values were good in the three geometries of airlift reactors. However, the riser liquid velocity was underpredicted by the model in the internal loop airlift (1) and (2) reactors (Figs 5.97 and 5.99). A $K_f$ value of 0.1 was employed in the model. A low $K_f$ value in the model gives a higher superficial liquid velocity (Fig 3.4). Unusually high liquid velocities occurred in the internal loop airlift contactors with low viscosity liquids.
Results and Discussion

(eespecially in the internal loop airlift (1) vessel). Hence, although the lowest possible value for $K_f$ was employed, the model still underpredicted the superficial liquid velocity by more than 20%. In the external loop airlift contactor, with water and salt solutions, a $K_f$ value of 0.1 was used. This gave relatively good predictions of the superficial liquid velocity. But, with the salt solutions, the riser gas holdup was slightly underpredicted by the model especially at higher superficial gas velocities.

The energy losses in the internal loop airlift devices used in the present study were expected to be very low. Since the bottom section of the airlift vessel was designed to minimise the energy losses occurring at the bottom. Chisti et al. (1988) also reported that the energy loss at the top of an internal loop airlift device was negligible compared to the bottom section. Therefore, in the internal loop airlift vessels with water and salt solutions energy losses due to fluid turn around were extremely small. A low $K_f$ value in the external loop airlift vessel, with water and salt solutions, indicated that the energy losses due to fluid turn around were relatively low.

a) Water and electrolytes

<table>
<thead>
<tr>
<th>Liquid</th>
<th>Viscosity (Pa s)</th>
<th>Terminal bubble rise velocity (m/s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Water</td>
<td>0.0009</td>
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</tr>
<tr>
<td>0.1 M NaCl</td>
<td>0.0010</td>
<td>0.257</td>
</tr>
<tr>
<td>0.2 M NaCl</td>
<td>0.0011</td>
<td>0.256</td>
</tr>
<tr>
<td>0.5 M NaCl</td>
<td>0.0012</td>
<td>0.259</td>
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b) Glycerol solutions

<table>
<thead>
<tr>
<th>Glycerol Concentration (%)</th>
<th>Viscosity (Pa s)</th>
<th>Terminal bubble rise velocity (m/s)</th>
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<tr>
<td>48.5</td>
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<td>0.242</td>
</tr>
<tr>
<td>57.5</td>
<td>0.0125</td>
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<td>71.5</td>
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<tr>
<td>82.0</td>
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<td>0.233</td>
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<tr>
<td>86.0</td>
<td>0.1444</td>
<td>0.228</td>
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</tbody>
</table>

c) CMC solutions

<table>
<thead>
<tr>
<th>CMC Concentration (%)</th>
<th>Flow behaviour index (-)</th>
<th>Terminal bubble rise velocity (m/s)</th>
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<td>0.2</td>
<td>0.80</td>
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<td>0.4</td>
<td>0.64</td>
<td>0.239</td>
</tr>
<tr>
<td>0.6</td>
<td>0.57</td>
<td>0.228</td>
</tr>
</tbody>
</table>

Table 5.6: The rise velocity of a bubble ($U_{br}$) with different fluids in the external loop airlift reactor.
There could be several reasons why the model significantly underpredicted (by more than 20%) the superficial liquid velocity with low viscosity liquids in the internal loop airlift contactors (Figures 5.97, 5.99, 5.103, 5.105, 5.109, 5.111, 5.115 and 5.117):

1) Unusually high liquid velocities in the internal loop airlift contactors occurred with low viscosity liquids primarily due to negligible energy losses. The model is unable to predict these liquid velocities even with a $K_f$ value of 0.1, indicating that the ‘upper limit’ (for $U_{sl}$) of the model has been reached with regard to the specified conditions.

2) With low viscosity liquids in the internal loop airlift contactor, the flow regime in the device might be in the fully developed churn turbulent flow. While the model was only developed for bubbly, coalesced bubble and ‘lower’ churn turbulent flow regimes.

3) The hydrodynamic model assumes that the gas holdup in the downcomer is negligible. However, experimentally, there was a significant amount of gas recirculation into the downcomer.
Figure 5.96: Riser gas holdup in the internal loop airlift (1). Experimental versus predicted gas holdup. The thick solid line indicates exact prediction.

Figure 5.97: Riser liquid circulation velocity in the internal loop airlift (1). Experimental versus predicted liquid velocity. The thick solid line indicates exact prediction.
Results and Discussion

Figure 5.98: Riser gas holdup in the internal loop airlift (2). Experimental versus predicted gas holdup. The thick solid line indicates exact prediction.

Figure 5.99: Riser liquid circulation velocity in the internal loop airlift (2). Experimental versus predicted liquid velocity. The thick solid line indicates exact prediction.
Results and Discussion

Figure 5.100: Riser gas holdup in the external loop airlift. Experimental versus predicted gas holdup. The thick solid line indicates exact prediction.

Figure 5.101: Riser liquid circulation velocity in the external loop airlift. Experimental versus predicted liquid velocity. The thick solid line indicates exact prediction.
Results and Discussion

The experimental riser gas holdup and superficial liquid velocity values against the predicted ones for the six glycerol solutions are displayed in Figures 5.102 to 5.107. In general, the agreement between the experimental and predicted gas holdup and liquid velocity values were reasonable in the three geometries of airlift contactors. But, the riser liquid velocity was again underpredicted by the model in the internal loop airlift (1) and (2) columns (Figs 5.103 and 5.105) with low viscosity glycerol solutions (especially 48.5% and 57.5% glycerol solutions). Even though the lowest possible value for $K_f$ (0.1) was used in the model. With low viscosity liquids in the internal loop airlift reactors, abnormally high liquid velocities occurred due to very low frictional losses in the vessels. Hence, the hydrodynamic model is unable to closely predict the high liquid velocities occurring in the internal loop airlift reactors. In the internal loop airlift (1) vessel with all six glycerol solutions the same $K_f$ value (0.1) was employed. This indicated that in the internal loop airlift (1) device, even with viscous glycerol solutions the energy losses remained relatively low. However, in the internal loop airlift (2) and the external loop airlift reactors, the loss coefficient increased with an increase in the glycerol concentration. Better liquid velocity predictions were obtained in the internal loop airlift (2) and the external loop airlift contactors than in the internal loop airlift (1) vessel with glycerol solutions. It is apparent that for a given glycerol solution, a higher value for $K_f$ was used for the external loop airlift contactor in the hydrodynamic model than for the two internal loop airlift vessels. This indicated that, for a given glycerol solution, higher energy losses occurred in the external loop airlift vessel than in the internal loop airlift contactors.

The gas holdup was predicted reasonably well in both the internal loop airlift contactors (Figs 5.102 and 5.104). With 48.5% glycerol solution in the internal loop airlift (1) contactor the gas holdup was slightly overpredicted at low gas velocities. While, in the internal loop airlift (2) vessel with 48.5% glycerol solution the riser gas holdup was slightly underpredicted. In the external loop airlift vessel with 48.5% and 57.5% glycerol solutions the gas holdup was underpredicted by more than 20% (especially at high gas velocities). This might be due to the model assuming that there is an insignificant amount of holdup in the downcomer. But, in practice more than 40% of the riser gas was recirculating into the downcomer. The gas in the downcomer then recirculates into the riser section thereby giving a higher experimental riser holdup. With a more viscous glycerol solution (86.0%) the gas holdup was overpredicted by more than 20% at low gas velocities (Fig 5.106).
Results and Discussion

Figure 5.102: Riser gas holdup in the internal loop airlift (1). Experimental versus predicted gas holdup. The thick solid line indicates exact prediction.

Figure 5.103: Riser liquid circulation velocity in the internal loop airlift (1). Experimental versus predicted liquid velocity. The thick solid line indicates exact prediction.
Results and Discussion

Figure 5.104: Riser gas holdup in the internal loop airlift (2). Experimental versus predicted gas holdup. The thick solid line indicates exact prediction.

Figure 5.105: Riser liquid circulation velocity in the internal loop airlift (2). Experimental versus predicted liquid velocity. The thick solid line indicates exact prediction.
Results and Discussion

Figure 5.106: Riser gas holdup in the external loop airlift. Experimental versus predicted gas holdup. The thick solid line indicates exact prediction.

Figure 5.107: Riser liquid circulation velocity in the external loop airlift. Experimental versus predicted liquid velocity. The thick solid line indicates exact prediction.
The experimental riser gas holdup and superficial liquid velocity results against the model generated values for the five CMC solutions are given in Figures 5.108 to 5.113. The agreement between the experimental and predicted gas holdup and liquid velocity values are usually within ± 20%. In the internal loop airlift (1) vessel, with 0.2% and 0.3% CMC solutions, the hydrodynamic model overpredicted the riser gas holdup by more than 20% (Fig 5.108) at low gas velocities. However, the gas holdup was predicted to within ± 20% in the internal loop airlift (2) vessel and the external loop airlift contactor with the five CMC solutions.

The riser liquid velocity was underpredicted by the model in the internal loop airlifts (1) and (2), Figures 5.109 and 5.111, with low apparent viscosity CMC solutions (0.05% and 0.2% CMC solutions). This was similar to the results obtained in the internal loop airlifts with water, salt solutions and low concentrations of glycerol solutions. With 0.05% and 0.2% CMC solutions, high liquid velocities occurred in the internal loop airlifts (1) and (2) due to low energy losses in the vessels. Hence, the model predictions of liquid velocity was lower than the experimental values. This problem did not occur in the external loop airlift contactor, where the model gave relatively good predictions of the superficial liquid velocity with the five CMC solutions (Fig 5.113). In all three airlift vessels, with increasing concentration of the CMC solution the loss coefficient is also increasing. This meant that the energy losses in the column were increasing with increasing apparent viscosity of the liquid medium. In the internal loop airlift (1) vessel this was opposite to what was observed with the glycerol solutions, where the loss coefficient did not change with an increase in the Newtonian viscosity. For a given CMC solution, the internal loop airlift (1) geometry produced the lowest energy loss. At low CMC concentrations (0.05% and 0.2%), higher energy losses occurred in the external loop airlift vessel than in the internal loop airlift (2) contactor, while at higher CMC concentrations the opposite was true. This indicated that the type of liquid in the column has a profound influence on the energy losses occurring in an airlift column and therefore on the hydrodynamics and liquid mixing in the airlift vessel. In the internal loop airlift, for a CMC solution, the loss coefficient was significantly affected by changing the diameter of the draft tube [i.e. the difference between the internal loop airlift (1) and the internal loop airlift (2)].

Generally, with CMC solutions, the hydrodynamic model gave better predictions of the riser gas holdup than the riser superficial liquid velocity.
Results and Discussion

Figure 5.108: Riser gas holdup in the internal loop airlift (1). Experimental versus predicted gas holdup. The thick solid line indicates exact prediction.

Figure 5.109: Riser liquid circulation velocity in the internal loop airlift (1). Experimental versus predicted liquid velocity. The thick solid line indicates exact prediction.
Figure 5.110: Riser gas holdup in the internal loop airlift (2). Experimental versus predicted gas holdup. The thick solid line indicates exact prediction.

Figure 5.111: Riser liquid circulation velocity in the internal loop airlift (2). Experimental versus predicted liquid velocity. The thick solid line indicates exact prediction.
Results and Discussion

Figure 5.112: Riser gas holdup in the external loop airlift. Experimental versus predicted gas holdup. The thick solid line indicates exact prediction.

Figure 5.113: Riser liquid circulation velocity in the external loop airlift. Experimental versus predicted liquid velocity. The thick solid line indicates exact prediction.
Results and Discussion

The experimental gas holdup and superficial liquid velocity in the riser section versus that predicted using the hydrodynamic model for the four oil-in-water emulsions and pure oil are depicted in Figures 5.114 to 5.119. In the case of the pure oil, the agreement between the experimental and predicted gas holdup and liquid velocity values are generally within ± 20%. However, with the four oil-in-water emulsions the agreement between the predicted and experimental values is not so good. This might be due to the fact that the hydrodynamic model was developed for gas-liquid systems only and not for gas-liquid-liquid systems. The fluid dynamics in the latter is more complex than the former system and this could be why the hydrodynamic model cannot give reasonable predictions.

Usually, in all three airlift contactors with pure oil (two-phase system) relatively good predictions of the riser gas holdup was obtained (Figs 5.114, 5.116 and 5.118). In the two internal loop airlift vessels, the liquid velocity was predicted to within ± 20% with pure oil. However, with pure oil, in the external loop airlift reactor the model slightly overpredicted the superficial liquid velocity. With pure oil, in the internal loop airlift (1) vessel a loss coefficient value of 2.0 was used, while the loss coefficient in the internal loop airlift (2) reactor was forty times the value used in the internal loop airlift (1) vessel. This indicated that the loss coefficient was affected by changing the draft tube in the internal loop airlift vessel.

With all the four oil-in-water emulsions, the riser gas holdup was overpredicted by more than 20% in the internal loop airlift (1) contactor, especially at higher gas velocities. But, the liquid velocity was underpredicted by the model although a very low value for the loss coefficient was used. High liquid velocities occurred in the internal loop airlift (1) vessel due to low energy losses and the model could not closely predict the high liquid velocity. In the internal loop airlift (2) reactor, with 60% oil-in-water emulsion, the gas holdup was overpredicted by the model. With 10 to 40% oil-in-water emulsions, the superficial liquid velocity was underpredicted by the model (Fig 5.117) even though a loss coefficient value of 0.1 was employed. In the internal loop airlift (2) contactor, especially with low viscosity liquids, relatively high liquid velocities occurred and the model underpredicted these velocities. With 10% oil-in-water emulsion, the gas holdup was underpredicted by more than 20% in the external loop airlift vessel, while with a higher concentration of oil-in-water emulsion (60%) the gas holdup was overpredicted by the model. The superficial liquid velocity was underpredicted by more than 20% in the external loop airlift contactor with 10% and 20% oil-in-water emulsions.
Results and Discussion

Figure 5.114: Riser gas holdup in the internal loop airlift (1). Experimental versus predicted gas holdup. The thick solid line indicates exact prediction.

Figure 5.115: Riser liquid circulation velocity in the internal loop airlift (1). Experimental versus predicted liquid velocity. The thick solid line indicates exact prediction.
Results and Discussion

Figure 5.116: Riser gas holdup in the internal loop airlift (2). Experimental versus predicted gas holdup. The thick solid line indicates exact prediction.

Figure 5.117: Riser liquid circulation velocity in the internal loop airlift (2). Experimental versus predicted liquid velocity. The thick solid line indicates exact prediction.
Results and Discussion

Figure 5.118: Riser gas holdup in the external loop airlift. Experimental versus predicted gas holdup. The thick solid line indicates exact prediction.

Figure 5.119: Riser liquid circulation velocity in the external loop airlift. Experimental versus predicted liquid velocity. The thick solid line indicates exact prediction.
The experimental data of riser gas holdup and superficial liquid velocity obtained by earlier studies were also compared with the predictions from the hydrodynamic model. The details of the airlift reactors used by the previous researchers together with the fluid properties are given in Table 5.7. Experimental data obtained by previous investigators in a wide range of airlift reactor geometry in both external and internal (draft tube and split cylinder) loop airlift contactors were chosen. The range of the volume of the contactors was from 0.033 to 1.058 m³, while the height range was from 1.220 to 5.076 m. The downcomer to riser cross sectional area ratio was between 0.11 and 1.45. Both Newtonian and non-Newtonian liquids were used. The range of the flow behaviour index was from 0.240 to 1, whereas the consistency index range was from 0.00089 to 6.800 Pa s⁻¹.

The graphs of experimental riser gas holdup and liquid velocity values against the predicted ones are shown in Figures 5.120 to 5.126. The thick solid line shows exact agreement between experimental and predicted values. The two solid lines on either side of the thick solid line indicate the ± 30% values. The key in these graphs is given in Table 5.7. The values of the loss coefficients used in the model for the different systems are also given in Table 5.7. For external loop airlift reactors (Table 5.7), 0.9 was taken to be the value of the distribution parameter. And the distribution parameter in the internal loop airlift vessels was taken to be 0.7. The terminal bubble rise velocity was taken to be 0.25 m/s in all the model predictions.

The experimental riser gas holdup values of Kemblowski et al. (1993) with water, glycol and CMC solutions were lower than the predicted values by more than 30% (Fig 5.120). While Kemblowski et al.'s riser superficial liquid velocity values with water and glycol solution were underpredicted by the hydrodynamic model (Fig 5.123) even though a very low value (0.1) for the loss coefficient was used. It seems that with low viscosity liquids, the hydrodynamic model is unable to closely predict the relatively high liquid velocities occurring in the airlift vessel. Similarly, the hydrodynamic model underpredicted Glennon et al.'s (1993) experimental liquid velocity values obtained with water (Fig 5.123). Especially with a higher A_d/A_r ratio external loop airlift reactor. A higher loss coefficient was employed with the lower A_d/A_r ratio external loop airlift than with the higher A_d/A_r ratio. [This was in agreement with findings of this study, where usually for a given liquid a higher value for the loss coefficient was obtained with the internal loop airlift (2) vessel than with the internal loop airlift (1) configuration.] With the external loop airlift configuration where the valve was partially closed, a much higher value (100.0) of
loss coefficient was used. This was in agreement with Glennon et al.’s observations. For this configuration of airlift vessel, they calculated a much higher value for the effective resistance of the reactor compared to the other two geometries. The model also underpredicted Glennon et al.’s experimental gas holdup values (Fig 5.120). Relatively good agreement between Choi and Lee’s (1993) experimental gas holdup values and model predictions were obtained (Fig 5.120). However, Choi and Lee’s superficial liquid velocity values were overpredicted by the model (Fig 5.123). This was due to the high value of loss coefficient employed in the model. With Chisti’s (1989) experimental gas holdup and superficial liquid velocity data obtained in internal (draft tube and split cylinder) and external loop airlift vessels, the hydrodynamic model gave relatively good predictions (Figs 5.121 and 5.124). But, Young et al.’s (1991) gas holdup and liquid velocity experimental data obtained in an external loop airlift vessel with water were underpredicted by more than 30% (Figs 5.122 and 5.125). A lower loss coefficient was used with the higher $A_d/A_r$ ratio (0.543) external loop airlift configuration than with the lower $A_d/A_r$ ratio (0.219) geometry. In the higher $A_d/A_r$ ratio airlift vessel, the model was unable to closely predict the liquid velocity values even though a low value (0.1) for the loss coefficient was employed. Garcia Calvo and Leton (1991) measured the riser gas holdup in an external loop airlift vessel with air/water system. It is apparent from Figure 5.122 that the hydrodynamic model gave good predictions of the gas holdup. Gas holdup and liquid circulation velocity were measured by Choi et al. (1995) in an internal loop airlift contactor with water. The agreement between the measured values and the predicted ones were relatively good. The unusual design of the bottom section of the airlift vessel resulted in relatively high energy losses (the base of the downcomer was inclined at an angle of 60° to the horizontal). Kawase and Moo-Young (1986a) obtained gas holdup measurements in an internal loop airlift vessel with water ($K=0.00089 \text{ Pa s}^n$) and glycerine solution ($K=0.00150 \text{ Pa s}^n$). The hydrodynamic model was able to predict their holdup values to within +30% as shown in Fig 5.122. A relatively high value for the loss coefficient was used. This was due to the vessel having an inverted conical bottom, which resulted in higher energy losses. The experimental superficial liquid velocity values of Hatch (1973) were predicted to within ±30% (Fig 5.124). Jones (1985) also measured the liquid velocity in an internal loop airlift vessel with water. There was good agreement between his experimental data and the model predicted values (Fig 5.124). The bottom section of the draft tube internal loop airlift vessel employed by Jones (1985) was similar to the bottom section of the internal loop airlift contactor used in this work.
Results and Discussion

Hence, it was not surprising that a loss coefficient value of 0.1 gave the best agreement between Jones's experimental data and predicted values. Bello's (1981) experimental liquid velocity data were overpredicted by the model as shown in Figure 5.124. Vatai and Tekic (1986) measured the liquid velocity in an external loop airlift vessel with CMC solutions. With $A_d/A_r = 0.5$ and 1.0, the model underpredicted the liquid velocity in the column by more than 30% although a low value for the loss coefficient was used. Popovic and Robinson (1988) also measured the superficial liquid velocity in an external loop airlift reactor with viscous CMC solutions. It is apparent from Figure 5.126 that the model predicted values were higher than the experimental values even though a relatively high value (100.0) for the loss coefficient was used. It seems that the low $A_d/A_r$ ratios of Popovic and Robinson's vessel together with the relatively viscous liquids produced exceptionally low liquid velocities.

Figure 5.120: Riser gas holdup in airlift reactors. Experimental versus predicted gas holdup. The thick solid line indicates exact prediction.
Results and Discussion

Figure 5.121: Riser gas holdup in airlift reactors. Experimental versus predicted gas holdup. The thick solid line indicates exact prediction.

Figure 5.122: Riser gas holdup in airlift reactors. Experimental versus predicted gas holdup. The thick solid line indicates exact prediction.
Results and Discussion

Figure 5.123: Riser superficial liquid velocity in airlift reactors. Experimental versus predicted liquid velocity. The thick solid line indicates exact prediction.

Figure 5.124: Riser superficial liquid velocity in airlift reactors. Experimental versus predicted liquid velocity. The thick solid line indicates exact prediction.
Figure 5.125: Riser superficial liquid velocity in airlift reactors. Experimental versus predicted liquid velocity. The thick solid line indicates exact prediction.

Figure 5.126: Riser superficial liquid velocity in airlift reactors. Experimental versus predicted liquid velocity. The thick solid line indicates exact prediction.
<table>
<thead>
<tr>
<th>Reference</th>
<th>Airlift reactor</th>
<th>Reactor geometry</th>
<th>Liquid property</th>
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</thead>
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<td></td>
<td>External</td>
<td>Internal</td>
<td>$K_{f}$</td>
</tr>
<tr>
<td></td>
<td>Draft tube</td>
<td>Split cylinder</td>
<td>(m$^3$)</td>
</tr>
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</tr>
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<tr>
<td>(c)</td>
<td>x</td>
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</tr>
<tr>
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<td>0.054</td>
</tr>
<tr>
<td>(d)</td>
<td>x</td>
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<tr>
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<td>(h)</td>
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<td>(j)</td>
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<td>100.0</td>
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<td>Reactor geometry</td>
<td>Liquid property</td>
</tr>
<tr>
<td>-----------</td>
<td>----------------</td>
<td>-----------------</td>
<td>----------------</td>
</tr>
<tr>
<td></td>
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<td>Internal</td>
<td>$K_f$</td>
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<tr>
<td>Draft tube</td>
<td>Split cylinder</td>
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<td>(a)</td>
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<td>(b)</td>
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<tr>
<td>(6) Garcia Calvo and Leton (1991)</td>
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<td>(7) Choi et al. (1995)</td>
<td></td>
<td>x</td>
<td>100.0</td>
</tr>
<tr>
<td>(8) Kawase and Moo-Young (1986a)</td>
<td>(a)</td>
<td>x</td>
<td>100.0</td>
</tr>
<tr>
<td>(b)</td>
<td>x</td>
<td>100.0</td>
<td>0.040</td>
</tr>
<tr>
<td>(9) Hatch (1973)</td>
<td></td>
<td>x</td>
<td>100.0</td>
</tr>
<tr>
<td>(10) Jones (1985)</td>
<td></td>
<td>x</td>
<td>0.1</td>
</tr>
<tr>
<td>(11) Bello (1981)</td>
<td>(a)</td>
<td>x</td>
<td>100.0</td>
</tr>
<tr>
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<td>0.033</td>
</tr>
<tr>
<td>(c)</td>
<td>x</td>
<td>0.1</td>
<td>0.033</td>
</tr>
<tr>
<td>(12) Vatai and Tekic (1986)</td>
<td>$A_n/A_d = 0.3$</td>
<td>(a)</td>
<td>x</td>
</tr>
<tr>
<td>(b)</td>
<td>x</td>
<td>0.1</td>
<td>-</td>
</tr>
<tr>
<td>(c)</td>
<td>x</td>
<td>0.1</td>
<td>-</td>
</tr>
<tr>
<td>(13) Popovic and Robinson (1988)</td>
<td>(a)</td>
<td>x</td>
<td>100.0</td>
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<tr>
<td>(b)</td>
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</tr>
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<td>(d)</td>
<td>x</td>
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</tr>
<tr>
<td>(e)</td>
<td>x</td>
<td>100.0</td>
<td>-</td>
</tr>
</tbody>
</table>

Table 5.7: Previous researchers’ airlift reactor geometries and the properties of fluids employed in their airlift vessels.
5.5.2 Liquid Circulation Time

The liquid circulation time was measured with all twenty liquids in the three airlift vessels at various superficial gas velocities as described in Section 4.6. Equation 3.50 (Section 3.3.1) was used to calculate the theoretical circulation time in the airlift vessel. For a given superficial gas velocity, airlift geometry and fluid properties, the riser gas holdup and superficial liquid velocity values were obtained from the hydrodynamic model. These values, together with the loop length (L) of the reactor and \( A_r/A_d \) ratio was inserted into Equation 3.50 to obtain a value for the liquid circulation time.

Figures 5.127 to 5.138 show the measured liquid circulation time against the predicted value for all 20 liquids in the three airlift vessels. The thick solid line indicates exact agreement between predicted and measured values. The two solid lines on either side of the thick solid line give the ± 20% values. Generally, there was a reasonable agreement between the measured and predicted circulation time values. In the internal loop airlift (1) column, the liquid circulation time correlation is overpredicting (by more than 20%) the circulation time with water and salt solutions (Fig 5.127). This was due to the hydrodynamic model underpredicting the riser superficial liquid velocities with water and salt solutions in the internal loop airlift (1) column (Fig 5.97). A lower superficial liquid velocity in Eqn 3.50 gives a higher circulation time. Also, with 48.5% and 57.5% glycerol solutions the predicted circulation times were higher (by more than 20%) than the measured ones (Fig 5.130) due to lower predicted liquid velocities than measured values. The predicted circulation time was more than 20% lower than the measured time in all three airlift contactors with 82.0% and 86.0% glycerol solutions. This was probably due to the hydrodynamic model overpredicting (but within -20%) the superficial liquid velocity with 82.0% and 86.0% glycerol solutions. With CMC solutions in the two internal loop airlift vessels, relatively good predictions of circulations times were observed. However, in the external loop airlift contactor (Fig 5.135), the circulation time equation underpredicted the circulation time by more than 20%, especially with 0.4% and 0.6% CMC solutions.  

With 10 to 60% oil-in-water emulsions, in the internal loop airlift (1) contactor, the predicted values of circulation time were higher than the measured values by more than 20% (Fig 5.136). This was due to the hydrodynamic model underpredicting the superficial liquid velocity (Fig 5.115). In the internal loop airlift (2) vessel and the external loop airlift device with pure oil, the experimental values of the circulation time were considerably
Results and Discussion

higher than the predicted values. This was due to the hydrodynamic model slightly overpredicting the superficial liquid velocity values with pure oil (Figs 5.117 and 5.119).

Figure 5.127: Liquid circulation time in the internal loop airlift (1). Experimental versus predicted circulation time. The thick solid line indicates exact prediction.

Figure 5.128: Liquid circulation time in the internal loop airlift (2). Experimental versus predicted circulation time. The thick solid line indicates exact prediction.
Results and Discussion

Figure 5.129: Liquid circulation time in the external loop airlift. Experimental versus predicted circulation time. The thick solid line indicates exact prediction.

Figure 5.130: Liquid circulation time in the internal loop airlift (1). Experimental versus predicted circulation time. The thick solid line indicates exact prediction.
Results and Discussion

Figure 5.131: Liquid circulation time in the internal loop airlift (2). Experimental versus predicted circulation time. The thick solid line indicates exact prediction.

Figure 5.132: Liquid circulation time in the external loop airlift. Experimental versus predicted circulation time. The thick solid line indicates exact prediction.
Results and Discussion

Figure 5.133: Liquid circulation time in the internal loop airlift (1). Experimental versus predicted circulation time. The thick solid line indicates exact prediction.

Figure 5.134: Liquid circulation time in the internal loop airlift (2). Experimental versus predicted circulation time. The thick solid line indicates exact prediction.
Results and Discussion

Figure 5.135: Liquid circulation time in the external loop airlift. Experimental versus predicted circulation time. The thick solid line indicates exact prediction.

Figure 5.136: Liquid circulation time in the internal loop airlift (1). Experimental versus predicted circulation time. The thick solid line indicates exact prediction.
Results and Discussion

Figure 5.137: Liquid circulation time in the internal loop airlift (2). Experimental versus predicted circulation time. The thick solid line indicates exact prediction.

Figure 5.138: Liquid circulation time in the external loop airlift. Experimental versus predicted circulation time. The thick solid line indicates exact prediction.
5.6 Shear Rate in Airlift Reactors

In this work, the average shear rate ($\gamma$) in an airlift reactor with non-Newtonian media was calculated using Equation 3.46. For the five different concentrations of CMC solutions in the three airlift vessels, the calculated average shear rate is presented as a function of superficial gas velocity in Figs 5.139 to 5.141. It is apparent that the average shear rate is dependent on the superficial gas velocity and the type of non-Newtonian fluid in the reactor. The average shear rate in the airlift vessel increases with increasing gas velocity. This was due to increasing riser gas holdup and superficial liquid velocity with increasing gas velocity (Section 5.1). Since in Eqn 3.46, a higher gas holdup and liquid velocity will give a higher average shear rate than a lower gas holdup and liquid velocity.

![Figure 5.139: The average shear rate as a function of the superficial gas velocity in the internal loop airlift (1) contactor.](image)

Generally, for a given gas velocity, a low viscosity non-Newtonian fluid will produce a higher average shear rate than a higher viscosity liquid (Figs 5.139 - 5.141). Equation 3.46 includes the flow behaviour index of the liquid and one would expect this to contribute to the above effect. However, the increase in the average shear rate with
Results and Discussion

decreasing flow behaviour index was not because of the n term in Eqn 3.46. Since for a n value of 0.90, the term \((1+3n)/4n\) is 1.028 and for a flow behaviour index of 0.57 the term \((1+3n)/4n\) is 1.189. That is the expression \((1+3n)/4n\) is increasing with decreasing flow behaviour index and this would imply that the average shear rate is also increasing with decreasing n. With decreasing flow behaviour index of the liquid, the riser superficial liquid velocity is also decreasing (Section 5.3.2.2) but the riser gas holdup is increasing (Section 5.3.2.1). According to Equation 3.46, decreasing the liquid velocity (keeping other parameters constant) would decrease the average shear rate in the reactor. But increasing the gas holdup (keeping other parameters constant) would tend to have an opposite effect. From Figs 5.139 - 5.141, it is clear that the average shear rate is primarily influenced by the liquid velocity in the column. A higher liquid velocity gives a higher shear rate in the airlift vessel than a lower liquid velocity. This was anticipated, since a higher liquid velocity in the airlift would produce a higher turbulence intensity and hence a higher average shear rate in the vessel.

![Graph](image)

Figure 5.140: The average shear rate as a function of the superficial gas velocity in the internal loop airlift (2) contactor.
Results and Discussion

In Figure 5.142, the average shear rate in the external loop airlift contactor is plotted against the superficial gas velocity for 0.05% and 0.6% CMC solutions. The average shear rates obtained from the equation presented in this work is compared with the correlations from previous studies (Table 2.2). For 0.05% and 0.6% CMC solutions, our equation (3.46) gave different average shear rates (shown by two lines in Fig 5.142). The other researchers’ (Henzler, 1980; Kawase and Moo-Young, 1986a; Nishikawa et al., 1977; Schumpe and Deckwer, 1977; Shi et al., 1990) correlations gave only one line irrespective of whether the 0.05% or the 0.6% CMC solution was used. This was because their correlations considered the average shear rate in the vessel as just a function of the superficial gas velocity. It is interesting to note that the shear rate obtained in the external loop airlift vessel using our equation is lower than the shear rates acquired employing the correlations of Henzler (1980), Nishikawa et al. (1977) and Schumpe and Deckwer (1977). These latter correlations were obtained in bubble columns where the average shear rates were higher than in airlift reactors (Shi et al., 1990). Kawase and Moo-Young’s (1986a) correlation seems to give very low values of average shear rates (Fig 5.142). This

Figure 5.141: The average shear rate as a function of the superficial gas velocity in the external loop airlift contactor.
Results and Discussion

correlation was obtained in a 20 cm diameter bubble column and might not be applicable to an airlift reactor. The correlation of Shi et al. (1990) gives a curve for the average shear rate with a minimum occurring at a superficial gas velocity of between 0.01 and 0.02 m/s. This was not surprising since Shi et al.’s (1990) correlation was a quadratic equation (Table 2.2).

Figure 5.142: The effect of superficial gas velocity on the average shear rate in the external loop airlift contactor with 0.05% and 0.6% CMC solutions. The shear rate equation presented in this work is compared with the shear rate correlations from the published works.
6 CONCLUSIONS AND RECOMMENDATIONS

6.1 Conclusions
The following conclusions are based on the results of the experiments with the three geometries of airlift contactors and the modelling work as described in this thesis:

1) The hydrodynamics and liquid mixing in the airlift vessels improve with an increase in the gas velocity (within the ranges used in this study). As the gas flow rate increases, a change in the flow regime from a homogeneous bubbly flow to a more turbulent heterogeneous flow occurs and this reduces the dependency of the gas holdup, liquid velocity and mixing and circulation times on the superficial gas velocity.

2) For a given airlift reactor configuration, the average amount of riser gas recirculating into the downcomer section increases with an increase in the viscosity of the liquid phase. The riser gas recirculation rate also increases with an increase in the liquid circulation velocity.

3) Higher Bo numbers were observed in the riser section of the airlift device compared to the downcomer for a fixed gas flow rate. Conversely, the liquid phase axial dispersion coefficient in the downcomer section was higher than in the riser. The overall Bo number for the vessel was in-between the riser and downcomer Bo numbers for a given gas velocity. The same was true with the overall dispersion coefficient.

4) The fluid flow in the upflow region (riser) of the airlift contactor approaches plug flow behaviour, while well mixedness in the downflow region (downcomer). The fluid flow in the vessel as a whole verges upon plug flow behaviour.

5) The ratio of the mixing to circulation time in the airlift vessel was not affected by the gas flow rate indicating that the rate of mixing is mainly dependent on the frequency of passage of the fluid through the end sections of the reactor. The mixing to circulation time ratio was also not influenced by either the liquid phase electrolyte concentration or the liquid phase Newtonian and apparent viscosity or the two immiscible liquid systems.

6) The riser and downcomer gas holdups improve with an increase in the electrolyte concentration of the liquid phase (within the ranges studied) for a given gas velocity. However, the liquid velocity and mixing deteriorate with increasing salt concentration. Higher salt concentrations gave lower riser and downcomer liquid velocities and therefore
Conclusions and Recommendations

higher mixing and circulation times. The poor mixing with increasing salt concentration was also confirmed by the axial dispersion studies.

7) For a fixed gas velocity, the stable values of riser and downcomer gas holdups improve with an increase in the Newtonian and pseudoplastic non-Newtonian liquid phase viscosity (within the ranges studied). But, the riser and downcomer liquid velocities worsen with increasing Newtonian and apparent viscosities. Hence, the liquid mixing deteriorates with increasing viscosity of the liquid phase. The mixing and circulation times were observed to increase with an increase in the liquid viscosity, while the liquid phase axial dispersion coefficients decreased.

8) The effect of two immiscible liquid systems on riser and downcomer gas holdups was complex. On addition of an immiscible liquid (oil) to the continuous phase (water) there was a sharp increase in both the riser and downcomer gas holdups, at a given gas velocity. But, as the oil-in-water concentration increases the riser and downcomer gas holdups decrease. Then as the pure oil (dispersed phase) limit was reached the riser and downcomer holdups increase. The riser and downcomer liquid velocities and liquid mixing deteriorate with an increase in the oil-in-water concentration.

9) The $A_d/A_r$ ratio has a strong influence on the hydrodynamics and liquid mixing in the airlift vessel. A lower $A_d/A_r$ ratio in the draft tube internal loop airlift contactor resulted in a higher gas holdup and a lower liquid circulation velocity. Poor liquid mixing was also observed with the lower $A_d/A_r$ ratio airlift geometry, primarily due to the reduced liquid circulation velocity.

10) The riser gas holdup and liquid velocity data obtained experimentally have been analysed in terms of a hydrodynamic model. Relatively good agreement between model predictions and experimental measurements were observed. Additionally the hydrodynamic model was also used on appropriate experimental data in the literature. In most cases, the agreement between the experimental data and model predictions was reasonable.

11) An equation to predict the liquid circulation time in an airlift reactor is given. The predicted liquid circulation time values were compared with the measured ones. The agreement between the two were typically within ±20%.

12) For non-Newtonian media, an equation for the average shear rate in an airlift vessel was developed. The effect of gas holdup, liquid velocity, column diameter and flow behaviour index on the shear rate are considered in this correlation. At a given gas flow rate, a higher apparent liquid viscosity resulted in a lower average shear rate value in the
vessel than a lower apparent liquid viscosity. This was due to the fact that a lower apparent liquid viscosity gave a higher liquid velocity and hence a higher turbulence intensity in the airlift vessel.

6.2 Recommendations

The following suggestions for any possible future research can be made:

1) The gas-liquid separator can have an important effect on the liquid mixing in an airlift reactor. For this reason, values of Bodenstein numbers and liquid phase axial dispersion coefficients in the top section of the vessel with different Newtonian and non-Newtonian media need to be obtained. It would be interesting to see what effect, if any, the liquid phase properties have on the separator Bo numbers and dispersion coefficients.

2) Further study is needed on a larger scale (i.e. pilot plant) to investigate the effect of liquid phase electrolyte concentration and Newtonian and non-Newtonian liquid viscosity on airlift hydrodynamics and mixing. This would indicate whether scale-up has any effect on the influence of the liquid phase properties on airlift hydrodynamics and liquid mixing.

3) Local Bo number and liquid phase axial dispersion studies in a pilot plant airlift reactor with fermentation media should prove to be very useful. This would indicate which sections of the vessel approach plug flow behaviour and which sections are well mixed.

4) More work needs to be carried out with two immiscible liquid systems in airlift vessels. Knowledge is required on the bubble size distribution in an airlift contactor with oil-in-water emulsions.

5) The hydrodynamic model presented in this work can be extended to include a three phase situation (solid-liquid-gas) in an airlift contactor. For example, Ganzeveld et al. (1995) have applied Chisti et al.'s (1988) two phase (gas-liquid) model to a three phase system (gas-liquid-microcarrier). Ganzeveld et al. concluded that the induced liquid circulation velocity in the three phase airlift reactors can be described by the equation of Chisti et al. even though the equation was originally developed for gas-liquid flow systems.

6) The applicability of the shear rate equation developed in this work must be demonstrated over a broad range of airlift vessel geometry and non-Newtonian media.
Appendices

7 APPENDICES

Appendix 1 Physical properties of the liquid phase

This appendix contains the physical properties of the various liquids employed in this study (Section 4.3). Twenty different liquids were used: (a) water and three different concentrations of electrolyte solutions (b) six various concentrations of glycerol-water solutions (c) five different concentrations of CMC solutions (d) food-grade oil and four different concentrations of oil-in-water emulsions.

The liquid physical properties in the internal loop airlift (1) contactor is given in Table A.1, while the liquid properties in the internal loop airlift (2) vessel is presented in Table A.2. Table A.3 gives the liquid physical properties in the external loop airlift reactor. For Newtonian fluids the temperature, density and viscosity of the liquid is given. The temperature and density of the liquid is also given for the non-Newtonian fluids, together with the flow behaviour index, consistency index and the apparent viscosity of the liquid.

With glycerol, the highest concentration of glycerol-water solution which could be employed was deemed to be about 86.0% by volume. This was because of local gas holdup measurements, where with very viscous liquids (above 86.0% vol.) problems occurred (the flow of the liquid through the 3 mm diameter PVC tubing was extremely slow). Also, the liquid circulation velocity and liquid mixing measurements could not be performed satisfactorily with very viscous glycerol-water solutions. Similarly, with CMC solutions the highest concentration was found to be about 0.6% vol. which would provide reasonable gas holdup, liquid velocity and liquid mixing experimental data.
### a) Water and electrolytes

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<th>Temperature (°C)</th>
<th>Density (kg/m³)</th>
<th>Viscosity (Pa s)</th>
</tr>
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<tr>
<td>Water</td>
<td>18.0</td>
<td>1000.0</td>
<td>0.0010</td>
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<tr>
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<td>19.0</td>
<td>1001.5</td>
<td>0.0009</td>
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<td>0.2 M NaCl</td>
<td>18.5</td>
<td>1009.0</td>
<td>0.0012</td>
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<td>0.5 M NaCl</td>
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<td>1014.0</td>
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### b) Glycerol solutions

<table>
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<th>Glycerol Concentration (%)</th>
<th>Temperature (°C)</th>
<th>Density (kg/m³)</th>
<th>Viscosity (Pa s)</th>
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<td>86.0</td>
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### c) CMC solutions

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<th>Density (kg/m³)</th>
<th>Flow behaviour index (-)</th>
<th>Consistency index (Pa sⁿ)</th>
<th>Apparent viscosity range (Pa s)</th>
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<td>0.4</td>
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### d) Oil/water emulsions

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<th>Flow behaviour index (-)</th>
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<td>0.0762</td>
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Table A.1: Liquid phase physical properties: Internal loop airlift (1) contactor.
### Appendices

#### a) Water and electrolytes

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<tr>
<th>Liquid</th>
<th>Temperature (°C)</th>
<th>Density (kg/m³)</th>
<th>Viscosity (Pa s)</th>
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</thead>
<tbody>
<tr>
<td>Water</td>
<td>23.2</td>
<td>1000.0</td>
<td>0.0009</td>
</tr>
<tr>
<td>0.1 M NaCl</td>
<td>23.5</td>
<td>1001.0</td>
<td>0.0010</td>
</tr>
<tr>
<td>0.2 M NaCl</td>
<td>24.3</td>
<td>1005.0</td>
<td>0.0011</td>
</tr>
<tr>
<td>0.5 M NaCl</td>
<td>24.3</td>
<td>1018.0</td>
<td>0.0012</td>
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#### b) Glycerol solutions

<table>
<thead>
<tr>
<th>Glycerol Concentration (%)</th>
<th>Temperature (°C)</th>
<th>Density (kg/m³)</th>
<th>Viscosity (Pa s)</th>
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<td>48.5</td>
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<td>86.0</td>
<td>25.3</td>
<td>1217.0</td>
<td>0.0759</td>
</tr>
</tbody>
</table>

#### c) CMC solutions

<table>
<thead>
<tr>
<th>CMC Concentration (%)</th>
<th>Temperature (°C)</th>
<th>Density (kg/m³)</th>
<th>Flow behaviour index (-)</th>
<th>Consistency index (Pa sⁿ)</th>
<th>Apparent viscosity range (Pa s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.05</td>
<td>28.4</td>
<td>999.9</td>
<td>0.94</td>
<td>0.0173</td>
<td>0.0145-0.0154</td>
</tr>
<tr>
<td>0.2</td>
<td>27.6</td>
<td>1000.3</td>
<td>0.81</td>
<td>0.0526</td>
<td>0.0300-0.0364</td>
</tr>
<tr>
<td>0.3</td>
<td>27.6</td>
<td>1003.5</td>
<td>0.71</td>
<td>0.1029</td>
<td>0.0562-0.0773</td>
</tr>
<tr>
<td>0.4</td>
<td>29.0</td>
<td>1006.5</td>
<td>0.61</td>
<td>0.2620</td>
<td>0.1482-0.2049</td>
</tr>
<tr>
<td>0.6</td>
<td>28.4</td>
<td>1007.5</td>
<td>0.58</td>
<td>0.5610</td>
<td>0.3112-0.4593</td>
</tr>
</tbody>
</table>

#### d) Oil/water emulsions

<table>
<thead>
<tr>
<th>Oil Concentration in water (%)</th>
<th>Temperature (°C)</th>
<th>Density (kg/m³)</th>
<th>Flow behaviour index (-)</th>
<th>Consistency index (Pa sⁿ)</th>
</tr>
</thead>
<tbody>
<tr>
<td>10</td>
<td>20.8</td>
<td>993.8</td>
<td>1.13</td>
<td>0.0007</td>
</tr>
<tr>
<td>20</td>
<td>24.1</td>
<td>985.8</td>
<td>0.96</td>
<td>0.0023</td>
</tr>
<tr>
<td>40</td>
<td>24.1</td>
<td>971.8</td>
<td>0.88</td>
<td>0.0146</td>
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<tr>
<td>60</td>
<td>20.8</td>
<td>960.5</td>
<td>0.83</td>
<td>0.0226</td>
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<tr>
<td>100 (no water)</td>
<td>22.3</td>
<td>923.8</td>
<td>0.94</td>
<td>0.0965</td>
</tr>
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</table>

Table A.2: Liquid phase physical properties: Internal loop airlift (2) contactor.
### a) Water and electrolytes

<table>
<thead>
<tr>
<th>Liquid</th>
<th>Temperature (°C)</th>
<th>Density (kg/m³)</th>
<th>Viscosity (Pa s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Water</td>
<td>17.8</td>
<td>1000.0</td>
<td>0.0008</td>
</tr>
<tr>
<td>0.1 M NaCl</td>
<td>18.5</td>
<td>1002.0</td>
<td>0.0009</td>
</tr>
<tr>
<td>0.2 M NaCl</td>
<td>18.0</td>
<td>1008.0</td>
<td>0.0011</td>
</tr>
<tr>
<td>0.5 M NaCl</td>
<td>18.3</td>
<td>1016.0</td>
<td>0.0012</td>
</tr>
</tbody>
</table>

### b) Glycerol solutions

<table>
<thead>
<tr>
<th>Glycerol Concentration (%)</th>
<th>Temperature (°C)</th>
<th>Density (kg/m³)</th>
<th>Viscosity (Pa s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>48.5</td>
<td>19.9</td>
<td>1108.3</td>
<td>0.0044</td>
</tr>
<tr>
<td>57.5</td>
<td>19.4</td>
<td>1159.1</td>
<td>0.0125</td>
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<tr>
<td>71.5</td>
<td>19.3</td>
<td>1180.7</td>
<td>0.0220</td>
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<tr>
<td>78.5</td>
<td>20.8</td>
<td>1209.3</td>
<td>0.0638</td>
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<tr>
<td>82.0</td>
<td>17.8</td>
<td>1214.0</td>
<td>0.0824</td>
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<tr>
<td>86.0</td>
<td>18.0</td>
<td>1226.0</td>
<td>0.1444</td>
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### c) CMC solutions

<table>
<thead>
<tr>
<th>CMC Concentration (%)</th>
<th>Temperature (°C)</th>
<th>Density (kg/m³)</th>
<th>Flow behaviour index (-)</th>
<th>Consistency index (Pa sⁿ)</th>
<th>Apparent viscosity range (Pa s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.05</td>
<td>19.0</td>
<td>1000.0</td>
<td>0.90</td>
<td>0.0119</td>
<td>0.0094-0.0103</td>
</tr>
<tr>
<td>0.2</td>
<td>17.5</td>
<td>998.0</td>
<td>0.80</td>
<td>0.0471</td>
<td>0.0291-0.0358</td>
</tr>
<tr>
<td>0.3</td>
<td>17.3</td>
<td>998.0</td>
<td>0.73</td>
<td>0.0929</td>
<td>0.0491-0.0697</td>
</tr>
<tr>
<td>0.4</td>
<td>16.9</td>
<td>998.9</td>
<td>0.64</td>
<td>0.2740</td>
<td>0.1304-0.1515</td>
</tr>
<tr>
<td>0.6</td>
<td>18.3</td>
<td>998.0</td>
<td>0.57</td>
<td>0.5840</td>
<td>0.2335-0.3602</td>
</tr>
</tbody>
</table>

### d) Oil/water emulsions

<table>
<thead>
<tr>
<th>Oil Concentration in water (%)</th>
<th>Temperature (°C)</th>
<th>Density (kg/m³)</th>
<th>Flow behaviour index (-)</th>
<th>Consistency index (Pa sⁿ)</th>
</tr>
</thead>
<tbody>
<tr>
<td>10</td>
<td>21.6</td>
<td>990.0</td>
<td>1.16</td>
<td>0.0007</td>
</tr>
<tr>
<td>20</td>
<td>26.5</td>
<td>992.3</td>
<td>0.97</td>
<td>0.0019</td>
</tr>
<tr>
<td>40</td>
<td>23.9</td>
<td>971.0</td>
<td>0.90</td>
<td>0.0110</td>
</tr>
<tr>
<td>60</td>
<td>21.6</td>
<td>964.8</td>
<td>0.85</td>
<td>0.0309</td>
</tr>
<tr>
<td>100 (no water)</td>
<td>20.9</td>
<td>919.5</td>
<td>0.92</td>
<td>0.0864</td>
</tr>
</tbody>
</table>

Table A.3: Liquid phase physical properties: External loop airlift contactor.
Appendix 2 The effect of CMC and oil-in-water concentrations on the flow behaviour index and the consistency index

This appendix contains two figures:
(1) The typical variation of the flow behaviour index and the consistency index as a function of the concentration of CMC solution used in this study.
(2) The effect of the oil-in-water concentration, employed in this work, on the flow behaviour index and the consistency index.

Figure A.1: The effect of CMC concentration on the flow behaviour index and the consistency index in the external loop airlift contactor.
Figure A.2: The variation of the flow behaviour index and the consistency index as a function of the oil-in-water concentration. The experimental data were obtained in the external loop airlift vessel.
Appendices

Appendix 3 The effect of electrolyte concentration, liquid viscosity and flow behaviour index on the terminal bubble rise velocity

This appendix contains the following figures:
(1) The influence of liquid phase salt concentration on the terminal bubble rise velocity.
(2) The effect of liquid phase viscosity on the terminal bubble rise velocity.
(3) How the terminal bubble rise velocity is influenced by the flow behaviour index of the non-Newtonian liquid phase.

Figure A.3: The effect of electrolyte concentration on the terminal bubble rise velocity. The experimental data were obtained in the external loop airlift reactor.
Figure A.4: The influence of liquid viscosity on the terminal bubble rise velocity. The experimental data were obtained in the external loop airlift vessel.

Figure A.5: The effect of flow behaviour index on the terminal bubble rise velocity. The experimental measurements were obtained in the external loop airlift contactor.
Appendices

Appendix 4 Published papers and conference presentations

In this appendix, a list of the published papers and conference presentations undertaken during the course of this doctoral work (from the 4th of October 1993 to the 12th of July 1996) is given.

(1) Rajarajan, J., Djeyvedjian, O. and Ayazi Shamlou, P. 1995
Effect of viscosity on gas holdup and liquid circulation rate in airlift devices

(2) Rajarajan, J. 1995
Gas-liquid mixing in airlift columns

(3) Rajarajan, J. and Ayazi Shamlou, P. 1995
Hydrodynamic modelling of an external loop airlift reactor for wastewater treatment
Paper presented (oral presentation) at the 31st AWRA Annual Conference, Advances in the Development and Use of Models in Water Resources, Houston, Texas, 4-11 Nov., p.283-288. [This paper won the 1995 Hydrolab Corporation's Best Student Paper Award.]

(4) Rajarajan, J. and Ayazi Shamlou, P. 1995
Hydrodynamic aspects of two-phase airlift contactors

Gas holdup and liquid velocity in airlift bioreactors containing viscous Newtonian liquids

(6) Rajarajan, J. and Ayazi Shamlou, P. 1996
Non-Newtonian two phase circulation in airlift contactors
Appendices

(7) Rajarajan, J. 1996
Mixing in airlift devices
An oral presentation given at the *IChemE Fluid Mixing Subject Group Meeting*, Birmingham, 22nd of May.

(8) Rajarajan, J. and Ayazi Shamlou, P. 1996
Mixing and dispersion studies in an external loop airlift reactor
Paper presented (oral presentation) at the *UK Conference on Mixing (Fluid Mixing 5)*, IChemE Symposium Series No. 140, Bradford, 4-5 Jul., p.271-282.

(9) Rajarajan, J. and Ayazi Shamlou, P. 1996
Hydrodynamic aspects of airlift reactors
Paper to be presented (poster presentation) at the *Fifth World Congress of Chemical Engineering* (Advances in Mathematical Modelling in Biotechnology), San Diego, California, 14-18 Jul.

(10) Ayazi Shamlou, P., Rajarajan, J. and Ison A. P. 1996
Gas holdup, liquid circulation velocity and shear rate in airlift reactors
Paper recently submitted to the *Bioprocess Eng.* for publication.
Nomenclature

8 NOMENCLATURE

Roman Symbols

<table>
<thead>
<tr>
<th>Symbol</th>
<th>Description</th>
<th>Units</th>
</tr>
</thead>
<tbody>
<tr>
<td>a</td>
<td>Gas-liquid interfacial area per unit liquid volume</td>
<td>( m^{-1} )</td>
</tr>
<tr>
<td>A</td>
<td>Total cross-sectional area of reactor</td>
<td>( m^2 )</td>
</tr>
<tr>
<td>Ab</td>
<td>Free area below baffle/draught-tube</td>
<td>( m^2 )</td>
</tr>
<tr>
<td>Ad</td>
<td>Downcomer cross-sectional area</td>
<td>( m^2 )</td>
</tr>
<tr>
<td>An</td>
<td>Cross-sectional area of one nozzle opening</td>
<td>( m^2 )</td>
</tr>
<tr>
<td>Ar</td>
<td>Riser cross-sectional area</td>
<td>( m^2 )</td>
</tr>
<tr>
<td>A_u</td>
<td>Unit area</td>
<td>( m^2 )</td>
</tr>
<tr>
<td>Bo</td>
<td>Bodenstein number</td>
<td>(-)</td>
</tr>
<tr>
<td>C</td>
<td>Instantaneous tracer concentration</td>
<td>( kgm^{-3} )</td>
</tr>
<tr>
<td>( \dot{C} )</td>
<td>Dimensionless concentration</td>
<td>(-)</td>
</tr>
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<td>C_e</td>
<td>Equilibrium tracer concentration</td>
<td>( kgm^{-3} )</td>
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<tr>
<td>C_l</td>
<td>Concentration of the transferring component in the liquid phase</td>
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</tr>
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<td>C_o</td>
<td>Distribution parameter</td>
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<tr>
<td>d_b</td>
<td>Bubble diameter</td>
<td>(m)</td>
</tr>
<tr>
<td>d_c</td>
<td>Column diameter (or equivalent hydraulic diameter)</td>
<td>(m)</td>
</tr>
<tr>
<td>d_d</td>
<td>Downcomer diameter (or equivalent hydraulic diameter)</td>
<td>(m)</td>
</tr>
<tr>
<td>d_p</td>
<td>Pipe diameter</td>
<td>(m)</td>
</tr>
<tr>
<td>d_s</td>
<td>Diameter of a sphere having the same volume as a particle or a bubble</td>
<td>((6v_b / \pi)^{1/3})</td>
</tr>
<tr>
<td>d_sh</td>
<td>Diameter of the sparger hole</td>
<td>(m)</td>
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<tr>
<td>d_l</td>
<td>Diameter of the gas-liquid separator</td>
<td>(m)</td>
</tr>
<tr>
<td>d_vs</td>
<td>Volume-to-surface mean bubble diameter</td>
<td>(m)</td>
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<tr>
<td>D</td>
<td>Liquid phase axial dispersion coefficient</td>
<td>( m^2s^{-1} )</td>
</tr>
<tr>
<td>E_b</td>
<td>Energy dissipation in liquid motion</td>
<td>(W)</td>
</tr>
<tr>
<td>E_f</td>
<td>Energy dissipation resulting from change in flow direction</td>
<td>(W)</td>
</tr>
<tr>
<td>E_i</td>
<td>Energy input by the gas</td>
<td>(W)</td>
</tr>
<tr>
<td>E_w</td>
<td>Energy dissipation due to bubble wake</td>
<td>(W)</td>
</tr>
</tbody>
</table>
## Nomenclature

<table>
<thead>
<tr>
<th>Symbol</th>
<th>Description</th>
<th>Units</th>
</tr>
</thead>
<tbody>
<tr>
<td>F</td>
<td>Shearing force</td>
<td>(kgms⁻²)</td>
</tr>
<tr>
<td>Fr</td>
<td>Overall Froude number ([U_{sgo} / (gd_c)^{0.5}])</td>
<td>(-)</td>
</tr>
<tr>
<td>g</td>
<td>Gravitational acceleration</td>
<td>(ms⁻²)</td>
</tr>
<tr>
<td>&lt;j&gt;</td>
<td>Average volumetric flux density of the mixture</td>
<td>(ms⁻¹)</td>
</tr>
<tr>
<td>J_{gl}</td>
<td>Gas drift flux</td>
<td>(ms⁻¹)</td>
</tr>
<tr>
<td>hₚ</td>
<td>Distance between two thermocouple probes</td>
<td>(m)</td>
</tr>
<tr>
<td>H</td>
<td>Unaerated riser liquid height</td>
<td>(m)</td>
</tr>
<tr>
<td>H_d</td>
<td>Unaerated downcomer liquid height</td>
<td>(m)</td>
</tr>
<tr>
<td>H_{do}</td>
<td>Height of the downcomer section</td>
<td>(m)</td>
</tr>
<tr>
<td>H_D</td>
<td>Height of gas-liquid dispersion</td>
<td>(m)</td>
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<td>H_r</td>
<td>Height of the riser section</td>
<td>(m)</td>
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<td>H_t</td>
<td>Overall (total) height</td>
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<tr>
<td>k</td>
<td>Ratio of liquid-wake volume to bubble volume</td>
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<tr>
<td>K</td>
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<td>K_b</td>
<td>Loss coefficient for the bottom</td>
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<td>K_f</td>
<td>Frictional loss coefficient</td>
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<tr>
<td>K_t</td>
<td>Loss coefficient for the top</td>
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<td>K_l</td>
<td>Mass transfer coefficient based on liquid-film</td>
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<td>K_{L,a}</td>
<td>Volumetric mass transfer coefficient</td>
<td>(s⁻¹)</td>
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<td>L</td>
<td>Circulation path length of the reactor</td>
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<td>Draught-tube length</td>
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<td>L_{cd}</td>
<td>Equivalent length of the downcomer</td>
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<td>L_{er}</td>
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<tr>
<td>M</td>
<td>Molar mass</td>
<td>(kgkmol⁻¹)</td>
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<tr>
<td>Mo</td>
<td>Morton number ([gμₜ^4 / ρₜ \sigmaₜ^3])</td>
<td>(-)</td>
</tr>
<tr>
<td>n</td>
<td>Flow behaviour index</td>
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<tr>
<td>N</td>
<td>Number of circulations required for complete mixing</td>
<td>(-)</td>
</tr>
<tr>
<td>P</td>
<td>Pressure</td>
<td>(Pa)</td>
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<tr>
<td>P_b</td>
<td>Pressure at the bottom of the fluid in the reactor</td>
<td>(Pa)</td>
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<tr>
<td>P_g</td>
<td>Power input due to gassing</td>
<td>(W)</td>
</tr>
<tr>
<td>P_h</td>
<td>Headspace pressure of the fluid in the reactor</td>
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</tr>
<tr>
<td>P_e</td>
<td>Peclet number</td>
<td>(-)</td>
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<tr>
<td>Symbol</td>
<td>Description</td>
<td>Unit</td>
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<td>--------</td>
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<tr>
<td>( \dot{Q}_g )</td>
<td>Volumetric gas flow rate</td>
<td>( (m^3 \cdot s^{-1}) )</td>
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<td>( \dot{Q}_m )</td>
<td>Molar gas flow rate</td>
<td>( (kmol^{-1}) )</td>
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<tr>
<td>( r )</td>
<td>Radial position</td>
<td>(m)</td>
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<tr>
<td>( R )</td>
<td>Column radius</td>
<td>(m)</td>
</tr>
<tr>
<td>( R_b )</td>
<td>Equivalent bubble radius</td>
<td>(m)</td>
</tr>
<tr>
<td>( R_g )</td>
<td>Gas constant (= 8314 JK(^{-1})kmol(^{-1}))</td>
<td></td>
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<tr>
<td>( R_i )</td>
<td>Radius of inner cylinder</td>
<td>(m)</td>
</tr>
<tr>
<td>( R_o )</td>
<td>Radius of outer cylinder</td>
<td>(m)</td>
</tr>
<tr>
<td>( \text{Re} )</td>
<td>Riser Reynolds number ( [\rho_l H_r V_{lr} / \mu_l] )</td>
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<tr>
<td>( t )</td>
<td>Time</td>
<td>(s)</td>
</tr>
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<td>( t_c )</td>
<td>Circulation time</td>
<td>(s)</td>
</tr>
<tr>
<td>( t_m )</td>
<td>Mixing time</td>
<td>(s)</td>
</tr>
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<td>( t_t )</td>
<td>Reactor temperature</td>
<td>(K)</td>
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<tr>
<td>( T )</td>
<td>Diameter of the riser section</td>
<td>(m)</td>
</tr>
<tr>
<td>( U_{bl} )</td>
<td>Linear liquid velocity due to density difference between riser and downcomer</td>
<td>(ms(^{-1}))</td>
</tr>
<tr>
<td>( U_{blr} )</td>
<td>Axial liquid velocity in a bubble column as a function of the radial position ( r )</td>
<td>(ms(^{-1}))</td>
</tr>
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<td>( U_B )</td>
<td>Bubble rise velocity relative to average mixture velocity</td>
<td>(ms(^{-1}))</td>
</tr>
<tr>
<td>( U_{bt} )</td>
<td>Terminal rise velocity of an isolated bubble</td>
<td>(ms(^{-1}))</td>
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<td>( U_c )</td>
<td>Linear liquid velocity at the column axis</td>
<td>(ms(^{-1}))</td>
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<tr>
<td>( U_{cr} )</td>
<td>Speed of the chart recorder</td>
<td>(ms(^{-1}))</td>
</tr>
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<td>Gas velocity in the sparger orifice</td>
<td>(ms(^{-1}))</td>
</tr>
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<td>( U_{sg} )</td>
<td>Superficial gas velocity in the riser</td>
<td>(ms(^{-1}))</td>
</tr>
<tr>
<td>( U_{sgo} )</td>
<td>Superficial gas velocity based on overall column cross section</td>
<td>(ms(^{-1}))</td>
</tr>
<tr>
<td>( U_{sg}^o )</td>
<td>( U_{sg} ) corrected for recirculating gas</td>
<td>(ms(^{-1}))</td>
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<td>( U_{sl} )</td>
<td>Superficial liquid velocity in the riser</td>
<td>(ms(^{-1}))</td>
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<tr>
<td>( U_{sld} )</td>
<td>Superficial liquid velocity in the downcomer</td>
<td>(ms(^{-1}))</td>
</tr>
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<td>( U_{slo} )</td>
<td>Overall superficial liquid velocity</td>
<td>(ms(^{-1}))</td>
</tr>
<tr>
<td>( U_w )</td>
<td>Linear liquid velocity due to liquid-wake</td>
<td>(ms(^{-1}))</td>
</tr>
<tr>
<td>( v_b )</td>
<td>Volume of a bubble</td>
<td>(m(^3))</td>
</tr>
<tr>
<td>( v_r )</td>
<td>Gas-liquid slip velocity</td>
<td>(ms(^{-1}))</td>
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<tr>
<td>( V )</td>
<td>Volume of liquid in reactor</td>
<td>(m(^3))</td>
</tr>
</tbody>
</table>
### Nomenclature

<table>
<thead>
<tr>
<th>Symbol</th>
<th>Definition</th>
<th>Unit</th>
</tr>
</thead>
<tbody>
<tr>
<td>$V_d$</td>
<td>Volume of dispersion</td>
<td>($m^3$)</td>
</tr>
<tr>
<td>$V_g$</td>
<td>Gas volume in reactor</td>
<td>($m^3$)</td>
</tr>
<tr>
<td>$V_l$</td>
<td>Volume of two phase mixture</td>
<td>($m^3$)</td>
</tr>
<tr>
<td>$V_1$</td>
<td>Average overall liquid circulation velocity</td>
<td>($ms^{-1}$)</td>
</tr>
<tr>
<td>$V_{ld}$</td>
<td>Interstitial liquid velocity in the downcomer</td>
<td>($ms^{-1}$)</td>
</tr>
<tr>
<td>$V_{lr}$</td>
<td>Interstitial liquid velocity in the riser</td>
<td>($ms^{-1}$)</td>
</tr>
<tr>
<td>$W_e$</td>
<td>Weber number $[V_l(H_l\rho_l / \sigma_l)^{0.5}]$</td>
<td>(-)</td>
</tr>
<tr>
<td>$X$</td>
<td>Dimensionless distance</td>
<td>(-)</td>
</tr>
<tr>
<td>$X_{cr}$</td>
<td>Distance between two consecutive peaks on the chart</td>
<td>(m)</td>
</tr>
<tr>
<td>$z$</td>
<td>Total distance travelled</td>
<td>(m)</td>
</tr>
</tbody>
</table>

### Greek Symbols

<table>
<thead>
<tr>
<th>Symbol</th>
<th>Definition</th>
<th>Unit</th>
</tr>
</thead>
<tbody>
<tr>
<td>$\alpha$</td>
<td>Volume concentration</td>
<td>(-)</td>
</tr>
<tr>
<td>$\alpha_{av}$</td>
<td>Average volume concentration</td>
<td>(-)</td>
</tr>
<tr>
<td>$\alpha_c$</td>
<td>$\alpha$ at the centre line of the circular pipe</td>
<td>(-)</td>
</tr>
<tr>
<td>$\alpha_w$</td>
<td>$\alpha$ at the wall of the circular pipe</td>
<td>(-)</td>
</tr>
<tr>
<td>$\gamma$</td>
<td>Average shear rate</td>
<td>($s^{-1}$)</td>
</tr>
<tr>
<td>$\gamma_w$</td>
<td>Shear rate at the wall</td>
<td>($s^{-1}$)</td>
</tr>
<tr>
<td>$\Gamma$</td>
<td>Torque per unit length</td>
<td>(kgm$s^{-2}$)</td>
</tr>
<tr>
<td>$\Delta\sigma_{d}^2$</td>
<td>Change in variance of pulse of tracer due to passage through downcomer</td>
<td>($s^2$)</td>
</tr>
<tr>
<td>$\Delta\sigma_o^2$</td>
<td>The overall change in variance of pulse of tracer due to passage through the reactor</td>
<td>($s^2$)</td>
</tr>
<tr>
<td>$\Delta\sigma_r^2$</td>
<td>Change in variance of pulse of tracer due to passage through riser</td>
<td>($s^2$)</td>
</tr>
<tr>
<td>$\Delta\sigma_g^2$</td>
<td>Change in variance of pulse of tracer due to passage through gas-liquid separator</td>
<td>($s^2$)</td>
</tr>
<tr>
<td>$\Delta\rho$</td>
<td>Gas-liquid density difference</td>
<td>(kgm$^{-3}$)</td>
</tr>
<tr>
<td>$\varepsilon_d$</td>
<td>Gas holdup in the downcomer</td>
<td>(-)</td>
</tr>
<tr>
<td>$\varepsilon_g$ (or $E_g$)</td>
<td>Gas holdup in the riser</td>
<td>(-)</td>
</tr>
<tr>
<td>$\varepsilon_l$</td>
<td>Liquid holdup</td>
<td>(-)</td>
</tr>
</tbody>
</table>
Nomenclature

<table>
<thead>
<tr>
<th>Symbol</th>
<th>Description</th>
<th>Unit</th>
</tr>
</thead>
<tbody>
<tr>
<td>$\varepsilon_o$</td>
<td>Overall gas holdup</td>
<td>(-)</td>
</tr>
<tr>
<td>$\varepsilon_w$</td>
<td>Liquid-wake holdup</td>
<td>(-)</td>
</tr>
<tr>
<td>$\eta$</td>
<td>Efficiency (in Eqn 2.1)</td>
<td>(-)</td>
</tr>
<tr>
<td>$\theta$</td>
<td>Dimensionless time</td>
<td>(-)</td>
</tr>
<tr>
<td>$\theta_v$</td>
<td>Angular velocity</td>
<td>(s$^{-1}$)</td>
</tr>
<tr>
<td>$\mu$</td>
<td>Newtonian viscosity</td>
<td>(Pa s)</td>
</tr>
<tr>
<td>$\mu_a$</td>
<td>Apparent viscosity</td>
<td>(Pa s)</td>
</tr>
<tr>
<td>$\mu_L$</td>
<td>Viscosity of liquid</td>
<td>(Pa s)</td>
</tr>
<tr>
<td>$\mu_v$</td>
<td>Plastic viscosity</td>
<td>(Pa s)</td>
</tr>
<tr>
<td>$\mu_w$</td>
<td>Viscosity of water</td>
<td>(Pa s)</td>
</tr>
<tr>
<td>$\mu_{oo}$</td>
<td>Viscosity at infinite shear</td>
<td>(Pa s)</td>
</tr>
<tr>
<td>$\nu$</td>
<td>Kinematic viscosity</td>
<td>(m$^2$s$^{-1}$)</td>
</tr>
<tr>
<td>$\nu_a$</td>
<td>Apparent kinematic viscosity</td>
<td>(m$^2$s$^{-1}$)</td>
</tr>
<tr>
<td>$\rho_G$</td>
<td>Density of gas</td>
<td>(kgm$^{-3}$)</td>
</tr>
<tr>
<td>$\rho_L$</td>
<td>Density of liquid</td>
<td>(kgm$^{-3}$)</td>
</tr>
<tr>
<td>$\rho_m$</td>
<td>Density of manometer fluid</td>
<td>(kgm$^{-3}$)</td>
</tr>
<tr>
<td>$\sigma$</td>
<td>Surface tension</td>
<td>(Nm$^{-1}$)</td>
</tr>
<tr>
<td>$\sigma_L$</td>
<td>Surface tension of liquid</td>
<td>(Nm$^{-1}$)</td>
</tr>
<tr>
<td>$\tau$</td>
<td>Shear stress</td>
<td>(Pa)</td>
</tr>
<tr>
<td>$\tau_w$</td>
<td>Shear stress at the wall</td>
<td>(Pa)</td>
</tr>
<tr>
<td>$\tau_y$</td>
<td>Yield stress</td>
<td>(Pa)</td>
</tr>
</tbody>
</table>
9 REFERENCES


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BHRA *International Conference on Bioreactor Fluid Dynamics I* 1986. Cambridge, UK, Apr., Editor: King, R.

BHRA *International Conference on Bioreactor Fluid Dynamics II* 1988. Cambridge, UK, Sep., Editor: King, R.


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