THE EFFECT OF LIQUID PHYSICAL PROPERTIES ON GAS-LIQUID EJECTOR PERFORMANCE

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This thesis extends previous work that has been carried out in an air-water system in a gas-liquid ejector. The thesis presents the experimental programme undertaken and contains a survey of the available literature on the effects of liquid physical properties in gas-liquid ejector systems. The survey concentrates on the effects of coalescence inhibition and touches on viscosity effects. Due to the lack of research in such systems, the literature review was extended to look at the work carried out in stagnant fluids and in other systems such as stirred tank reactors and bubble columns. This was used to identify a fluid that would represent a non-coalescing system.

The experimental work was split into three parts: the effect a) of flow parameters on gas holdup and bubble size in an air-water system; b) of coalescence inhibition and c) of liquid viscosity on ejector performance. Parameters measured in order to establish the ejector performance in the coalescence inhibited and viscous solutions were gas entrainment, ejector pumping efficiency, bubble size, gas holdup and mass transfer.

A non-coalescing fluid was simulated using a 0.08M Magnesium Sulphate solution. It is believed, based on results reported by various researchers that this concentration was sufficient to achieve a near non-coalescing fluid. Various experiments were carried out and the above mentioned parameters measured to assess the ejector performance against a pure water system.

Experimental results showed that coalescence inhibition has little effect on gas entrainment or gas holdup (and hence slip-velocity). However, the Sauter mean diameter decreases with the non-coalescing nature and the bubbles were seen to be more spherical and appeared more rigid. Values of $k_f a$ are similar between coalescing and non-coalescing systems: this is due to an increase in $a$ and at the same time a corresponding decrease in $k_L$, which may be due to a decrease in diffusivity.

Newtonian CMC solution provided a viscous system over a range of viscosities (3 - 10 mPas). Comparisons were made with water (1 mPas) for the ejector performance parameters. Experimental results showed that viscosity has a small effect on gas entrainment and ejector pumping efficiency and little effect on gas holdup (and hence slip-velocity). Sauter mean diameter was found to decrease with an increase in liquid viscosity, producing more spherical bubbles. Values of $k_f a$ were smaller for an increase in viscosity, despite an increase in $a$. This is due to a more significant decrease in $k_L$, due to the effect of liquid viscosity reducing the diffusivity.
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# NOMENCLATURE

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<th>SYMBOL</th>
<th>MEANING</th>
<th>UNITS</th>
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<td>Specific interfacial area</td>
<td>m(^{-1})</td>
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<tr>
<td>$A_m$</td>
<td>Mixing tube cross-sectional area</td>
<td>m(^2)</td>
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<tr>
<td>$b$</td>
<td>Constant in Equations 6 and 7</td>
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<tr>
<td>$c$</td>
<td>Viscosity exponent in relation to $k_L$</td>
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<td>Concentration of ion, $i$</td>
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<td>kg m(^{-3})</td>
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<td>$C_2$</td>
<td>Oxygen concentration at mixer outlet</td>
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<td>$d_{\text{p}}$</td>
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<td>$d_{32}$</td>
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<td>Oxygen diffusion coefficient in pure water</td>
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<td>$P$</td>
<td>Power</td>
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<td>Power used for mixing and friction</td>
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<td>Radiation count for when pipe section is empty</td>
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<td>$R_{lv}$</td>
<td>Radiation count for when pipe section is filled with liquid</td>
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<td>Peak-peak residence time between conductivity probes 1 &amp; 2, liquid only mixture</td>
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<td>Time liquid spends on spinner</td>
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<td>$u_{r,0}$</td>
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<td>$v_G$</td>
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<td>Liquid velocity</td>
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<td>Superficial gas velocity</td>
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<td>Mixing tube pressure drop</td>
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<td>Energy dissipation rate</td>
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<td>Ejector pumping efficiency</td>
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<td>$\eta(d_e,d_o)$</td>
<td>Coalescence efficiency</td>
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<td>$\rho$</td>
<td>Liquid density</td>
<td>kgm⁻³</td>
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<tr>
<td>$\sigma$</td>
<td>Standard deviation of bubble diameter</td>
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<td>Angular velocity at nozzle</td>
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Chapter 1 - INTRODUCTION

Gas-liquid mixing is an important factor to consider in many industrial processes. Examples include oxidations, chlorinations, hydrogenations and gas stripping. They all involve the transfer of solute between the gas and liquid phases, and are quite often followed by a chemical reaction. The contacting between the gas and liquid phases can be carried out in a wide range of equipment, such as stirred tank reactors, bubble columns, sieve tray contactors, motionless mixers and ejectors. The mass transfer rates attainable in motionless mixers and ejectors tend to be higher than in the other devices.

The mass transfer rate required between the phases dictates the type and volume of equipment needed. If a high mass transfer rate is available then equipment volume may be reduced. If mass transfer is followed by a chemical reaction, the mass transfer may be the limiting step, and the use of high intensity mixers may be required.

In industrial practice, both the mass transfer and the total specific interfacial area are of interest to the designer. The product of the liquid-side mass transfer coefficient and specific interfacial area gives the important design parameter, $k_L a$. $k_L$ is a function of the diffusion coefficient of the solute in the liquid and the bubble size and $a$ depends on the bubble size and distribution. Both $k_L$ and $a$ are influenced by the fluid dynamics in the equipment, i.e. the flow regimes, turbulent energy dissipation rate and flow patterns. $k_L a$ is an important parameter that can fairly easily be measured. It is dependent on many different parameters: fluid dynamics in the equipment, turbulent energy dissipation rates and also liquid phase physical properties. For example, one important variable in determining the mass transfer in gas-liquid contacting devices is the coalescence rate between the gas bubbles. Industrial processes are rarely carried out in a pure liquid. In most cases suspended particles and dissolved inorganic and/or organic material are present which decisively influence the bubble coalescence.
Optimal reactor design therefore requires a fundamental understanding of the gas dispersion process, interphase mass transfer phenomenon, chemical reaction kinetics, liquid mixing and turbulence and how all of these factors combine and interrelate to produce the required result. Lack of fundamental understanding makes optimal design difficult.

Before attempting to design a reactor for gas dispersion, the process results must be clearly defined e.g. product yield or product quality and more importantly, how this process result is affected by the degree of mixing.

Also, the rate limiting part of the process must be identified. When a reaction requires interphase mass transfer, the rate of production of a desired product depends on both the intrinsic reaction rate and the rate of mass transfer. For processes that are not mass transfer limited, e.g. slow reactions in the liquid phase, the degree of gas dispersion is not critical and a coarse dispersion of gas will suffice - some degree of mechanical agitation will be adequate. For gas-liquid processes where rapid mass transfer is required, e.g. fast chemical reactions in the liquid phase, the interfacial area needs to be maximised. In low-intensity mixing equipment the mass transfer rate is often slower than the intrinsic reaction rate and therefore becomes the rate limiting step. Hence, the use of high-intensity equipment such as ejectors which can increase the mass transfer to a rate that more closely matches the reaction rate is much more suitable.

An ejector consists of a nozzle and gas chamber followed by a mixing tube. The nozzle provides a high velocity liquid jet which creates suction in the gas chamber, entraining the gas phase into the device. The two phases experience an intensive mixing zone in the early part of the mixing tube, known as the mixing shock. This intensive mixing zone is a result of rapid dissipation of the kinetic energy of the liquid jet as it attaches to the mixing tube wall. A fine dispersion of bubbles forms after the mixing zone, created by the high turbulence and resulting in high interfacial areas in the ejector mixing tube. The device is known to have near plug flow characteristics for both gas and liquid
phases, therefore is effective for complex reactions where the intermediate product is the desired one.

The use of ejectors as reactors can bring other benefits. The device is compact, thus requiring small sites. There are no moving parts within the mixer itself, reducing problems of sealing, high pressure conditions etc. Due to the characteristics of the ejector, a compressor is not required to introduce the gas into the mixing device. Large gas flowrates can be passed through the ejector, whereas with stirred tanks, high gas flowrates would lead to flooding. High, uniform energy dissipation rates are achievable in ejectors as well as narrow residence time distributions, corresponding to characteristics of plug flow.

To date, ejectors have received a limited amount of attention from researchers. Previous work has primarily concentrated on the momentum transfer characteristics of the ejector, such as the amount of gas that may be entrained into the liquid stream at a given flowrate. In more recent work, there has been increased interest in the mass transfer characteristics of ejectors and how the geometry affects the performance. There is reasonable agreement between the various investigations, but also significant inconsistencies. It appears that no two researchers have investigated exactly the same geometries, and the ranges of operating variables e.g. back pressures and gas-liquid flow ratios are quite different.

The majority of this previous research was in air-water or coalescing systems, with a very minor amount carried out in non-coalescing - coalescing systems or systems with increased liquid viscosity. Again, consistency between researchers as to the geometries of the ejector type device were practically non existent. It is seen from the survey that in the available literature, the effects of physical properties such as viscosity, interfacial tension and coalescence behaviour is lacking and there is no work done on the effect of coalescence and viscosity on ejector mass transfer performance.
A considerable amount of research on the effects of liquid physical properties has however been carried out in other types of gas-liquid contacting equipment such as stirred tank reactors and bubble columns and is discussed in the following chapter.

This thesis is based on a continuation of research carried out by Zhu (Z2, Z3). Zhu’s work in a gas-liquid ejector was undertaken in one gas-liquid system only (air-water). The effects of flow parameters and ejector geometry on the entrainment and mass transfer performance of the ejector were studied. Ejector geometry changes were made on the spinner, nozzle, mixing tube length and diameter and scale. The principle conclusions of Zhu’s investigations are as follows:

1) Use of an appropriate spinner upstream of the ejector nozzle stabilises the two-phase flow and reduces equipment size for a given mass transfer rate without additional power consumption.

2) Entrainment performance drops if the swirl number (described in Appendix A) exceeds 0.18. For a swirl number above 0.29, mass transfer also decreases.

3) The ejector’s entrainment and mass transfer performance can be enhanced by increasing either liquid flowrate or nozzle upstream pressure at a given liquid flowrate. For a given geometry and liquid flowrate, the pumping efficiency of the ejector is a function of gas flowrate and exhibits a maximum. The higher the power input is, the higher the maximum efficiency will be.

4) Mixing tube diameter has a complicated effect on the ejector’s performance and selection depends on the actual requirement, whether it is $k_2a$ value, solute transfer rates or entrainment.

5) Reducing mixing tube length such that the mixing tube diameter to length ratio is below 24 impairs the ejector’s performance, but its effect on mass transfer coefficient depends on gas flowrate.
6) A smaller scale ejector, which is geometrically similar to the large scale appears to be less efficient in both entrainment and mass transfer performance.

7) An annular flow pattern develops in the mixing tube at low liquid flowrates or high gas flowrates. In this flow regime, $k_l\alpha$ is independent of gas flowrate. In the bubble flow regime, $k_l\alpha$ was correlated with liquid power input per unit volume and a gas-liquid flow ratio as follows:

$$k_l\alpha = 2.2 \times 10^{-3} \left( \frac{P_T}{V_d} \right)^{0.6} \left( \frac{Q_G}{Q_G + Q_L} \right)^{0.55} \text{ for } \frac{Q_G}{Q_L} \leq 2$$  \hspace{1cm} (Eqn. 1)

$$k_l\alpha = 1.8 \times 10^{-3} \left( \frac{P_T}{V_d} \right)^{0.6} \text{ for } \frac{Q_G}{Q_L} > 2$$  \hspace{1cm} (Eqn. 2)

The objective of the present work was to extend the experimental work on the air-water system to measurement of bubble size and gas holdup and to examine the effect of viscosity and coalescence. The aims were to establish correlations in a gas-liquid ejector enabling the prediction of bubble size, gas holdup and overall mass transfer coefficient in an air-water system, non-coalescing system and viscous system. These correlations would enable the design of a gas-liquid ejector used for contacting a non-coalescing or viscous liquid with a gas phase.
The approach was to carry out a detailed experimental programme to investigate the following in a down flowing gas-liquid ejector:

1) The effect of flow parameters on bubble size and gas holdup in air-water.

2) The effect of coalescence inhibition on gas entrainment, mass transfer, gas holdup and bubble size.

3) The effect of liquid viscosity on gas entrainment, bubble size, gas holdup and mass transfer.

This research therefore sets out to investigate the effect of liquid phase physical properties on gas-liquid ejector performance for a specific geometry of ejector that remains consistent for the coalescing, non-coalescing and viscous systems. The thesis covers investigation of a non-coalescing system or close to that, where the defined coalescence behaviour differs greatly with that of pure water, represented by the use of an aqueous salt solution. The viscosity was increased by the use of various concentrations of sodium carboxymethyl cellulose (CMC) solution up to a viscosity of 10 mPas. The results for these different liquid systems were compared to a pure water system.
Chapter 2 - LITERATURE SURVEY AND REVIEW OF PREVIOUS WORK

A literature survey was carried out to determine what research had been carried out in the area of gas-liquid ejector mixing, preferably regarding the effect of liquid physical properties. It was found that even though a reasonable amount of research has been done in the gas-liquid ejector type mixing area, very little work has been carried out looking at the coalescence effects and even less on viscosity. The literature survey therefore concentrates more on the effect of coalescence behaviour in a gas-liquid system than the viscous effects.

Due to the lack of work on liquid physical properties in gas-liquid ejector type systems, the literature survey was therefore extended to other gas-liquid mixing equipment e.g. stirred tank reactors (STR) and bubble columns. It was found that there has been a considerable amount of research on the effect of salt solutions on bubble coalescence in stagnant fluids and consequently a large number of models have been developed. The information from this literature was used in deciding which salt to use to simulate a non-coalescing fluid and also to help gain an understanding on coalescence between bubbles in differing degrees of coalescing liquids.

The types of mixing devices that the survey extended to included stirred tank reactors and bubble columns where a great deal of research that has been carried out relating to coalescence and viscosity.

The knowledge gained from the work carried out in stagnant fluids, stirred tank reactors and bubble columns, will in later chapters be related to the effects seen in gas-liquid ejector mixing. Since there are no comparable investigations for gas-liquid ejectors, only a qualitative comparison with these devices will be presented.
2.1 GENERAL EFFECTS ON COALESCENCE BEHAVIOUR OF GAS BUBBLES

The interfacial mass transfer rate in a gas-liquid system is controlled by the interfacial area. One of the parameters that affects the interfacial area is the frequency of bubble coalescence within the system.

Salts are found to inhibit bubble coalescence by retarding the thinning of the intervening liquid film between bubble pairs (P3). At sufficiently high concentrations of salt, the gas-liquid interface between coalescing bubbles is immobilized by the surface tension gradient which results from the thinning process. When this occurs, there is a dramatic increase in the time required for coalescence. Film thinning times for bubbles with immobile interfaces are in the order of seconds. Since contact times in turbulent dispersions are in the order of milliseconds, bubble coalescence is not envisaged where the salt concentration is sufficient to immobilize the gas-liquid interface (P4).

Bubble coalescence is considered to be a three step process (O2):

1. The bubbles must come into contact with each other within the liquid phase. The contact is distinguished by a flattening of the bubble surfaces against each other, leaving a thin liquid layer separating them. The initial thickness of this film is controlled by the hydrodynamics of the bulk liquid phase.

2. The intervening liquid film must thin to a dimension of around \(10^{-4}\text{cm}\) before it will rupture. If this thinning time/mechanism takes longer than the bubble contact time, then coalescence will not occur. This step is controlled by the hydrodynamics of the liquid film.

3. Once the film is sufficiently thin, it will rupture via an instability mechanism. This step is very rapid compared to the previous two.
One important factor that distinguishes various film thinning theories is the assumption made about the mobility of the film surface. The surface mobility is restricted due to a gradient in surfactant concentration developing, leading to a gradient in surface tension along the thinning film.

For pure liquid phases, such as water, the bubbles within the liquid are thought to be deformable with fully mobile interfaces. However, surface active materials will frequently be present in the liquid phase which lead to the restriction of the mobility of the film surface. In this case, the film thinning process will be controlled by the viscous effects and occur more slowly.

There are a number of people who have researched the effects of various organic and inorganic fluids on the coalescence behaviour of gas bubbles. Lessard and Zieminski (LI, 1971) studied the effects of inorganic electrolytes on bubble coalescence and interfacial mass transfer. Coalescence was studied by contacting bubble pairs in an aqueous solution, and the coalescence percentage was evaluated as a function of concentration.

In pure water, bubbles were seen to coalesce spontaneously on contact. As a salt was added to the water, a concentration was reached at which coalescence was drastically reduced. This sharp transition concentration was a function of the valency combinations of the salts. This was an effect seen by Zieminski and Whittemore (Z6, 1971), who found that the coalescence was suppressed on addition of inorganic salts in various degrees, depending on the valence of the respective ions and concentration of the salt. These trends can be seen in Figure 1 which show the affect of molar concentration on percentage coalescence frequency for a range of different salts (L1), together with a similar graph which was determined for alcohols (O2, 1986). The transition concentrations were found to correlate well with ionic entropy of solution and with self-diffusion ability of water in solution. Film thinning models have been developed by Oolman and Blanch (O2) for predicting the effects of liquid phase composition on the rate bubbles coalesce in a stagnant fluid based on theories and experiments. Several
experiments were carried out, using a syringe pump which produced pairs of bubbles which grew simultaneously on adjacent orifices in a cell. Visual observations were made of coalescence frequency for a variety of salt and surfactant solutions in water. Systems used in order to vary the degree of coalescence were air-pure liquid phases, air-salt solutions, air-organic surfactants and microbial broths. Organic and microbial broth solutions are not part of the current research, hence they are not covered in detail in this survey.

Oolman and Blanch’s (O2) results were compared to the work by Marrucci and Nicodemo (M3, 1967) and Lessard and Zieminski (L1, 1971) which showed that there is an extremely sharp transition from 100% to very low coalescence rates at an ionic concentration which is dependent on the salt used. These results were used as a basis for the choice of identifying a salt for representing a non-coalescing fluid.

Prince and Blanch (P3) also looked at various salt solutions and examined the amount of salt required to immobilise the gas-liquid interface between coalescing bubbles. Through their work, models were developed using data for various salts and solutions and an expression for the transition concentration of salt necessary to immobilise the gas-liquid interface in coalescing bubbles was determined. Similar effects of various salts on coalescing times and frequencies were also found by Keitel and Onken (K2, 1982), Sagert and Quinn (S1, 1978) and Zahradnik et al. (Z1, 1995). Sagert and Quinn (S1) also correlated coalescing times with respect to the ionic concentration and introduced a dimensionless concentration parameter which was a convenient measure for expressing range of validity of models describing coalescence under conditions where viscous effects are not important. Again, models investigated were those proposed by Marrucci (M2, 1969).

Models were also reviewed by Chesters (C4, 1991). The models concerned were related to the coalescence probability in three particles in viscous shear, drops with partially mobile interfaces in viscous shear and bubbles in turbulent flow. Most of this work was based on liquid-liquid dispersions and only bears some resemblance to a gas-liquid
system. However, Calderbank (C2, 1958) showed that a single relationship cannot suffice for an adequate description of the behaviour of a gas both in electrolyte and non-electrolyte solutions.

The addition of a salt to a liquid, as well as suppressing coalescence, also influences oxygen diffusion coefficients (A1, J2, J3, H3 and R1) and measurements of oxygen diffusion coefficients in single electrolyte solutions were made by Ju and Ho (J3). It was found that for both single and mixed electrolyte solutions, oxygen diffusion coefficients are found to be well correlated by the square root of total ionic strengths, \( I \)

where:

\[
I = \frac{1}{2} \sum_i c_i z_i^2
\]

(Eqn. 3)

and can be expressed as:

\[
D = D_0 (1 - \sqrt{I})
\]

(Eqn. 4)

Where \( D_0 \) = oxygen diffusion coefficient in pure water and values of \( f' \) depend on the nature of the salts in the electrolyte solutions. For MgSO\(_4\), \( f' \) was measured to be 0.284 (moll\(^{-1}\))\(^{-1/6}\), showing that an increase in ionic strength led to a decrease in diffusivity. Similarly, Ho et al. (H2) observed that oxygen diffusion coefficient is dependent on the solution properties and species involved.

The effect of single and mixed electrolyte solutions on oxygen diffusion coefficient was measured and related to \( k_L \), as follows:

\[
k_L \propto D^{0.67}
\]

(Eqn. 5)

This corresponds to boundary layer theory predictions of oxygen transfer for rigid spheres. The addition of surfactants creates a rigid film surrounding the gas bubbles (M2). The bubbles then act as having rigid surfaces, greatly reducing the thinning process. The work by Ho et al. (H2) agrees with this, finding that the small gas bubbles
could be treated as rigid spheres for mass transfer purposes, with the relationship between $k_L$ and $D$ as shown in Equation 5. Similarly, Andrew (A2) reported that surface rigidity was found to have a marked effect on $k_L$, particularly as bubble size decreases. These findings (H2 and J3) combined show that increasing the ionic strength leads to a decrease in $k_L$. A similar result was seen by Zieminski and Whittemore (Z6), observing that an increase in either salt concentration or ionic strength tended to decrease the mass transfer coefficient, $k_L$.

In liquid-liquid dispersions (C1), coalescence efficiency is also affected by the type of drop. Applying similar theories to those used for a gas-liquid system, then, for deformable bubbles, whether the interface is fully mobile, partially mobile or immobile, the coalescence efficiency can be expressed as:

$$\eta(d_{pL}, d_{pL}) \propto \exp \left[ -K_1 e^b \right]$$  \hspace{1cm} (Eqn. 6)

where $b$ is a positive number (either 1 or $\%$) and $K_1$ is a function of liquid viscosity, density and bubble diameter.

It follows that as the energy dissipation rate, $e$ increases, then the coalescence efficiency decreases and vice versa. Coalescence efficiency is related to the proportion of collisions that result in coalescence, with an increased coalescence efficiency corresponding to a greater likelihood of bubbles coalescing.

For rigid bubbles,

$$\eta(d_{pL}, d_{pL}) \propto \exp \left[ -K_2 e^b \right]$$  \hspace{1cm} (Eqn. 7)

where $b$ is a positive number and $K_2$ is a function of liquid viscosity and density, bubble diameter and interfacial tension.
This implies that as energy dissipation rate increases, the coalescence efficiency increases.

It has also been found that liquid viscosity can affect the coalescence behaviour of gas bubbles. An increase in viscosity affects the thickness of the film and consequently the film thinning time as the two bubbles come together, hence affecting the coalescence rates.

Due to the increased mass transfer film thickness, viscosity has an inverse effect on $k_L$ (Eqn. 8). Also, as seen by Calderbank (C3), viscosity has an inverse effect on diffusivity and hence $k_L$. i.e.:

$$k_L = \mu_L^c$$

(Eqn. 8)

A similar relationship was seen by Andrew (A2).

This section has covered the general effects due to the addition of a salt or increase in liquid viscosity on bubble coalescence and related parameters. The following sections in this chapter discuss what has been researched in some more common gas-liquid devices on the effect of coalescence and how the changes in the liquid physical properties affect the key parameters.

### 2.2 GAS-LIQUID EJECTORS

It has been found from the literature survey that there has been considerable work carried out on pumping characteristics of gas-liquid ejectors or similar devices. More recently, mass transfer characteristics of these devices has been investigated, but mainly in air-water systems (Zhu (Z5), Schwartz et al. (S3) Cramers et al. (C5 and C6), Dirix and van der Wiele (D1) and Wong et al. (W1)).
This recent work carried out by various researchers on mass transfer in gas-liquid ejectors (or similar devices) showed similar trends. As reviewed by Zhu et al. (Z4, 1992), correlations of $k_{La}$ have been developed by various workers for ejector systems. All featured the influence of flow parameters but due to the use of different configurations and geometries, other, different parameters were included which made it difficult to make any good comparisons.

Despite differing geometries, similar trends relating gas and liquid flowrates to overall mass transfer were seen by Zhu (Z5), Schwartz et al. (S3), Cramers et al. (C6), and Wong et al. (W1). They all found that increasing both gas and liquid flowrates increased the mass transfer coefficient ($k_{La}$). Both Wong et al. (W1) and Schwartz et al. (S3) found that increasing the gas and liquid flowrates also increased the specific interfacial area in the device. However, Cramers et al. (C6) and Zhu (Z5) both discovered that above certain gas flowrates, increasing gas flowrate did not continue to give enhanced $k_{La}$: either $k_{La}$ would remain constant throughout (Z5) or $k_{La}$ would decrease for an increasing gas flowrate (C6). Both discussed that the change in the trend when increasing the gas flowrate was related to the change in flow pattern in the mixing tube. Increasing gas flowrate changed the flow pattern from bubble flow, through an intermediate condition to annular flow. All this work however was carried out in a coalescing system.

A similar results was reported by Dirix and van der Wiele (D1) in that the flow pattern is again shown to have an influence on the mass transfer: mass transfer is linearly dependent on $Q_{G}(Q_{G}+Q_{L})$ in the bubble flow regime and in the jet (or annular) flow regime, $k_{La}$ is independent of flowrates.

Practically no work has been done on the effects of viscosity on ejector characteristics. However, some investigative work on the effect of coalescence behaviour has been carried out in a 'slot injector' (Z7). This is of similar design and working principle to the gas-liquid ejector and Figure 2 shows a comparison. The slot injector was used to create the initial gas-liquid mixture which was then introduced to a vessel. Hence there
were some similarities to the ejector and also some dissimilarities in that the gas-liquid mixture exiting the slot injector was introduced into a vessel where the two phases were allowed to mix further and also with the surrounding liquid. This was where the mass transfer was measured.

An air-water system was used in Zlokarnik's work (Z7) with the coalescence behaviour changed by the addition of various amounts of Sodium Chloride, consequently varying the concentration up to 20 g l\(^{-1}\). The mass transfer coefficient was measured in the system under steady state conditions by continuous addition of hydrazine to remove absorbed oxygen. It was seen that the coalescence was inhibited slightly at low concentrations of NaCl. At slightly higher concentrations, the inhibition was very marked and for a further concentration increase the inhibition could not be substantially raised.

The effect seen was not thought to be attributable to the salt concentration alone and was principally connected to the device used to disperse the gas. Hence when looking at the gas-liquid ejector as a stand alone device, there may be a different behaviour seen when coalescence inhibition is present. This again shows the need for research on a device that can compare between different liquid phase physical properties using a consistent device and geometry set up.

In the work by Zlokarnik (Z7), it was found that mass transfer was enhanced by the addition of a salt which inhibited coalescence. It is important to remember however that the slot injector studied was not a stand alone device and following the formation of the gas bubbles within the injector, the bubbles were introduced into a vessel in a number of positions. Hence a direct comparison is not possible. Neither can predictions made on the effect of coalescence in ejectors be based on these findings. Hence again highlighting the need for the current research.
The following table summarises the areas of work discussed in this section:

**TABLE 1  Summary of literature reviewed on gas-liquid ejectors**

<table>
<thead>
<tr>
<th>Author(s)</th>
<th>Solutions used</th>
<th>Parameters studied</th>
<th>Observed effects</th>
</tr>
</thead>
<tbody>
<tr>
<td>Cramers et al. (C6)</td>
<td>air/nitrogen-water</td>
<td>$k_{L\alpha}$</td>
<td>$k_{L\alpha}$ increases then decreases as $Q_g$ increases.</td>
</tr>
<tr>
<td>Dirix and van der Wiele (D1)</td>
<td>nitrogen-water</td>
<td>Flow regimes $k_{R\alpha}$</td>
<td>$k_{L\alpha}$ is linearly dependent on $Q_g/(Q_g+Q_L)$ in the bubble flow regime.</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td>$k_{L\alpha}$ is independent of flowrates in the annular flow regime.</td>
</tr>
<tr>
<td>Schwartz et al. (S3)</td>
<td>air-water $CO_2$ in air into aqueous $K_2CO_3$ and $BiCO_3$ solution</td>
<td>$k_{L\alpha}$ $\alpha$</td>
<td>$k_{L\alpha}$ increases with increasing gas and liquid (constant gas) flowrates.</td>
</tr>
<tr>
<td>Wong et al. (W1)</td>
<td>dilute $CO_2$ - $Na_2CO_3$ and $NaHCO_3$ / $NaOH (aq)$</td>
<td>$k_{L\alpha}$ $\alpha$</td>
<td>$k_{L\alpha}$ increases with increasing gas and liquid velocities. $\alpha$ increases as gas and liquid velocities increase.</td>
</tr>
<tr>
<td>Zhu (Z5)</td>
<td>nitrogen-water</td>
<td>$k_{L\alpha}$</td>
<td>For constant $Q_g/Q_L$, increasing $Q_L$ leads to an increase in $k_{L\alpha}$. Increase $Q_g$ increases $k_{L\alpha}$ then effect flattens.</td>
</tr>
<tr>
<td>Zlokarnik (Z7)</td>
<td>Air-water $NaCl$ solutions</td>
<td>$k_{L\alpha}$</td>
<td>$k_{L\alpha}$ was enhanced when the $NaCl$ concentration is increased.</td>
</tr>
</tbody>
</table>
2.3 GAS-LIQUID STIRRED TANK REACTORS

Bubble coalescence effects in gas-liquid stirred tank reactors have received a lot of attention. In this work, a wide variety of salts and organic solutes were used to inhibit coalescence by researchers such as Calderbank (C2) and (C3), Reith and Beek (R2), Robinson and Wilke (R3), Machon et al. (M1), Van't Riet (V1), Keitel and Onken (K1), Ho et al. (H2), Linek et al. (L2) and (L3), Solanki et al. (S6), Barigou and Greaves (B1) and Martin et al. (M4).

Reith and Beek (R2, 1970) studied the effect of coalescence rates in a salt solution compared to water in a stirred tank, and found that if the gas phase was well mixed, the coalescence rates in water were found to be higher than in ionic solutions. Linek et al. (L2, 1988) found that a lot of work had been carried out on the effect of power and gas superficial velocity on $k_a^L$ in stirred tanks, but did not necessarily fit the dependence of $k_a^L$ on physical properties, such as the rate of coalescence of gas bubbles. From several experimental tests carried out using various electrolyte solutions (Na$_2$SO$_4$, KI, NaCl etc) and concentrations, the rate of coalescence for the aqueous electrolyte solutions was seen to be well described on the basis of the Gibbs theory of adsorption of the solute in the surface layer of the solution (L2). Lessard and Zieminiski (L1, 1971) tried to use a coalescence parameter that related to the viscosity of the experimental solution relative to the viscosity of water, the entropy of a solution, the self-diffusion parameter of water and the ionic strength of the solution. However none of these correlating parameters was found to be particularly successful.

The effect of electrolytes on bubble size and bubble size distribution has been studied by various researchers. Small changes in the physical properties of the solution, including the surface tension are not important enough to influence the bubble disruption forces in turbulent flow of a gas-liquid dispersion (B1). Thus the principal action of the impurities is to retard bubble coalescence, with the consequence that small bubbles, once they are formed, tend to preserve their individual identities. Barigou and Greaves (B1, 1992) carried out work in NaCl solution at a concentration that provided a liquid that
was coalescence suppressing rather than coalescence inhibiting. It was found that finer and more uniform distributions were produced in the salt solution with a reduction in Sauter mean diameter of 22 - 34% and increased holdup up to 70% leading to a large increase in interfacial area. As well as the salt solution having effects on $d_{50}$ in the bulk of the tank due to coalescence suppression, it was also noticed that the bubble size was also smaller near the impeller. The effect was not quite so significant but did imply that the presence of the salt in the water had an effect on bubble breakage and that it does not repress coalescence completely.

Bubble size distributions were also affected due to the presence of salt. The spread of the distributions were considerably reduced compared with those for water giving evidence of greatly hindered coalescence. The distributions were positively skewed for the ionic solution.

Solanki et al. (S4) simulated a non-coalescing medium using KBr solution and studied the effect on bubble size and bubble size distributions. Bubbles were formed in a tank from a filter cloth distributor. The addition of this electrolyte gave rise to a bimodal size distribution, particularly at larger gas flow rates. The difference in the effect of salt on bubble size distribution is probably due to the different method of introduction of gas into the vessel and the lack of dispersion of the gas in the tank.

Robinson and Wilke (R3, 1973) and Van't Riet (V1, 1979) previously found that in electrolyte solutions in stirred vessels, the average bubble size was significantly smaller than in a pure water system. This again was mainly due to the reduction in bubble coalescence frequency in stirred tanks and the decrease in bubble size, together with an increase in gas holdup, contributed to a significant increase in interfacial area. $k_a$ was observed to be directly proportional to bubble size, indicating that a decrease in bubble size reduced $k_a$. It was suggested (R4) that this relationship may be attributed to the change in bubble hydrodynamic regimes from deformable circulating bubbles of large diameter to rigid spherical bubbles of small diameter. Another possibility for this observation may be due to the effect of the salt solution on $k_a$ (V1). It is thought that
the value of $k_\text{l}$ is affected by the addition of ions - both directly by the presence of ions on the bubble surface and indirectly by changes in the bubble diameter. It was also found that raising the ion concentration in the STR increases $k_\text{l}$ considerably however in the ionic solution, $k_\text{l}$ is more dependent on power per unit volume than in a water system.

Ho et al. (H2) investigated and correlated diffusivity with $k_\text{l}$ (as also discussed in Section 2.1) and observed the effects of salt solutions on various parameters that have been previously discussed, such as $k_\text{l}$, holdup and $d_{32}$. It was seen that the gas holdup values did not vary for ionic strengths ranging from 0 - 0.1 mol\text{l}^{-1}. However, for ionic strengths above 0.1 mol\text{l}^{-1} up to 0.42 mol\text{l}^{-1}, gas holdup was seen to vary with increased strength and above ionic strengths of 0.42 mol\text{l}^{-1}, gas holdup levelled off. This effect also seemed to be independent of salt used. The bubble diameter was observed to decrease with increasing ionic strength, again levelling off at an ionic strength of 0.42 mol\text{l}^{-1}. As seen by other authors, bubble size is directly related to the decrease in bubble coalescence and increased ionic strength. Consequently the interfacial area increased with increasing ionic strength, plateauing at 0.42 mol\text{l}^{-1}. Marrucci and Nicodemo (M3) also saw that bubble diameter decreased with increasing electrolyte concentration until a minimum value was reached at which point bubble diameter was unaffected by further increases in electrolyte concentration.

Calderbank (C3, 1959) found that $k_\text{l}$ was independent of the power dissipated by the impeller, size of the bubbles, their number and mean rising velocity. $k_\text{l}$ was correlated with diffusivity for small (<2mm) and large (2 - 5mm) bubbles. The critical bubble diameter of 2mm differentiated between mobile surface (larger) bubbles and rigid surface (smaller) bubbles. The difference in the correlations for the mobile and rigid surface bubbles is probably due to the reasons previously discussed.

Similarly, Ho et al. (H2) observed that $k_\text{l}$ increased with increasing ionic concentration and was thought to be a function of the solution properties and the type of species involved rather than the power input per unit volume or superficial gas velocities.
Martin et al.'s (M4, 1995) investigations in a stirred tank for two types of impeller modified the coalescence behaviour by using solutions of both surfactants and an electrolyte. Various concentrations of Na₂SO₄ were used for coalescence inhibition and found to give a corresponding increase in gas holdup in the vessel when all other parameters were constant. This is a similar result to that found by Machon et al. (M1, 1977) in that overall gas holdup in stirred tanks was strongly influenced by the presence of dissolved inorganic salts but beyond a certain salt concentration, with all other conditions constant, the gas holdup remained constant. Also, as found by Van't Riet (V1, 1979), $k_L \alpha$ was much higher in the salt solution that inhibited coalescence than in water. This was explained by the holdup increasing and bubble size decreasing, enhancing specific interfacial area and hence $k_L \alpha$. It was suggested by Andrews (A2) that $k_L$ is largely unaffected hence holdup can be used as a guideline for mass transfer according to Martin et al. (M4). This however is not necessarily the case, since they (M4) measured $k_L \alpha$ and holdup and the corresponding increases for the addition of the salt solution are very similar. Bubble size is also affected by the presence of a salt and contributes to the enhancement of specific interfacial area. Hence, for the $k_L \alpha$ in the system to increase as seen, this must be due to a decrease in $k_L$ as well as the known increase in $\alpha$.

Keitel and Onken (K1, 1981) studied aqueous solutions of NaCl in various concentrations to inhibit coalescence in a stirred loop reactor. The effect was compared to pure water on gas holdup and $k_L \alpha$ as well as power. The density and interfacial tension of the salt solution were found to be about the same as for water. The presence of the electrolyte led to an increase in mass transfer up to a factor of 3, but had no effect on gas holdup up to 0.08 mol/l. Only at a higher concentration of 0.5 mol/l gave a significant increase in gas holdup.

Robinson and Wilke (R3, 1973) similarly observed that $k_L \alpha$ obtained for various non reactive electrolyte solutions are considerably greater than $k_L \alpha$ in water, the difference becoming larger with increasing ionic strength until an upper limiting value was reached.
It was found that the increased ionic strength in the electrolyte solutions led to a larger exponent on power per unit volume in relation to $k_Ia$.

It has also been observed (R3) that the increase in the amount of ionic solute decreased the bubble diameter and increased the gas holdup appreciably (also found by other researchers (M4)). It was attributed that the variation of $k_Ia$ with varying ionic strength was primarily due to variations in $a$. Holdup was seen to increase less than $a$ and it was thought that there was a secondary effect on $k_I$ due to variation in $k_Ia$ which came about by the decrease in bubble size.

It can be summarised that the addition of a salt to water at increasing concentrations in a stirred tank reactor has significant effects on various parameters. From the literature it is seen that there are similarities and discrepancies between researchers on exactly how the addition of salt affects the bubble coalescence and consequently bubble size, gas holdup and mass transfer.

However, this particular part of the literature survey is very valuable in providing some insight as to how a non-coalescing system differs from a coalescing system in a gas-liquid mixing device. Some of the effects seen will be able to relate to the ejector system and others will not.
The following tables summarise the effects on bubble size, gas holdup and $k_{l,a}$ determined by different researchers as discussed in this section:

**TABLE 2  Summary of literature reviewed on the effects on bubble size in stirred tank reactors**

<table>
<thead>
<tr>
<th>Author(s)</th>
<th>Solutions used</th>
<th>Effects observed</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ho et al. (H2)</td>
<td>Various concentrations of $NaCl$ solutions; $MgSO_4$, $KCl$ and $(NH_4)_2SO_4$ solutions</td>
<td>Bubble diameter decreased with increasing ionic strength. Bubble diameter decrease levelled off at ionic strength of 0.42 mol l$^{-1}$. Gas bubbles acted like rigid spheres in the electrolyte solutions.</td>
</tr>
<tr>
<td>Martin et al. (M4)</td>
<td>0.05, 0.1 and 0.2M $Na_2SO_4$</td>
<td>Bubble size reduced with addition of salts.</td>
</tr>
<tr>
<td>Reith and Beek (R2)</td>
<td>100 g l$^{-1}$ $Na_2SO_3$ in distilled water</td>
<td>Coalescence rates markedly higher in water than ionic solution.</td>
</tr>
<tr>
<td>Robinson and Wilke (R3)</td>
<td>Solutions of $KCl$, $KCl$ with KOH–$K_2CO_3$ and various combinations of $Na_2SO_4$ and $CuSO_4$</td>
<td>Visual observations indicated that upon addition of ionic solute, average bubble diameter decreased.</td>
</tr>
<tr>
<td>Van’t Riet (V1)</td>
<td>Water with ions</td>
<td>Addition of ions to water reduces bubble size.</td>
</tr>
</tbody>
</table>
**TABLE 3**  Summary of the literature reviewed on the effect on gas holdup in stirred tank reactors

<table>
<thead>
<tr>
<th>Author(s)</th>
<th>Solutions used</th>
<th>Effects observed</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ho et al. (H2)</td>
<td>Various concentrations of NaCl solutions; MgSO₄, KCl, (NH₄)₂SO₄ solutions</td>
<td>Gas holdup values did not vary from species to species for ionic strength ranging from 0 to 0.1 mol⁻¹, but did vary with increased ionic strength up to a value of 0.42 mol⁻¹ where the gas holdup began to level off.</td>
</tr>
<tr>
<td>Keitel and Onken (K1)</td>
<td>0.08 and 0.5M NaCl</td>
<td>No effect on gas holdup compared to air-water for 0.08M NaCl. 40% increase in gas holdup in 0.5M NaCl.</td>
</tr>
<tr>
<td>Machon et al. (M1)</td>
<td>5 kmolm⁻³ Na₂SO₄, 0.5, 1.5 and 2.5 kmolm⁻³ MgCl₂, 1, 3 and 5 kmolm⁻³ NaSCN</td>
<td>Overall holdup strongly influenced by the presence of dissolved inorganic salts, up to a factor of 2.</td>
</tr>
<tr>
<td>Martin et al. (M4)</td>
<td>0.05, 0.1 and 0.2M Na₂SO₄</td>
<td>Gas holdup increased with increasing coalescence inhibition (all other parameters constant).</td>
</tr>
<tr>
<td>Robinson and Wilke (R3)</td>
<td>Solutions of KCl, KCl with KOH-K₂CO₃ and various combinations of Na₂SO₄ and CuSO₄</td>
<td>Visual observations indicated that upon addition of ionic solute, dispersed fractional gas holdup increased appreciably.</td>
</tr>
<tr>
<td>Van’t Riet (V1)</td>
<td>Water with ions</td>
<td>Gas holdup becomes larger in the solution of ions in water, compared to pure water.</td>
</tr>
</tbody>
</table>
TABLE 4  Summary of the literature reviewed on the effects on mass transfer in stirred tank reactors

<table>
<thead>
<tr>
<th>Author(s)</th>
<th>Solutions used</th>
<th>Effects observed</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ho et al. (H2)</td>
<td>Various concentrations of NaCl solutions, MgSO$_4$, KCl, (NH$_4$)$_2$SO$_4$ solutions</td>
<td>$k_L \alpha$ increased with increasing ionic strength. $k_L$ correlated well with diffusivity.</td>
</tr>
<tr>
<td>Keitel and Onken (K1)</td>
<td>0.08 and 0.5M NaCl</td>
<td>Presence of the NaCl solutions increases $k_L \alpha$ up to a factor of 3.</td>
</tr>
<tr>
<td>Martin et al. (M4)</td>
<td>0.05, 0.1 and 0.2M Na$_2$SO$_4$</td>
<td>$k_L \alpha$ in salt solution much higher compared to water. Concentration of the salt solution has little difference.</td>
</tr>
<tr>
<td>Robinson and Wilke (R3)</td>
<td>Solutions of KCl, KCl with KOH-K$_2$CO$_3$ and various combinations of Na$_2$SO$_4$ and CuSO$_4$</td>
<td>$k_L \alpha$ values found to be significantly dependent on total ionic strength. For all solutions, $k_L \alpha$ correlated well with $(P/V_l)^n$, $n$ being dependent on the nature of the solution. Values of $k_L \alpha$ obtained for various electrolyte solutions are considerably greater than $k_L \alpha$ values obtained in water.</td>
</tr>
<tr>
<td>Van’t Riet (V1)</td>
<td>Water with ions</td>
<td>Raising the ion concentration in solution increases $k_L \alpha$ considerably. $k_L \alpha$ values for ionic solutions are more dependent on power per unit volume than those for pure water.</td>
</tr>
</tbody>
</table>
2.4 BUBBLE COLUMNS

As with stirred tank reactors, many researchers have investigated the effects of liquid physical properties on bubble column reactor performance. This section will summarise some of the work that has been carried out in this field.

Heijnen and Van't Riet in 1984 (H1), reviewed the work carried out in bubble columns. As with a STR, it was found that bubble sizes were larger in coalescing media than in non-coalescing media and $k_La$ was higher in non-coalescing fluids. An important conclusion for bubble columns was that for $d_{32} > 2\text{mm}$, $k_L = (3 - 4) \times 10^{-4}\text{ms}^{-1}$ while for smaller bubble diameters, $k_L$ values could be significantly lower, varying with bubble rigidity. It was said that for $d_{32} < 0.8\text{mm}$, $k_L$ was probably about $1 \times 10^{-4}\text{ms}^{-1}$.

Keitel and Onken (K2, 1982) also studied coalescence effects in bubble columns with respect to bubble size and gas holdup by the use of various salts and alcohols. As with a stirred tank reactors, the addition of a solute reduces the bubble size or Sauter mean diameter. In salt solutions, increasing the salt concentration decreases the bubble diameter until there is a levelling off at a small bubble size for ionic strengths greater than 1 mol/l. However, it was found that gas holdup was practically unaffected by electrolyte concentration in a bubble column.

This was not case found by Zahradnik (Z1, 1995) et al. who carried out work in a bubble column reactor using various salt solutions who found that the gas holdup increased with increasing electrolyte concentration but was independent of type of electrolyte. As well as carrying out holdup measurements for nine different electrolyte solutions, the bubble coalescence frequency was determined in a coalescence cell, which led to the evaluation of the transition concentration corresponding to the switch from complete coalescence to virtually total coalescence suppression. These results were similar to those discussed in Section 2.1.
From their observations they (Z1) concluded that bubble coalescence was almost completely suppressed as concentrations approached the respective transition concentrations for individual electrolytes, and subsequently the hydrodynamics of such non-coalescing bubble beds is not further altered by an increased solute concentration.

For all electrolytes studied, the appropriate limiting dependencies of holdup against superficial gas velocity corresponding to the respective bubbling regimes were achieved at concentrations close to the transition concentrations of respective electrolytes. Further increase of electrolyte concentration had a negligible effect on the bubble bed voidage. The characteristics of the holdup in the beds were also strongly influenced by the presence of the electrolytes, which differs from the conclusion reached by Keitel and Onken (K2).

Prince and Blanch (P4, 1990) carried out research in an air-sparged bubble column in order to test the validity of a phenomenological model for the rates of bubble coalescence and bubble break up in turbulent gas-liquid dispersions. It was shown that inorganic electrolytes significantly decrease the degree of bubble coalescence in turbulent gas-liquid dispersions due to contact times in the order of milliseconds compared to film thinning times for immobile interface bubbles in the order of seconds.

Two salts (NaCl and Na$_2$SO$_4$) at various concentrations were compared and their effect of concentration on coalescence inhibition. Na$_2$SO$_4$ was found to be more effective at reducing coalescence at low concentrations compared to NaCl. This was thought to be due to the lower diffusivity of Na$_2$SO$_4$ and therefore its lessened ability to relieve surface tension gradients during the film thinning process.

However the model for bubble coalescence rates was inadequate, particularly at higher salt concentrations, despite the transition concentration of salts in stagnant liquids being well correlated. Hence this model could not be adapted and applied to a gas-liquid ejector system.
Gas holdup and bubble size together with bubble rise velocity was also studied in various systems by Jamialahmadi and Müller-Steinhagen (Jl). They looked at the effect of various alcohols and organic acids as well as various concentrations of KCl. Experimental tests were performed with KCl solutions where the surface tension at low concentrations was similar to that of water. For large gas flowrates, the bubbles created at the gas distributor plate broke up into a large number of small bubbles. It was found that there were considerable effects on bubble size.

As found by Zahradnik et al. (Zl), the presence of an electrolyte may change the holdup from that in pure water, not only in absolute terms, but also showing a different trend. For higher gas superficial velocities, an increase in the concentration of an electrolyte increased the gas holdup.

Zieminski and Whittemore (Z6, 1971) also investigated the effect of various salt solutions and concentrations on the interfacial area (calculated from bubble size and gas holdup) and oxygen transfer coefficient in a bubble column. Again, the salt affected the bubble size compared to that in pure water, with much reduced bubble size and an increased number of bubbles present. An increase in salt concentration increased the surface area; the trend appeared to be affected by the valency of the salt. The 3-2, 3-1 and 1-3 and 2-2 electrolytes were most effective while the 1-1 electrolytes showed the smallest effect. Magnesium Sulphate has an intermediate to large effect on the interfacial area. Their (Z6) results of surface area however were found to correlate well with ionic strength. In this work, it was observed that for the range of bubble sizes measured, the holdups were almost the same.

Again, as with stirred tank reactor, the addition of a salt to water at increasing concentrations in a bubble column has significant effects on various parameters. Some of these effects may be relevant to the ejector system.
The following table summarises the literature reviewed in this section:

TABLE 5  Summary of literature reviewed on bubble columns

<table>
<thead>
<tr>
<th>Author(s)</th>
<th>Solutions used</th>
<th>Parameters studied</th>
<th>Observed effects</th>
</tr>
</thead>
<tbody>
<tr>
<td>Jamialahmadi and Müller-Steinhagen (J1)</td>
<td>5 - 150 g\textsuperscript{1} \text{KCl} solutions</td>
<td>Bubble size</td>
<td>Gas holdup increases for low concentrations of \text{KCl} solution compared with water.</td>
</tr>
<tr>
<td></td>
<td>Various alcohols and organic acids</td>
<td>Gas holdup</td>
<td>High concentration of \text{KCl} gives an increase in interfacial tension leading to an increase in bubble size and decrease in gas holdup.</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Bubble rise velocity</td>
<td>At high superficial gas velocities, unstable bubbles break-up which leads to an increase in holdup.</td>
</tr>
<tr>
<td>Keitel and Onken (K2)</td>
<td>Solutions of \text{NaCl}, \text{Na}_2\text{SO}_4, \text{Al}_2(\text{SO}_4)_3 and \text{NaOH} and organic compounds</td>
<td>Bubble size</td>
<td>Addition of solute reduces bubble size.</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Gas holdup</td>
<td>In salt solutions, increase in salt concentration decreases bubble diameter then levels off for ionic strengths &gt; 1 mol\textsuperscript{1}.</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td>Gas holdup was practically unaffected by electrolyte concentration.</td>
</tr>
<tr>
<td>Source</td>
<td>Electrolyte Solutions</td>
<td>Phenomenon</td>
<td>Comments</td>
</tr>
<tr>
<td>--------</td>
<td>------------------------</td>
<td>------------</td>
<td>----------</td>
</tr>
<tr>
<td>Prince and Blanch (P4)</td>
<td>Solutions of up to 0.1M NaCl and up to 0.04M Na&lt;sub&gt;2&lt;/sub&gt;SO&lt;sub&gt;4&lt;/sub&gt;</td>
<td>Bubble size</td>
<td>Changes in bubble size may result from the influence of salts on either bubble break-up or bubble coalescence rates. Inorganic electrolytes significantly decrease degree of bubble coalescence in turbulent gas-liquid dispersions.</td>
</tr>
<tr>
<td>Zahradnik et al. (Z1)</td>
<td>Solutions of MgSO&lt;sub&gt;4&lt;/sub&gt;, Na&lt;sub&gt;2&lt;/sub&gt;SO&lt;sub&gt;4&lt;/sub&gt;, NaOH, NaCl, KCl, KI, CaCl&lt;sub&gt;2&lt;/sub&gt;, BaCl&lt;sub&gt;2&lt;/sub&gt;</td>
<td>Gas holdup</td>
<td>Gas holdup increased with increasing electrolyte concentration.</td>
</tr>
<tr>
<td>Zieminski and Whittemore (Z6)</td>
<td>Various electrolyte solutions including: Na&lt;sub&gt;2&lt;/sub&gt;SO&lt;sub&gt;4&lt;/sub&gt;, MgSO&lt;sub&gt;4&lt;/sub&gt;, NaOH, NaCl, MgCl&lt;sub&gt;2&lt;/sub&gt;, Al&lt;sub&gt;2&lt;/sub&gt;(SO&lt;sub&gt;4&lt;/sub&gt;)&lt;sub&gt;3&lt;/sub&gt;</td>
<td>Bubble size</td>
<td>The degree of coalescence of air bubbles in water decreases on addition of inorganic salts. The decrease was found to be dependent on the valence of the respective ions. Presence of salt reduced bubble size. Gas holdup almost unaffected.</td>
</tr>
</tbody>
</table>
Chapter 3 - EXPERIMENTAL METHODOLOGY

3.1 EXPERIMENTAL SET UP

Figures 3 and 4 show a schematic diagram of the downflow experimental set up used in this work. The supply tank was equipped with an immersion electric heater used to control the liquid temperature. Liquid was pumped from a storage tank, through a 40mm pipe, to the top of a 3 metre vertical section where the ejector was installed. Gas was supplied either from nitrogen cylinders or from a locally situated air compressor, through a 25mm pipe into the ejector. The two phase mixture from the ejector passed through a section of pipeline and then to a vessel where the two phases separated: the gas to atmosphere and the liquid to the receiving tank.

Valves were installed to control and adjust both liquid and gas flowrates. The back pressure of the ejector was controlled using a valve situated downstream of the pipeline, before the separation vessel. The experiments were carried out by maintaining a constant downstream pressure, adjusting the liquid flowrate to a fixed value and varying the gas upstream pressure, which in turn affected the liquid flowrate and downstream pressure. Consequently, a considerable amount of time was spent in establishing a particular set of flow conditions.

In Figure 4, the positions of various measuring probes are labelled. The ejector included the use of a helical spinner (a specific geometry, previously found to be the optimum design (Z2)) upstream of the nozzle and a straight mixing tube. The spinner imparts a tangential velocity to the liquid jet, creating a swirl effect. A diffusor was not used in this study in order to keep the geometry simple and to be consistent with work previously carried out by Zhu (Z2 and Z3) to enable comparison.
The dimensions of the ejector are:

Mixing tube diameter, \( D_m = 0.029 \text{m} \)

Nozzle diameter/mixing tube diameter, \( D_n/D_m = 0.41 \)

Mixing length/mixing tube diameter, \( L_n/D_m = 24 \)

Swirl number of the spinner, \( Sw = 0.18 \) (Swirl number is described in Appendix A)

Temperature was controlled at \( 20 \pm 2^\circ\text{C} \) and the back pressure, \( p_2 \) at \( 4 \pm 0.5 \text{ psig} \).

### 3.2 OPERATING FLUIDS

#### 3.2.1 Coalescence effects

The choice of liquid for these experiments was based on the fact that it should have a non-coalescent behaviour consequently differing greatly from that of pure water.

As seen in Figure 1 (L1) and discussed in Chapter 2, as a salt is added to water, a concentration is reached where the coalescence is drastically reduced. This sharp transition is a function of the valency combinations of the salts (Z6). Magnesium Sulphate solutions have a very sharp transition from 100% coalescence to very low coalescence frequencies over a narrow concentration range. This occurred at a low molar concentration, and also gave a minimum disturbance to the system properties.

A molar concentration of 0.08M Magnesium Sulphate solution was used in order to ensure that the operating region was well within the low coalescence rate range. The density of this solution was measured to be 1004 kgm\(^{-3}\) and the surface tension measured to be 0.07 Nm\(^{-1}\) (cf. 998.2 kgm\(^{-3}\) at 20\(^\circ\text{C}\) and 0.072 Nm\(^{-1}\) respectively for pure water).

#### 3.2.2 Viscosity effects

In order to investigate the effects of liquid viscosity on the gas-liquid ejector performance, a Newtonian fluid was required with a density and interfacial tension
similar to water. Hence, it was known that any observations were due to viscosity alone and not a combination of various liquid physical properties.

A number of liquids were considered to increase the liquid viscosity that would not affect the system properties significantly and maintain a Newtonian fluid. The liquid chosen for these experiments was a Newtonian CMC solution (7ELC1), which at concentrations of 1, 2.5 and 3.8 % w/w gave liquid viscosities of 3, 6 and 10 mPas respectively, all of which enabled the flow to remain turbulent.

The liquid viscosity was increased to 10 mPas initially. Due to the significant effects seen on $k_o$, the decision was made to reduce rather than increase the viscosity for the remaining experiments.

The sodium CMC has an average molecular weight of 21,000 gmol$^{-1}$. Hence the molarity of CMC in these solutions were in the order of 0.001 mol$^{-1}$. Assuming complete dissociation of Na$^+$ ions in water, the ionic strength at these molarities is still far below the level which changes the fluid system from coalescing to non-coalescing. It was assumed that the presence of the Na$^+$ ions alone in these CMC solutions did not have any effect on the coalescence behaviour. It is assumed in this work that the size of the negative ions is unimportant.

The density and surface tension of these solutions were very similar and measured at 1017 kgm$^{-3}$ and 0.070 Nm$^{-1}$ respectively.

3.3 PERFORMANCE PARAMETERS

Performance of the ejector system was assessed in terms of liquid power input per unit volume, gas entrainment, ejector pumping efficiency and mass transfer. The parameters measured to determine this were: upstream pressure ($p_1$), temperature ($T_1$) and dissolved oxygen concentration ($C_1$), gas chamber pressure ($p'_s$); downstream pressure ($p_2$),
temperature ($T_2$) and dissolved oxygen concentration ($C_2$) and liquid and gas flowrates ($Q_L$ and $Q_G$).

These parameters were calculated as follows:

Liquid power input per unit volume $= P_f/V_d$ where:

\[
P_T = (p_z' - p_2)Q_L + \frac{1}{2} \rho Q_L (u_f^2 - u_{2L}^2) \tag{Eqn. 9}
\]

and

\[
V_d = 0.785 D_m^2 L_m \tag{Eqn. 10}
\]

Gas entrainment is dependent on $p_z'$ and related to mixing tube pressure drop, $\Delta p_m$:

\[
\Delta p_m = (p_z' - p_2) \tag{Eqn. 11}
\]

where:

\[
p_z' = p_z + \rho g L_m (1 - \phi) \tag{Eqn. 12}
\]

Ejector pumping efficiency:

\[
\eta = \frac{P_G}{P_T} \tag{Eqn. 13}
\]

where $P_G$ is the power gained by the gas, as detailed in Appendix B.
Overall mass transfer coefficient:

\[ k_{L,a} = \frac{Q_L (C_1-C_2)}{\Delta CV_d} \]  
(Eqn. 14)

Pressures were measured using pressure transducers and temperatures with platinum probes. Miniature dissolved oxygen probes were used to determine oxygen concentration in the liquid phase. The flow pattern in the mixing tube was determined with a fibre optic probe which measures the fraction of time the probe spends in each phase. Liquid flowrate was measured with an electromagnetic flowmeter and gas flowrate measured with turbine flowmeters.

The pressures, temperatures and flowrates were recorded for all experimental tests. The additional parameter, dissolved oxygen concentration, was measured in the specific set of tests designed to measure mass transfer performance in the ejector system. For these tests, the tanks were filled and the selected fluid recirculated around the whole test rig with air as the process gas. The idea of to completely saturate the fluid with air and this, depending on the liquid phase could take up to an hour.

The dissolved oxygen probes, having been calibrated during saturation of the liquid phase were then fixed in appropriate positions along the test section (Figure 4). The valve between the supply and receiver tank was closed and all the liquid was pumped to the supply tank for batch operation. Each mass transfer test was then carried out using nitrogen supplied from pressurised cylinders, as the process gas in order to strip the oxygen from the liquid phase, resulting in a decrease in dissolved oxygen concentration. In conjunction with the pressure and flowrate measurements, the mass transfer coefficient was determined.

Calculation of all the parameters mentioned in this section are detailed in Appendix B.
3.4 BUBBLE SIZE MEASUREMENT TECHNIQUES

A photographic technique was used to determine bubble size. For each set of flow conditions, numerous photographs of the bubbles were taken. An SLR camera was used with an extension tube using Kodak TMAX 400 ASA professional black and white film. A high-speed flash unit having a duration of 1μs, was used, which gave sufficiently short time to freeze the bubbles motion under all flow conditions tested. The flash was positioned opposite the camera in order to produce back lighting. This position was found to give clearer images of bubbles than side lighting. The photographs of gas bubbles were taken through a square window to eliminate errors due to the curvature of the pipe. The photographic set up is illustrated in Figure 5.

3.4.1 Calibration

Bubble size calibration was achieved by using a graticule with a 5mm scale positioned at depths of 2mm, 6mm and 10mm in the mixing tube as shown in Figure 6. The calibration took place in water, assuming that the set up would be the same for the separate fluids. A calibration block was made up to aid accurate positioning of the graticule. The graticule was first fixed into the block which was then placed in the mixing tube and the position of the camera lens was adjusted until the graticule was in focus. This distance between the lens and mixing tube was recorded for each depth of the graticule from the pipe wall. A photograph of the graticule was taken to be used as the scale for calibration.

3.4.2 Bubble Photographs

The majority of photographs were taken with the focal plane at a fixed radial position of 2mm from the pipe wall. Limited photographs were taken at larger radial positions in the air-water system to establish if the value of Sauter mean diameter, \( d_{32} \) at 2mm from the pipe wall can be taken as a good approximation which averages over the whole cross section of the pipe.
To take photographs of bubbles at the three given axial positions in the mixing tube, as in Figure 7, the camera was set up on the rig and positioned so that the distance between the lens and mixing tube was the same as that determined during calibration. For every set of flow conditions at least 20 photographs were taken at intervals of about 10 seconds. The films were then developed to produce negatives for analysis.

3.4.3 Analysis of Photographs

The photographs of bubbles were analysed semi-automatically using a Joyce Loeb Magiscan 2A Image Analyzer system (Y1). The first step in the measurement was to establish the required magnification of the image analyzer and the calibration factor at this magnification. This was done by using a negative of the graticule taken as described in Section 3.4.1. The magnification was set in such a way that the image of the whole negative captured by the camera just filled the screen, hence eliminating the possibility of duplicate counts. Once the calibration factor was established, negatives of the air bubbles were analysed. The size of a bubble was measured with a light pen by marking 3 points on the circumference of the bubble. The analyzer then automatically drew a circle passing through these three points. The diameter of the circle was the diameter of the bubble, assuming the bubble was spherical. For non-spherical bubbles, three points were marked in such a way that the area and diameter of the circle was a very good representation for the bubble. Due to the fact that the bubble measurements are very subjective, particularly for non-spherical bubbles, repeats of photographs of selected conditions were carried out both by the same observer and an independent observer. These repeats were found to be within 5% of each other. Given the difficulty and scatter in measuring bubble sizes, the repeatability of 5% is encouraging and presents users with confidence (Y2).

All the raw data was transferred to a PC and the data analysis carried out using a spreadsheet package.
3.5 Holdup Measurement Techniques

There are two techniques which can be used to measure gas holdup in a pipe: residence time distribution (RTD) and gamma ray densitometer. Figure 8 represents the experimental set up used for measuring gas holdup using the RTD technique in an air-water system.

Previous work has been carried out on single phase mixing and residence time distributions in a six element Kenics mixer (S4). The technique used was found to be successful and was applied to the gas-liquid ejector system to measure the liquid residence time and hence calculated gas holdup. The technique used a small amount of saturated NaCl solution that was injected as a pulse into the liquid phase directly upstream of the spinner. The salt concentrations in the single and multiphase streams were monitored using conductivity probes. These probes were connected to a conductivity box and then to a DASH 16 DA board into a computer. A conductivity data acquisition and analysis programme was set up in order to acquire the data. For each channel, 2500 samples were taken per second. A typical output of this graph can be seen in Figure 9. Trace (1) is from conductivity probe 1 for liquid only, and trace (2) is from conductivity probe 2. The second trace appears to be noisier but this is due to the presence of gas.

The programme automatically calculated the time gap between the two peaks on the trace curves as well as the time gap between the 50th percentiles. Both values were recorded, but the peak to peak time was more accurate due to the peaks being more distinguishable, particularly on the two phase trace, and hence used in subsequent calculations.
At every liquid flowrate, measurements were taken in the absence of gas as a basis and gas holdup was calculated using the following equation:

\[
\phi = 1 - \left( \frac{t_{g,1} - \frac{V_{Li}}{Q_L}}{t_{L,1} - \frac{V_{Li}}{Q_L}} \right)
\]

(Eqn. 15)

where

- \( Q_L \) = liquid flowrate
- \( t_{g,1} \) = peak-peak time between conductivity probes 1 & 2, gas-liquid mixture
- \( t_{L,1} \) = peak-peak time between conductivity probes 1 & 2, liquid only
- \( V_{Li} \) = volume of liquid between conductivity probe 1 and the tip of the ejector nozzle

This equation is derived as shown in Appendix C.

For each set of flow rates, three repeats were conducted to obtain an average value of peak to peak time. Previous work (V2) on gas-liquid dispersion motionless mixers has studied the repeatability of the peak to peak measurement and concluded that three repeats gave a sufficient degree of accuracy. Typically, the variations in peak to peak time between different repeats were less than 10%. Table 6 shows a selection of peak to peak times at one set of flow conditions measured and the respective calculated gas holdup.

For experiments where the bulk liquid was a salt solution, representing the non-coalescing fluid, a further salt injection into the system at such small quantities would have little effect on the conductivity probes. Also in the viscous system, the RTD technique was tried, but gave inaccurate results because of excessive noise in the system. This was probably a result of the sodium ions present in CMC and hence further salt...
injection into the system at such small quantities having little effect on the conductivity probes. An alternative technique was available using a gamma ray densitometer.

A set of measurements were carried out in an air-water system to compare the RTD technique and enhance confidence in the results when used for the salt solution and viscous solutions. Since the RTD technique measured overall gas holdup in the mixing tube whereas the gamma ray densitometer technique could only measure the local gas holdup, this comparison was necessary to confirm that the average of the limited local holdup measurements were representative of the overall gas holdup in the mixing tube.

The local gas holdup was measured at the same 3 axial positions as for bubble size measurement (see Figure 7) and an overall gas holdup was taken as the average of these 3 measurements. Figure 10 shows a diagram of the experimental set up. The radiation source was a line source of Americium 241 incorporated in ceramic beads sealed in a stainless steel capsule. This line source was 3 mm in diameter and 30 mm long. The detector system was a photomultiplier and scintillation crystal connected to a remote scale ratemeter where readings were taken.

Each time the gamma source/detector was re-positioned, and every hour during testing, calibration of the system was carried out to ensure that any drift was not significant enough to affect the results. This calibration was done at 2 points, one when the pipe was empty and the other when the pipe was full of the test liquid.

Measurements were then taken for the range of flow conditions tested. For each measurement, including calibration, a counting time of 60 seconds was set. Following this counting period, the scale ratemeter automatically averaged out the count rate and displayed this average value.
This reading, combined with the calibration readings, was used to calculate local gas holdup in the mixing tube using the following equation:

\[ \phi = \frac{\ln \left( \frac{R}{R_d} \right)}{\ln \left( \frac{R_l}{R_d} \right)} \]  

(Eqn. 16)

where \( R \) = reading for certain flow condition
\( R_d \) = reading when pipe section was empty
\( R_l \) = reading when pipe section was filled with liquid

3.6 CALCULATION OF \( k_L \) AND \( a \)

The results from bubble size and holdup measurements can be used to calculate the specific interfacial area \( a \). Assuming spherical bubbles, the following equation applied:

\[ a = \frac{6\phi}{d_{32}} \]  

(Eqn. 17)

where \( d_{32} \) is the Sauter mean diameter of the bubbles, along the whole mixing tube. With the knowledge of both \( a \) and \( k_La \), \( k_L \) can also be calculated.
Chapter 4 - AIR-WATER BUBBLE SIZE AND GAS HOLDUP - RESULTS AND DISCUSSION

The first approach in extending the work carried out by Zhu (Z2 and Z3) was to establish the effect of the flow parameters on bubble size and gas holdup in air-water. These parameters can be combined to estimate \( \alpha \), the specific interfacial area so that \( k_t \) can be calculated. This chapter discusses the experimental programme undertaken to achieve these results.

4.1 BUBBLE SIZE MEASUREMENTS

4.1.1 Minimum Bubble Count

Photographs of air bubbles in water were taken at different axial and radial positions in the mixing tube and at various liquid and gas flowrates. From the initial photographs developed, the negatives were analysed to determine the minimum bubble count required to give sufficiently accurate values of mean bubble size. This was an important parameter to determine for use in future experimental tests. In this study, the Sauter mean diameter, \( d_{32} \), is used as the mean bubble size because this is the parameter which is related to specific interfacial area (as in Equation 17).

The value of Sauter mean diameter \( (d_{32}) \) can be calculated as follows:

\[
\frac{d_{32}}{} = \frac{\Sigma d_i^3}{\Sigma d_i^2} \tag{Eqn. 18}
\]

where \( i = 1 \rightarrow n \)

\( n \) = the number of bubbles counted for the specific flow condition and position

Figure 11 shows examples of \( d_{32} \) against bubble count. This graph represents the value of \( d_{32} \) against the number of bubbles counted for one liquid flowrate and three different gas flowrates and is representative of the whole range of flow conditions tested. As can
be seen, the more bubbles counted, $d_{32}$ value becomes more stable. It is considered that an accuracy of ±5% in bubble size measurement is sufficient for the present purpose, therefore the minimum bubble count required, as seen in Figure 11 is 200 - 250.

4.1.2 Bubble Size Distribution (BSD)

For each set of flow conditions, a histogram was produced to observe the bubble size distribution, which also shows the number of bubbles in a particular size range. A graph showing cumulative number frequency against bubble size was also created at the flow conditions tested. This gave an idea of the actual percentage of bubbles that was below a particular size. Examples of each of these graphs at specific sets of flow conditions and positions are shown in Figures 12 and 13 respectively. The effect of gas flowrate can also be seen in these graphs. The bubble size distribution is close to being a normal distribution (P2 and V3). For all the conditions where bubble size measurements were taken, the distributions were observed to be close to normal.

4.1.3 Effects of Positions and Flow Parameters on $d_{32}$

Measurements were carried out at various liquid and gas flowrates and axial positions in the mixing tube. The flow regime in which these measurements were taken was bubble flow, verging on intermediate flow for one condition. Most photographs were taken at a radial position 2 mm from the pipe wall. For two flow conditions (different liquid flowrates at constant gas-liquid flow ratio), photographs were also taken at radial positions of 6mm and 10mm from the pipe wall.

The results at a radial position of 2mm are included in Tables 7 - 9.

**Effect of axial and radial positions**

For each flow condition, local $d_{32}$ for various positions was calculated as well as an overall $d_{32}$ where possible. The number of bubbles counted and value of standard deviation ($\sigma$) are also included in these tables. $\sigma$ is calculated using

$$\sigma = \sqrt{\frac{\sum (d_i - d_{32})^2}{n - 1}}$$  \hspace{1cm} (Eqn. 19)
As a guidance of what a particular value of $\sigma$ means, the following approximations apply assuming a normal distribution (K3):

About 66.7% of the values will be within $d_{32} \pm \sigma$.
About 95% of the values will be within $d_{32} \pm 2\sigma$.
About 99.75% of the values will be within $d_{32} \pm 3\sigma$.

Table 7 shows that at a given flow condition, local $d_{32}$ significantly increased as the flow progressed down the mixing tube especially from position 1 to position 2. This result indicates the bubble coalescence in the mixing tube. This effect was also seen by visual observation of the mixing tube, particularly at the lower gas flowrates.

For one flow condition ($Q_L = 2.1 \, \text{ls}^{-1}$, $Q_G = 1.5 \, \text{ls}^{-1}$), photographs were also taken at radial positions of 6mm and 10mm from the pipe wall at the bottom of the mixing tube (axial position 3). At a depth of 10mm into the pipe, it was not possible to capture bubbles on film. At a depth of 6mm into the pipe, the majority of the photographs were blurred with some bubbles clear and in focus. The number of bubbles able to be analysed in numerous photographs was 99, giving a value of $d_{32}$ about 5% different from the value at 2mm from the pipe wall. Despite the number of bubbles counted being less than the minimum bubble count, the value of $d_{32}$ had appeared to stable out. From this result, it is assumed that the value of $d_{32}$ taken at 2mm from the pipe wall can be taken as a good approximation for the whole cross section of the pipe.

**Effect of gas flowrate**

The effect of gas flowrate on $d_{32}$ was investigated over a range of gas-liquid flow ratios from 0.48 to 0.95 under liquid flowrates of 1.2 and 2.1 $\text{ls}^{-1}$. As shown in Table 8, $d_{32}$ decreased with an increase of gas flowrate. An increase in gas flowrate was seen to increase the liquid power input per unit mass at constant back pressure. This increase in specific energy input into the system implies that break-up of the gas bubbles is more likely. However, from the literature (B2), for static mixers:

$$d_b = \left( \frac{P}{M} \right)^{-0.4} (1 + 1.7\phi) \quad \text{(Eqn. 20)}$$
By using the measured values of average gas holdup and power per unit mass in this equation, an increase in gas flowrate indicated a small increase in bubble size. However, by using different gas holdup values by taking into consideration the experimental errors this trend is not necessarily consistent. This shows that bubble size prediction (using Equation 20) is very sensitive to the error incurred due to gas holdup. Additionally, the ejector system is more complex due to the interaction of break-up and coalescence, hence suggesting that the above correlation for bubble size prediction may not be applicable in this case.

**Effect of liquid flowrate**

Table 9 shows the effect of liquid flowrate on $d_{50}$ for different gas-liquid flow ratios and positions. At the bottom of the mixing tube (position 3) Sauter mean diameter decreased with increasing liquid flowrate. This effect was also seen half way down the mixing tube (position 2) at a gas-liquid flow ratio of 0.48. An increase in liquid flowrate also increases the liquid power input per unit mass. Again, this increase in specific energy input into the system implies increased break-up of the gas bubbles. Using measured values of gas holdup and power per unit mass (as in Equation 20), an increase in liquid flowrate at constant $Q_g/Q_L$ indicated a decrease in bubble size. This is consistent when taking into consideration the error in holdup. Hence there is reasonable agreement between the correlation as found in previous literature and the present experimental data with respect to the effect of liquid flowrate on bubble size.
Bubble size prediction

From the results of bubble sizes in an air-water system, as summarised in this section, a correlation has been developed for estimating the Sauter mean diameter at specific points in the mixing tube as follows:

\[
d_{32} = 10.62 \left( \frac{Q_G}{Q_G + Q_L} \right)^{1.5} \left( \frac{P_T}{V_d} \right)^{0.34} \left( \frac{L}{L_m} \right)^{0.67}
\]  

(Eqn. 21)  

\[R^2 = 0.967\]

for 1.2 \( \text{ls}^{-1} \leq Q_L \leq 2.1 \text{ls}^{-1}\) and \(Q_G/Q_L < 1.0\). \(d_{32}\) is in mm.

The quality of this correlation can be seen in Figure 14.

From this correlation, an overall Sauter mean diameter can be estimated in an ejector mixing tube of similar geometry in an air-water or coalescing system.

4.2 GAS HOLDUP MEASUREMENTS

Figure 15 shows the effect of gas flowrate on gas holdup for a range of liquid flowrates. As one would expect, an increase in gas flowrate increased the gas holdup in the mixing tube.

For the range of liquid flowrates investigated, at the lower gas flowrates where bubble flow was exhibited, an increase in gas flowrate had a greater effect on gas holdup. At relatively higher gas flowrates where the flow pattern reached annular flow, the effect of gas flowrate became less significant.

The data in Figure 15 were re-plotted by using gas-liquid flow ratio on the x-axis. As can be seen in Figure 16, for the whole range of liquid flowrates at a gas-liquid flow ratio of around 1.0, gas holdup appears to stabilise and with gas-liquid flow ratios above 1.5 gas holdup was constant and independent of both liquid and gas flowrate. This figure also shows that the data can be brought together well by plotting holdup against \(Q_G/Q_L\) and is not dependent on liquid flowrate alone.
The gas holdup was also measured using the gamma ray densitometer technique. Figure 17 shows the close agreement between the RTD and gamma ray densitometer techniques used in air-water.

4.2.1 Slip-velocity calculation

Figure 18 shows a comparison of the measured gas holdup and \( \frac{Q_g}{Q_g + Q_L} \). The graph shows that measured gas holdup in the mixing tube is consistently higher than \( \frac{Q_g}{Q_g + Q_L} \) generally by 20 - 30%. A positive difference between gas holdup and \( \frac{Q_g}{Q_g + Q_L} \) would be expected because of a difference in velocities between the gas and liquid phase, known as slip-velocity. The vertical downflow orientation gives a positive slip-velocity (i.e. gas travelling slower than the liquid), due to bubbles rising against the flow.

From the results of measured gas holdup in the mixing tube and the ratio \( \frac{Q_g}{Q_g + Q_L} \), it was possible to calculate the slip-velocity in the mixing tube, using the following equation:

\[
\nu_{\text{slip}} = \nu_{sg} \left( \frac{\phi}{\frac{Q_g}{Q_g + Q_L} - 1} \right) \frac{1}{\phi (1 - \phi)}
\]

(Eqn. 22)

This equation is derived as shown in Appendix D.

Figure 19 shows the effect of gas-liquid flow ratio on slip-velocity. For the range of flowrates tested, an increase in \( \frac{Q_g}{Q_L} \) increases the slip-velocity in the mixing tube, for bubble and intermediate flow regimes. At liquid flowrates up to \( Q_L = 1.5 \text{ls}^{-1} \), the slip-velocity does not increase much above \( 4 \text{ms}^{-1} \) at \( \frac{Q_g}{Q_L} \) higher than 1.0. This is due to the annular flow regime at these conditions. For the higher liquid and gas flowrates, the slip-velocity appears to approach \( 10 \text{ms}^{-1} \). This does seem larger than expected, but at high liquid and gas flowrate, slip-velocity calculation is very sensitive to small errors in holdup measurement. The error in holdup measurement (± 10%) could give up to \( 5 \text{ms}^{-1} \) difference in slip-velocity calculation.
However, even the moderate values of slip-velocity are much higher than might be expected from bubble rise velocities (less than 1 m s\(^{-1}\)). The high slip-velocities are still thought to be possible in the ejector system. Slip-velocity is calculated using a value of overall holdup, but there is a region in the system where there is a free jet in the pipe. This is the point where gas is entrained in the gas chamber and the free jet also exists in the beginning of the mixing tube. Very high slip velocities (approaching the liquid jet velocity) will occur in this local area hence influencing the overall slip-velocity.

From the results in Figure 19, it is suggested that the following equations are used for slip-velocity:

\[
\nu_{\text{slip}} = 5 \left( \frac{Q_G}{Q_L} \right) \text{ms}^{-1} \quad (\text{Eqn. 23})
\]

For \(0 \leq \frac{Q_G}{Q_L} \leq 0.8\), and

\[
\nu_{\text{slip}} = 4 \text{ms}^{-1} \quad (\text{Eqn. 24})
\]

for \(0.8 \leq \frac{Q_G}{Q_L} \leq 2.5\).

Apart from the extreme values of \(\nu_{\text{slip}}\), these equations approximate \(\nu_{\text{slip}}\) within \(\pm 1 \text{ms}^{-1}\).

4.3 SPECIFIC INTERFACIAL AREA AND LIQUID-SIDE MASS TRANSFER COEFFICIENTS

Values of \(k_L\alpha\) obtained previously by Zhu (Z2 and Z3) and holdup in this study are all average values over the whole mixing tube. To calculate specific interfacial area and \(k_L\), one also needs the average bubble size along the whole mixing tube. As shown in Table 7, bubble size varies considerably over the mixing tube. Unfortunately, most of the photographs were taken at one position (position 3). Therefore estimation of \(\alpha\) and comparison between different flow conditions are limited at this stage. Nevertheless, values of \(\alpha\) and \(k_L\) can be estimated for one flow condition (\(Q_L = 2.1 \text{ls}^{-1}\), \(Q_G/Q_L = 0.48\)) where 3 axial positions were examined. Table 10 shows values of overall \(\alpha, k_L\) for this particular flow condition.
Chapter 5 - EFFECT OF COALESCENCE INHIBITION - RESULTS AND DISCUSSION

In this work, experiments were carried out to examine the effects of flow parameters on gas entrainment, mass transfer, gas holdup and bubble size with a non-coalescing system and to compare these results to a coalescing system. The geometry of the ejector was kept constant throughout the experiments.

5.1 FLOW PATTERN

As part of the experimental tests, the flow pattern at the end of the mixing tube was measured using a fibre optic probe device as well as observing the presence of a jet. The attachment position of the jet was measured when present.

It was found that in the Magnesium Sulphate solution, even at high gas or liquid flow rates, the flow regime remained in bubbly flow. This is unlike an air-water or coalescing system where annular flow developed at high gas or low liquid flow rates.

A jet was not always present for low gas flow rates but was observed only when the gas-liquid flowrate ratio was above 0.5. When formed the jet always attached to the mixing tube wall at the same position, a distance of approximately 150 mm from the nozzle. The mixing shock region was observed to be present in the mixing tube immediately downstream of this attachment position.

5.2 ENTRAINMENT RESULTS

Figure 20 presents the effect of flow parameters on entrainment performance, for the air-salt solution system. The graph plotted is mixing tube pressure drop against average gas-liquid flow ratio in the mixing tube. The negative pressure drop indicates a pressure rise down the mixing tube, and, because downstream pressure \( (p_2) \) is held constant, the higher negative value indicates a smaller gas chamber pressure. The graph shows that an increase in liquid flowrate gave a lower gas chamber pressure and hence resulted in better entrainment. For a particular liquid flowrate, pressure rise down the mixing tube decreases approximately linearly with increasing gas flowrate.
The exception to this trend is at very low gas flowrates, where $Q_d/Q_L < 0.5$. This is due to the ejector not having a free jet through the first part of the mixing tube, i.e. the gas chamber is full of liquid as discussed in Section 5.1.

The effects described above are similar to what has been observed previously for an air-water system, as shown in Figure 21. This graph shows that there is little or no effect of coalescence inhibition on entrainment. Any differences between the curves is within the range of experimental error or due to the slight difference in density between the pure water and salt solutions.

Figure 22 presents the effect of flow parameters on ejector pumping efficiency for the air-salt solution. The graph plotted is efficiency against gas-liquid flow ratio for a range of liquid flowrates. The figure shows that up to $Q_d/Q_L$ of 1.0 the effect of liquid flowrate is negligible. At this ratio the pumping efficiency reached a maximum for $Q_L = 2.1$ l/s, but continued to rise for the higher liquid flowrates. As gas-liquid flow ratio increased above 1.0, the effect of liquid flowrate was more noticeable, with a higher liquid flowrate achieving higher pumping efficiencies.

These effects were similar to those observed for an air-water system (Z2). The comparison between the two systems can be seen in Figure 23. For the air-salt solution, efficiencies were lower than for an air-water system. For a liquid flowrate of 2.1 l/s, maximum efficiency is achieved at about $Q_d/Q_L = 1.0$ for both systems. At higher liquid flowrates, and high gas-liquid flow ratios, the maximum efficiency occurs later for the salt solution and hence gives a similar efficiency to the water system.

Therefore, calculation of gas chamber pressure in the design of a gas-liquid ejector will follow the same steps as for an air-water system.

### 5.3 BUBBLE SIZE MEASUREMENTS

Photographs of air bubbles in salt solution were taken at three axial positions in the mixing tube and at different liquid and gas flowrates. As for the air-water system, the Sauter mean diameter, $d_{12}$, was calculated as using Equation 18.
All photographs were taken at a fixed radial position of 2 mm from the pipe wall, since earlier results (Section 4.1.3) showed that the value of \( d_{32} \) taken at 2 mm from the pipe wall can be taken as a good approximation for the whole cross-section of the pipe.

5.3.1 Minimum Bubble Count

In the work covered in Chapter 4 on air-water systems in gas-liquid ejectors, the minimum bubble count was found to be 200 - 250 bubbles. For the range of flowrates studied in this chapter, approximately 20 - 30 photographs were taken to ensure that at least this minimum bubble count was achieved. For some conditions, many photographs were taken in order to check that this minimum bubble count was also appropriate for this system.

As for an air-water system, the minimum bubble count required to give an accurate value of mean bubble size/\( d_{32} \) was determined from analysis of negatives. For about half of the conditions tested, the minimum bubble count appeared to be about 100 but, for other conditions, this value was around 200 - 250. This is similar to the findings with an air-water system as shown in Section 4.1.1. Examples of these extremes of minimum bubble count estimation can be seen in Figure 24, which shows a graph of \( d_{32} \) against bubble count. Error bars of ±5% from the final values of \( d_{32} \) are also plotted on the same graph in order to observe the minimum bubble count required to achieve this accuracy.

5.3.2 Bubble Size Distribution (BSD)

For each set of flow conditions, a histogram was produced to observe the bubble size distribution. Figure 25 shows a typical example. The BSD was non-Gaussian and showed positive skewness. This can be explained by the salt solution having a very low coalescence rate. No system is completely non-coalescent, hence a limited number of larger bubbles form as indicated by the long tail. Figure 26 shows a comparison between the bubble size distributions for an air-salt solution and air-water system at the same experimental conditions. This clearly shows that the mean bubble diameter in the non-coalescing fluid was smaller than in air-water.
5.3.3 Effects of Positions and Flow Parameters

Measurements were carried out at various liquid and gas flowrates and axial positions in the mixing tube at a radial position 2 mm from the pipe wall. The flow regime in which these measurements were taken was bubble flow.

All the results are shown in Tables 7 - 9, with the corresponding results for an air-water system.

Effect of axial position

For each flow condition, local $d_{32}$ was calculated at the various axial positions and an overall $d_{32}$ determined where possible. The number of bubbles counted and the value of standard deviation ($\sigma$), calculated using Equation 19, are also included in these tables.

Table 7 shows that at a given flow condition, local $d_{32}$ increased as the flow progressed down the mixing tube. There are two possible reasons for the change in $d_{32}$ along the mixing tube: bubble coalescence and change in local pressure. Local pressure usually rises along the ejector mixing tube which would cause a decrease in bubble size. Therefore the increase in $d_{32}$ along the mixing tube can only be explained by the presence of bubble coalescence. This suggests that despite the aim of a non-coalescing system, there is a degree of bubble coalescence that occurs as flow progresses down the mixing tube.

Three main observations were made when comparing the results between air-salt solution and air-water system:

i) The bubbles observed in the salt solution are much more spherical than found in the air-water system.

ii) Sauter mean diameter for the salt solution is significantly lower than for an air-water system.

iii) The increase in $d_{32}$ down the mixing tube is less significant than for air-water.
Observations i) and ii) can be seen in Figure 27 which shows photographs of air bubbles in water and salt solutions under the same flow conditions. These photographs also demonstrate to some extent the difference in surface mobility. The air bubbles in water obviously appear to be deformable, whereas in the salt solution the bubbles are smaller and more spherical, implying the move towards a rigid surface.

In the photograph of the air bubbles in salt solution there appear to be bubbles inside bubbles. Analysis of several of these picture and the positions of the ‘internal’ bubbles confirms that there is always a corresponding ‘external’ bubble close by and at the correct angle to generate this image. This can be put down to the presence of more bubbles in the mixing tube and the bubbles being more spherical than for water. The bubbles closer to the mixing tube wall act as lenses (due to the spherical shape), causing the bubbles behind them to appear inside.

A smaller  in the early part of the mixing tube (axial position 1) for the salt solution implies that in this system bubbles are easier to break up. The initial break-up occurs in the mixing shock region forming primary bubbles in the mixing tube. Bhavaraju et al. (B3) found that in gas-liquid systems break-up is affected by surface tension and the density of the surrounding fluid.

The density and surface tension were measured and the values were found to be almost the same as for water. The differences between these values would not be expected to change the break-up significantly. The addition of salt therefore appears to introduce an additional factor influencing bubble break-up. This salt presence affects the charge distribution, as surfaces are extended and deformed. It is thought this may lead to the production of smaller initial bubbles than compared to pure water.

Comparing results at axial positions 2 and 3, although  increased along the mixing tube, the rate of increase was smaller than with air-water (Table 7). This means that with air-salt solution, bubble coalescence is still present, but has been suppressed to some extent. Figure 1 showed coalescence rates with respect to salt concentration for stagnant fluids. In the mixing tube there are high levels of turbulence in which coalescence is more probable than this figure suggests. In Figure 1, it shows that at the concentration of Magnesium Sulphate used in these tests gives low coalescence rates, where as for a water system, the coalescence rate is high. Hence, the increase in bubble size down the
mixing tube for this salt solution is less significant than for water due to the large difference in coalescence rates.

**Effect of gas flowrate**

The effect of gas flowrate on $d_{32}$ was investigated at position 3 at liquid flowrates of 1.2 and 2.1 ls$^{-1}$. Gas flowrate was changed such that the gas-liquid flow ratios tested were 0.24 to 0.95 for $Q_L = 2.1$ ls$^{-1}$ and a slightly lower range for $Q_L = 1.2$ ls$^{-1}$. An increase in gas flowrate decreases $d_{32}$ for both liquid flowrates, as can be seen in Table 8. This is similar to air-water and can be explained by the increase in power input per unit mass causing increased break-up of bubbles.

The $d_{32}$ in salt solution was again found to be smaller than in water. This is partly due to the non-coalescing nature of the system, because measurements at the end of the mixing tube were compared, as previously discussed. The results show that $d_{32}$ in salt solution does not change as rapidly with gas flowrate as for water. The results also reflect the direct effect of salt on bubble size, also previously discussed.

By using measured values of average gas holdup and power per unit mass in Equation 20 (found to enable bubble prediction in static mixers (B2)), an increase in gas flowrate should produce a small increase in bubble size. This is not what has been observed in these tests and shows again that this correlation for bubble size prediction is not applicable, due the complexity of the system.

**Effect of liquid flowrate**

Table 9 shows the effect of liquid flowrate on $d_{32}$ at constant gas-liquid flow ratios. At axial position 1, an increase in liquid flowrate, caused a decrease in $d_{32}$. This position is just below the mixing shock region, just below where initial break-up occurs. An increase in liquid flowrate increases the liquid power input per unit mass, hence there is more energy lost from the liquid in creating the mixing shock region and inducing an increase in break-up of gas into fine bubbles.

At axial positions 2 and 3, the effect of increasing liquid flowrate has negligible effect on $d_{32}$. Photographs illustrating this result can be seen in Figure 28.
Overall, $d_{32}$, calculated by combining raw data from all axial positions (using Equation 18) shows a very small reduction with liquid flowrate.

This is an unexpected result, due to the significant increase in liquid power input per unit mass over this range of liquid flowrates. Also when attempting to use Equation 20 to predict the trends, there shows a consistent decrease in bubble size (outside experimental error bands). Hence the salt solution must have such a severe effect on bubble size that it is unaffected by liquid flowrate changes as the flow developed down the mixing tube.

However, one reason for this observation and finding may be due to the difference in the type of bubbles in the air-water and air-salt solution systems. As discussed in Chapter 2, for rigid bubbles (as the air in salt solution is thought to create), increasing the energy dissipation rate causes an increase in coalescence efficiency (from Equation 7). In general, an increase in the energy dissipation rate would be expected to create more turbulence and hence increase break-up leading to a decrease in bubble size. If the coalescence efficiency also increases, then bubble size could remain constant if the two opposing effects are of similar magnitude.

**Bubble size prediction**

From the results of bubble sizes at various positions in the mixing tube, the following correlation has been developed:

$$
d_{32} = 0.66 \frac{P_r}{V_d}^{-0.055} \left( \frac{Q_G}{Q_G + Q_L} \right)^{0.58} \left( \frac{L}{L_m} \right)^{0.28}$$

(Eqn. 25)

$R^2 = 0.876$

for $1.2 \text{ls}^{-1} \leq Q_L \leq 2.7 \text{ls}^{-1}$ and $Q_G/Q_L < 1.0$ and $d_{32}$ is in mm.

The weak function of power per unit volume is slightly surprising but does reflect with the effects observed earlier in this section. The various terms in this correlation are consistent with $k_{j,a}$ and previous bubble size correlations and appear to give good predictions as demonstrated in Figure 29.
The exponent on the $L/L_m$ term is less than half that in an air-water (coalescing) system. This implies that down the mixing tube the size of the bubbles increases, but to a lesser extent than water. This again shows that the correlation bears a realistic representation of the results observed.

5.4 GAS HOLDUP MEASUREMENTS

The local gas holdup was measured at three axial positions in the mixing tube, as summarised in Section 3.5. For all conditions tested, the flow pattern in the mixing tube was bubbly flow. For most conditions tested, when a free jet was formed before the mixing tube, the local holdup was larger at axial position 1 and decreased down the mixing tube. This could partly be attributed to the large difference in pressures as shown in Figure 20. An example of these results can be seen in Figure 30 at $Q_G/Q_L = 0.92$ and 1.48.

The higher gas holdup at position 1 for higher $Q_G/Q_L$ is similar to the ratio of gas volume over the total volume as when the liquid jet remains intact. Calculations show that this volume ratio is 0.83. This similarity indicates that a bubble in liquid dispersion has not fully formed at this position.

For very low gas-liquid flow ratios of 0.22 and 0.48 the variation of holdup down the mixing tube was negligible. This was because the gas flow was so low that the liquid flooded the gas chamber and no free jet was formed.

The overall gas holdup in the mixing tube was calculated by averaging the three local measurements. This was considered a reasonable approach to take due to the way the local values were spread where the local holdup at axial position 2 was approximately half-way between the values at positions 1 and 3, both when a free jet was and was not present.

Figure 31 shows the effect of liquid and gas flowrate on overall gas holdup in the mixing tube. The graph plotted is gas holdup against gas-liquid flow ratio due to the findings in Chapter 4 showing this to be the best relationship. Increasing the gas flowrate significantly increased the gas holdup, up to the condition at $Q_G/Q_L \geq 1.0$ where gas holdup was independent of both gas and liquid flowrate.
Most of the data correlate well together, providing a free jet exists. However, for experimental conditions where there is no free jet ($Q_g/Q_L < 0.5$ for $Q_L \geq 2.1 \text{ ls}^{-1}$) and the gas chamber is flooded the correlation breaks down. Operation of the ejector in this mode is not recommended due to the comparatively low intensity mixing.

Figure 32 shows the direct comparison between the salt solution and water systems. Despite the use of different techniques, the comparison is feasible, as discussed in Section 4.2 and shown in Figure 17. When comparing gas holdup between these two systems, the trends observed are the same. In both systems, gas-liquid ratio has a large effect on holdup for $Q_g/Q_L$ up to 1.0, and only a slight effect above this ratio. The only significant difference occurred at $Q_g/Q_L = 0.5$, but this is the point where a free jet has just formed or is about to form.

It can be seen and concluded that salt in the system has little or no effect on gas holdup. This result is similar to the finding by Keitel and Onken (K2) who reported that gas holdup was practically unaffected by electrolyte concentration.

The following approximate correlation can be used to predict gas holdup for a coalescing or non-coalescing system:

$$\phi = 0.4 \frac{Q_g}{Q_L} + 0.3 \quad (\text{Eqn. 26})$$

for $0.48 \leq \frac{Q_g}{Q_L} \leq 1.25$ providing a free jet exists, and

$$\phi = 0.8 \quad (\text{Eqn. 27})$$

for $Q_g/Q_L > 1.25$

These equations approximate gas holdup within ±0.1.

5.4.1 Slip-velocity calculations

Figure 33 compares the measured gas holdup with $Q_g/(Q_g + Q_L)$. The graph shows that measured gas holdup in the mixing tube is generally consistently higher than $Q_g/(Q_g + Q_L)$, by about 20 - 30%. In some cases, as with air-water, at low gas-liquid
flow ratios, measured gas holdup is less than \( \frac{Q_G}{(Q_G + Q_L)} \). This again occurs when the gas chamber is flooded and no free jet is present. This difference is the same as that found with water and expectedly so, due to the negligible difference in gas holdup.

From the results of measured gas holdup in the mixing tube and the ratio \( \frac{Q_G}{Q_G + Q_L} \) it was possible to calculate the slip-velocity in the mixing tube, using Equation 22.

Figure 34 shows the effect of gas-liquid flow ratio on slip-velocity in Magnesium Sulphate solution. For the range of flowrates tested, an increase in \( \frac{Q_G}{Q_L} \) increases the slip-velocity significantly in the mixing tube up to \( \frac{Q_G}{Q_L} = 0.8 \), giving a slip-velocity of \( 4 \text{ ms}^{-1} \). As discussed in Section 4.2.1, for the higher liquid and gas flowrates, the slip-velocity appears to be up to twice this value. This does seem larger than expected, but is of the same order as that found with an air-water system and, at high liquid and gas flowrate, slip-velocity calculation is very sensitive to small errors in holdup measurement.

These calculated slip-velocities are of the same order of a water system. This was expected due to similar gas holdup values at corresponding gas and liquid flowrates. A direct comparison between these systems can be seen in Figure 35. This figure indicates that slip-velocity can be estimated using the same equations suggested for air-water in Section 4.2.1, i.e.:

\[
\text{Eqn. 28}
\]

\[
\nu_{\text{slip}} = 5 \left( \frac{Q_G}{Q_L} \right) \text{ ms}^{-1}
\]

For \( 0 \leq \frac{Q_G}{Q_L} \leq 0.8 \), and

\[
\text{Eqn. 29}
\]

\[
\nu_{\text{slip}} = 4 \text{ ms}^{-1}
\]

for \( 0.8 \leq \frac{Q_G}{Q_L} \leq 2.5 \).

Apart from the extreme values of \( \nu_{\text{slip}} \), these equations approximate \( \nu_{\text{slip}} \) within \( \pm 1 \text{ ms}^{-1} \).
5.5 MASS TRANSFER RESULTS

Figure 36 shows mass transfer coefficient against gas-liquid flow ratio for a range of liquid flowrates for the salt solution. Two sets of tests were carried out and the repeats all displayed in this figure.

At constant gas-liquid flow ratio, an increase in liquid flowrate gives a higher value for $k_La$, as seen previously for water (Z2). For each of the liquid flowrates tested, an increase in gas flowrate increases $k_La$ to a maximum at approximately $Q_G/Q_L = 1.0$. Following this, $k_La$ actually decreases with an increased gas input. This seems an odd result that a decrease in $k_La$ should occur despite the flow regime remaining bubbly.

Above this ratio of $Q_G/Q_L = 1.0$, holdup remains almost constant. Therefore changes in $k_La$ must be due to a change in bubble size. However, bubble size measurements were not made above $Q_G/Q_L = 1.0$ and hence no conclusive reasons for this effect can be achieved.

It was previously found in a gas-liquid ejector (Z2) that the relationship of mass transfer coefficient with gas flowrate depended on the flow pattern where, in bubble flow, $k_La$ increased rapidly with gas flowrate and, in annular flow, $k_La$ was independent of gas flowrate. However, for this salt solution where all conditions were in bubble flow regime, mass transfer coefficient was dependent on gas-liquid flow ratio, where $k_La$ increased up to $Q_G/Q_L = 1.0$, and then decreased slightly at higher $Q_G/Q_L$.

As suggested by Zhu (Z3), mass transfer coefficient could be correlated using power per unit volume ($P_r/V_d$) and gas holdup ($\phi$). From experimental data obtained in this work, over the whole range, a multiple regresional analysis produced the following correlation:

$$k_La = 2.4 \times 10^{-3} \left( \frac{P_r}{V_d} \right)^{0.56} \phi^{0.16}$$

(Eqn. 30)

$$R^2 = 0.658$$

where $\phi$ is the measured gas holdup.
As can be seen in Figure 37, the majority of the data lies within ± 20%, which is thought to be a reasonable error band for mass transfer.

\[ k_{L,a} \] can also be correlated with respect to flow ratio as follows:

\[
k_{L,a} = 7.6 \times 10^{-3} \left( \frac{P_T}{V_d} \right)^{0.50} \left( \frac{Q_G}{Q_G + Q_L} \right)^{0.48} \text{ for } \frac{Q_G}{Q_L} < 1.0 \quad \text{(Eqn. 31)}
\]

\[ R^2 = 0.815 \]

\[
k_{L,a} = 2.47 \times 10^{-4} \left( \frac{P_T}{V_d} \right)^{0.71} \text{ for } \frac{Q_G}{Q_L} > 1.0 \quad \text{(Eqn. 32)}
\]

\[ R^2 = 0.946 \]

Figure 38 gives a graphical representation of these two correlations and combined with the values of \( R^2 \), shows that they give a better prediction for \( k_{L,a} \) than Equation 30. However, the 2 correlations may not give the same value for \( k_{L,a} \) at \( Q_G/Q_L = 1.0 \), hence should be used cautiously, particularly at this condition.

Comparing with the correlations for pure water:

\[
k_{L,a} = 2.2 \times 10^{-3} \left( \frac{P_T}{V_d} \right)^{0.6} \left( \frac{Q_G}{Q_G + Q_L} \right)^{0.55} \text{ for } \frac{Q_G}{Q_L} \leq 2 \quad \text{(Eqn. 33)}
\]

\[
k_{L,a} = 1.8 \times 10^{-3} \left( \frac{P_T}{V_d} \right)^{0.6} \text{ for } \frac{Q_G}{Q_L} > 2 \quad \text{(Eqn. 34)}
\]

When comparing Equations 31 and 32 with 33 and 34 respectively, the exponent values overlap when taking into consideration the standard errors of coefficients.
Figure 39 shows that $k_L a$ values in pure water and salt solution are very similar for the same flow conditions, up to $Q_G/Q_L = 0.9$. Above this gas-liquid flow ratio, $k_L a$ is consistently lower for the salt solution.

These results are somewhat surprising since the Sauter mean diameter is significantly smaller for the salt solution (as discussed in Section 5.3). Hence, with gas holdup being constant for the two systems, the presence of salt increases specific interfacial area. This result indicates that for the addition of salt, which increases $a$, there must be a corresponding decrease in $k_L$, making the overall effect negligible.

As discussed in Chapter 2, the decrease in $k_L$ could be due to a number of reasons, as $k_L$ is dependent on many properties of the liquid, including surface tension, liquid density and gas diffusivity in the liquid phase.

The results discussed in this section provide useful correlations that can be used in predicting overall mass transfer in a non-coalescing gas-liquid ejector system. Hence it is possible to estimate this important parameter in the design of a gas-liquid ejector.

### 5.6 SPECIFIC INTERFACIAL AREA AND LIQUID-SIDE MASS TRANSFER COEFFICIENTS

Specific interfacial area was calculated using the overall Sauter mean diameter and overall gas holdup in the mixing tube, using Equation 17. Overall mass transfer coefficients were obtained as detailed in Section 5.5. By using these results, liquid volumetric mass transfer coefficient could then be calculated. Table 11 shows the limited corresponding results for $a$ and $k_L$. 
Chapter 6 - EFFECT OF LIQUID VISCOSITY - RESULTS AND DISCUSSION

In this area of work, experiments were carried out to examine the effects of flow parameters on gas entrainment, bubble size, gas holdup and mass transfer with a 'viscous' system and to compare these results to a water of viscosity 1 mPas. The geometry of the ejector was unchanged throughout the experiments.

6.1 FLOW PATTERN

As with the experimental tests carried out in the salt solution, flow pattern was measured at the end of the mixing tube together with observations as to whether a jet was present.

It was found, as with air-water (Z2 and Z3) that an annular flow develops in the mixing tube at low liquid flowrates or high gas flowrates. There were also similarities with the salt solution with respect to the free jet: at low gas flowrates, a jet was not always present and formed when the gas-liquid flow rate ratio was above 0.5; the jet attachment position to the mixing tube wall was approximately 150 mm downstream of the nozzle and the mixing shock region was present immediately after the attachment of the jet.

6.2 ENTRAINMENT RESULTS

Figures 40 - 42 present the effect of flow parameters on entrainment performance, for air-CMC solution of viscosity 3, 6 and 10 mPas respectively. The graph plotted is mixing tube pressure drop against gas-liquid flow ratio. As discussed earlier, the negative pressure drop indicates a pressure rise down the mixing tube, and because downstream pressure is held constant, the higher negative value indicates a smaller gas chamber pressure. The graphs show that for the range of viscosities an increase in liquid flowrate gave a lower gas chamber pressure and hence resulted in better entrainment. Also, for a particular liquid flowrate as the gas flowrate is increased, in general, the difference in pressure down the mixing tube decreased. All these three graphs show the same trends and very similar results. The effects described above are as expected and similar to the previous observations for an air-water system.
Figure 43 shows the comparison between all viscosities and water at a liquid flowrate of 2.1 $\text{ls}^{-1}$. For the viscosity of 10 mPas, the data points shown are the average of a number of repeats, with error bars. This graph shows that for the range of liquid viscosities, at $Q_g/Q_L < 0.75$, there seems to be negligible effect on entrainment. Any differences are within the bands of experimental error. However, at gas-liquid flow ratios above 0.75, for viscosities of 3 and 6 mPas there seems to be an effect on gas entrainment. Both of these viscosities show the same gas entrainment which was 10 - 30% less compared to water. For a viscosity of 10 mPas, there appears to be no difference in gas entrainment from 1 mPas. Included in this figure are error bars on the 10 mPas results which do not cause overlap with the 3 and 6 mPas results. Therefore an increase in viscosity up to 6 mPas has an effect on gas entrainment and an increase to 10 mPas appears to have no effect on gas entrainment when compared to water.

Figures 44 - 46 present the effect of flow parameters on ejector pumping efficiency for the air-CMC solution for viscosities of 3, 6 and 10 mPas respectively. The graphs plotted are efficiency against gas-liquid flow ratio for a range of liquid flowrates. The figures show that for $Q_L \geq 2.1 \text{ls}^{-1}$, an increase in gas flowrate initially increases the efficiency of the ejector, and up to $Q_g/Q_L$ of 1.0 the effect of liquid flowrate is negligible. Above this ratio, maximum pumping efficiencies were achieved, depending on flowrate and viscosity, as can be seen in these figures.

At liquid flowrates below 2.1 $\text{ls}^{-1}$, at very low gas-liquid flow ratios, pumping efficiencies were comparable to those of higher liquid flow and an increase in gas flowrate had a negative effect on pumping efficiency. These effects were also observed for an air-water system. The comparison between these systems at $Q_L = 2.1 \text{ls}^{-1}$ can be seen in Figure 47. At low gas-liquid flow ratios, viscosity had negligible effect on pumping efficiency. At higher flow ratios, for viscosities of 3 and 6 mPas, pumping efficiency was lower than for a water system (by 10 - 30%), as for gas entrainment. At a viscosity of 10 mPas compared to 1 mPas, there appeared to be little effect on efficiency. For all viscosities, the maximum pumping efficiency occurred at similar gas-liquid flow ratios.

For all the results discussed in this section, the 10 mPas data does not appear to be consistent with the trends shown with the 3 and 6 mPas results. For all the tests, the fluids in the mixing tube were in fully turbulent flow. The CMC used for the 10 mPas
tests was from a different batch and could be the reason for this slight discrepancy with respect to the entrainment and efficiency results.

From these results, it would be acceptable to adopt the same method as an air-water system in calculating the gas chamber pressure in the design of a gas-liquid ejector for liquid viscosities up to 10 mPas.

6.3 BUBBLE SIZE MEASUREMENTS

Photographs of air bubbles in a viscous solution were taken at various axial positions in the mixing tube and at different liquid and gas flowrates. Limited photographs were taken and some pictures analysable for 10 mPas CMC solution. A more thorough set of pictures were taken and analysed for 6 mPas CMC solution. In the following sections, the results will concentrate on the 6 mPas CMC solution and will be compared to 1 mPas. However, the limited results from 10 mPas are presented in Table 12.

6.3.1 Minimum Bubble Count

Again, as for the air-salt solution, approximately 20 - 30 photographs were taken to ensure that the bubble count was in excess of 200 - 250 bubbles as for the air-water minimum bubble count. For some conditions, many photographs were taken in order to check that this minimum bubble count was also appropriate for this system.

The minimum bubble count required to give an accurate value of mean bubble size/ $d_{32}$ was determined from analysis of negatives. For the conditions tested, the minimum bubble count was found to be around 250 - 300. This is slightly larger than the findings with air in water system (Section 4.1.1). A typical example of minimum bubble count estimation determined above can be seen in Figure 48, which shows a graph of $d_{32}$ against bubble count. Error bars of ±5% from the final values of $d_{32}$ are also plotted on the same graph in order to observe the minimum bubble count required to achieve this accuracy.
6.3.2 Bubble Size Distribution (BSD)

For each set of flow conditions, a histogram was produced to observe the bubble size distribution. Figure 49 shows a typical example that the BSD was non-Gaussian and showing positive skewness. Due to the liquid viscosity reducing coalescence rates following the initial bubble formation, limited larger bubbles are produced. Therefore, the mode of the distribution is at low bubble sizes and the BSD has a long tail.

6.3.3 Effects of Positions and Flow Parameters

Measurements were carried out at various liquid and gas flowrates and axial positions in the mixing tube at a radial position 2 mm from the pipe wall. The majority of these measurements were taken in the bubbly flow regime.

All the results for a viscosity of 6 mPas are shown in Tables 13 - 15 with the corresponding results for an air-water system.

Effect of axial position

As with previous experimental tests on bubble size, local \( d_{32} \) for various positions was calculated as well as an overall \( d_{32} \) where possible. The number of bubbles counted and value of standard deviation (\( \sigma \)) are also included in these tables where \( \sigma \) is calculated using Equation 19.

Table 13 shows that at a given flow condition, local \( d_{32} \) increased slightly as the flow progressed down the mixing tube, due to the presence of bubble coalescence as discussed in Chapter 5. This ionic solution does not reduce coalescence rates, as summarised in Section 3.2.2 however, the viscosity of the liquid in a film opposes the drainage of the film and its displacement by the approach of coalescing bubbles. The higher the viscosity, the slower will be the film thinning process. Hence, with the increase in viscosity there is a reduction in coalescence rate. This increase in \( d_{32} \) therefore suggests
that some bubble coalescence does occur as flow progresses down the mixing tube. Three main observations were made when comparing the results between air-CMC solution and air-water system:

i) the bubbles observed in the viscous system are much more spherical than found in the air-water system

ii) the Sauter mean diameter for dispersion of the viscous system is significantly less than for an air-water system

iii) the increase in $d_{32}$ along the mixing tube is less significant than for air-water.

Observations i) and ii) can be seen in Figure 50 which shows photographs of air bubbles in water and viscous solutions under the same flow conditions.

A smaller $d_{32}$ in the early part of the mixing tube (axial position 1) for the viscous solution indicates that in this system bubbles are easier to break up. The initial break-up occurs in the mixing shock region forming primary bubbles in the mixing tube. Bhavaraju et al. (B3) found that in gas-liquid systems break-up is affected by surface tension and the density of the surrounding fluid.

The density and surface tension were measured and the values found to be close to those for water (Section 3.2.2). The differences between these values would not be expected to cause significant change to break-up. Therefore there must be an additional factor influencing bubble break-up which is thought to be due to the increase in viscosity. The higher viscosity causes the formation of smaller bubbles and the salt presence affects the charge distribution, as surfaces are extended and deformed, leading to the production of smaller initial bubbles than in ordinary water.

Comparing results at axial positions 2 and 3, although $d_{32}$ increased along the mixing tube, the rate of increase was smaller than with air-water (Table 13). This means that with the air-viscous solution, bubble coalescence is still present, but has been suppressed to a large extent. Hence, the increase in bubble size down the mixing tube for this viscous solution is less significant than for water due to the difference in coalescence rates.
Figure 51 shows a comparison between 1 mPas and 6 mPas solutions and Magnesium Sulphate solution (from Chapter 5) with respect to the effect of axial position on $d_{32}$. The initial bubble size is the same for 6 mPas and salt solutions, indicating that the salt presence affecting the charge distribution has a more dominant effect on the break-up mechanism than viscosity. As flow develops down the mixing tube, the increase in bubble size for 6 mPas is less significant than for the salt solution showing that the increase in viscosity has a greater effect on coalescence inhibition than the presence of a salt.

The bubble sizes in the mixing tube for a viscosity of 10 mPas also show the same trend. There is an increase in bubble size as flow progresses down the mixing tube but again it is not as significant as for the 1 mPas solution. The bubble size for this viscosity of 10 mPas is however larger than for 6 mPas. This finding is consistent with the entrainment results. Due to better entrainment, there is less energy loss in the system for the 10 mPas solution leading to bigger bubbles and better entrainment performance. It is not clear whether this is a viscosity effect or due to the fact that the 10 mPas solution was made from a different batch of CMC and further investigation to confirm these findings would be advisable.

**Effect of gas flowrate**

The effect of gas flowrate on $d_{32}$ was investigated at position 3 at a liquid flowrate of 2.1 ls$^{-1}$ (Table 14). Gas flowrate was changed such that the gas-liquid flow ratios tested ranged from 0.24 to 0.95. An increase in gas flowrate was seen to increase $d_{32}$ up to the condition where $Q_g/Q_L = 0.71$. This is the opposite effect to air-water.

Increasing the gas flowrate such that $Q_g/Q_L$ increases from 0.71 to 0.95 has little effect on $d_{32}$. The difference is within experimental error bands and occurs when the flow pattern verges on intermediate. Again, this effect is different from air-water. These results (at axial position 3) are unexpected due to the low coalescence rates as seen in Figure 51 (increased gas increases the probability of coalescence due to a larger proportion of gas present in the mixing tube) and the fact that an increase of gas increased the power input per unit mass.
By using measured values of average gas holdup and power per unit mass in Equation 20, an increase in gas flowrate did not show any trend as observed for bubble size. This shows again that this correlation for bubble size prediction is not applicable for this complex system.

**Effect of liquid flowrate**

Table 15 shows the effect of liquid flowrate on \( d_{32} \) at constant gas-liquid flow ratios. At axial position 3, an increase in liquid flowrate caused a decrease in \( d_{32} \). An increase in liquid flowrate at constant \( Q_g/Q_L \), increases the liquid power input per unit mass hence there is more energy lost from the liquid in creating the mixing shock region and inducing more break-up of gas into fine bubbles. This produces smaller bubbles at axial position 1 and due to the small amount of coalescence occurring down the mixing tube, means that there will be correspondingly smaller bubbles at axial position 3, as liquid flowrate is increased.

The trend of the change in bubble diameter due to the change in liquid flowrate was consistent with Equation 20.
Bubble size prediction

The results obtained in this section can be correlated well together with power input per unit volume, \( Q_G/(Q_G+Q_L) \) and position in the mixing tube as follows:

\[
d_{32} = 35.1 \left( \frac{P_T}{V_d} \right)^{-0.23} \left( \frac{Q_G}{Q_G + Q_L} \right)^{1.1} \left( \frac{L}{L_m} \right)^{0.18}
\]  
(Eqn. 35)

\( R^2 = 0.991 \)

for \( 1.2 \text{ l/s} \leq Q_L \leq 2.7 \text{ l/s} \) and \( Q_G/Q_L < 1.0 \) and \( d_{32} \) is in mm.

A comparison between actual \( d_{32} \) and predicted \( d_{32} \), using the above equation can be seen in Figure 52.

The exponent on power input per unit volume is close to that found in an air-water system whereas the exponent on \( Q_G/(Q_G+Q_L) \) is of the opposite sign, thus demonstrating the observed effects of flowrates.

The relationship between \( d_{32} \) and \( (L/L_m) \) is significantly less than air-water and also less than the non-coalescing system. This again shows the effects observed with respect to coalescence as flow develops down the mixing tube, as displayed in Figure 51.

6.4 GAS HOLDUP MEASUREMENTS

The local gas holdup was measured for viscosities of 3 mPas and 6 mPas at three axial positions in the mixing tube, as summarised in Section 3.5. For most conditions tested, when a free jet was formed before the mixing tube, the holdup was larger at axial position 1 and decreased down the mixing tube. A typical example of these results achieved at both viscosities can be seen in Figure 53.

For very low gas-liquid flow ratios of 0.24 and 0.48 the variation of holdup down the mixing tube was negligible. This was because the gas flow was so low that the liquid flooded the gas chamber and no free jet was formed.
At conditions where a free jet existed, differences in the local gas holdup in the mixing tube were noticeable. This can be seen in Figure 53 at $Q_j/Q_L = 0.92$ and 1.38.

The higher gas holdup at position 1 for higher $Q_j/Q_i$ is again similar to the ratio of gas volume over the total volume as when the liquid jet remains intact (volume ratio 0.83). This similarity indicates that a bubble in liquid dispersion may not have formed at this position.

As with the measurements in an air-salt solution, the overall gas holdup in the mixing tube was calculated by averaging the three local measurements. Figures 54 and 55 show the effect of liquid and gas flowrate on overall gas holdup in the mixing tube for liquid viscosities of 3 mPas and 6 mPas respectively. The graphs plotted were gas holdup against gas-liquid flow ratio and show the same trends. Increasing the gas flowrate significantly increased the gas holdup, up to the condition where $Q_j/Q_i = 1.0$.

The majority of the data correlates well together providing a free jet exists. For experimental conditions where there is no free jet ($Q_j/Q_i < 0.5$ for $Q_L > 2.1 \text{ ls}^{-1}$, for both viscosities) the correlation is not applicable.

Figure 56 shows the direct comparison between the 1 mPas, 3 mPas and 6 mPas systems. When comparing gas holdup between these three systems, the trends observed are the same. For all systems, gas-liquid ratio has a large effect on holdup for $Q_j/Q_L$ up to 1.0, and only a slight effect above this ratio. Considering the difference in experimental techniques and the scatter for each group of data, the holdup values between the two sets of data can be regarded as being quite similar. The only significant difference occurred at $Q_j/Q_L = 0.5$, but this is the point where a free jet just forms or is about to form. It can be seen and concluded that an increase in viscosity up to 6 mPas has little or no effect on gas holdup.
Therefore, based on the above conclusion, the following approximate correlation can be used to predict gas holdup for a Newtonian system up to a viscosity of 6 mPas:

\[ \phi = 0.4 \frac{Q_g}{Q_L} + 0.3 \]  \hspace{1cm} (Eqn. 36)

for 0.48 \leq \frac{Q_g}{Q_L} \leq 1.25 providing a free jet exists, and

\[ \phi = 0.8 \]  \hspace{1cm} (Eqn. 37)

for \frac{Q_g}{Q_L} > 1.25

These equations approximate gas holdup within \( \pm 0.1 \). The equations are comparable with those found in the salt solution (Section 5.4) and indicate that both coalescence inhibition and viscosity have little or no effect on gas holdup.

6.4.1 Slip-velocity calculations

Figure 57 shows a typical comparison for different viscosities of the measured gas holdup and \( \frac{Q_g}{Q_g + Q_L} \). The graph shows that measured gas holdup in the mixing tube is generally higher than \( \frac{Q_g}{Q_g + Q_L} \), around 30+\% at \( \frac{Q_g}{Q_L} > 1.0 \). In some cases, at low gas-liquid flow ratios, measured gas holdup is less than \( \frac{Q_g}{Q_g + Q_L} \). This occurs (as seen before) when the gas chamber is flooded and no free jet is present. This difference is the same as that found with water and expectedly so, due to the negligible difference in gas holdup.

From the results of measured gas holdup in the mixing tube and the ratio \( \frac{Q_g}{Q_g + Q_L} \) it was possible to calculate the slip-velocity in the mixing tube, using Equation 22 as derived in Appendix D.

Figures 58 and 59 show the effect of gas-liquid flow ratio on slip-velocity for liquid viscosities of 3 mPas and 6 mPas respectively. For the range of flowrates tested, an increase in \( \frac{Q_g}{Q_L} \) increases the slip-velocity significantly in the mixing tube up to \( \frac{Q_g}{Q_L} = 0.8 \), where slip-velocity is estimated as 4 ms\(^{-1}\). For the higher liquid and gas flowrates, the slip-velocity appears to be up to twice this value. This does seem larger
than expected, but is of the same order as found with an air-water system and, at high liquid and gas flowrate, slip-velocity calculation is very sensitive to small errors in holdup measurement.

These calculated slip-velocities are of the same magnitude as those found with an air-water system. This was expected since similar flowrates and holdup values were involved. A direct comparison between these systems can be seen in Figure 60. Despite the scatter of the results, this figure indicates that slip-velocity can be estimated from the equations suggested in Section 5.4.1 i.e.:

\[ v_{slip} = 5 \left( \frac{Q_G}{Q_L} \right) \text{ms}^{-1} \]  
(Eqn. 38)

For \( 0 \leq \frac{Q_G}{Q_L} \leq 0.8 \), and

\[ v_{slip} = 4 \text{ms}^{-1} \]  
(Eqn. 39)

for \( 0.8 \leq \frac{Q_G}{Q_L} \leq 2.5 \).

Apart from the extreme values of \( v_{slip} \), these equations approximate \( v_{slip} \) within \( \pm 1 \text{ms}^{-1} \).

Note that if Equation 22 combined with either Equation 38 or 39 is used to predict holdup, large predictive errors in slip-velocity at high \( Q_G/Q_L \) only have a very small effect on predicted holdup.

6.5 MASS TRANSFER RESULTS

Figures 61 to 63 show mass transfer coefficient against gas-liquid flow ratio for a range of liquid flowrates for 3, 6 and 10 mPas respectively. Any repeat tests are displayed in these figures. Also shown on these graphs is the relevant flow pattern. At constant gas-liquid flow ratio, an increase in liquid flowrate gives a higher value for \( k_La \), as seen previously for water. For each of the liquid flowrates and viscosities tested, in the bubble flow regime, an increase in gas flowrate significantly increases \( k_La \). As the transition to intermediate to annular flow occurs, the increase in gas flowrate becomes less significant on \( k_La \) and in some instances causes a reduction in \( k_La \).
A similar result was previously found for water (Z2) in that the relationship of mass transfer coefficient with gas flowrate depended on the flow pattern where, in bubble flow, \(k_La\) increased rapidly with gas flowrate and, in annular flow, \(k_La\) was independent of gas flowrate.

Figures 64 and 65 show the effect of viscosity (up to 10 mPas) on \(k_La\) for liquid flowrates of 2.5 and 2.7 l/s respectively. These figures show that in general, an increase in viscosity decreases \(k_La\). These figures have been re-plotted as \(k_La\) against viscosity for various gas-liquid flow ratios, as shown in Figures 66 and 67.

An increase in viscosity due to the presence of CMC solution significantly decreases the Sauter mean diameter (as discussed in Section 6.3), with gas holdup remaining constant. Hence this increases specific interfacial area, \(a\). According to Onda et al. (O1), viscosity has an inverse effect on \(k_L\), due to increased mass transfer film thickness. Also, as seen by Calderbank, (C3), viscosity has an inverse effect on diffusivity and hence \(k_L\), i.e.:

\[
k_L \propto \mu_L^{-c}\]  \hspace{1cm} (Eqn. 40)

As suggested by Zhu (Z3), mass transfer coefficient could be correlated using power per unit volume and gas holdup. Due to mass transfer results being obtained for a number of viscosities and a significant effect seen, the data was correlated including a viscosity term. The flow ratio \(Q_G/(Q_G+Q_L)\) was used as in previous correlations (Equations 1 and 2) and limiting the data so that \(k_La\) still increased with gas flowrate, the following correlation was produced:

\[
k_La = 0.036 \left( \frac{P_T}{V_d} \right)^{0.41} \left( \frac{Q_G}{Q_G + Q_L} \right)^{0.48} \mu_L^{-0.18}\]  \hspace{1cm} (Eqn. 41)

\[R^2 = 0.801\]

for \(Q_G/Q_L \leq 1.0\) and \(\mu_L\) is in mPas.

The quality of the correlation is shown in Figure 68. The majority of the data points are within the ±20% limits. For gas-liquid flow ratios above 1.0, \(k_La\) remains constant or decreases, so use of Equation 41 would lead to over prediction. It is clear from this
figure that the above correlation may not be the optimum and further analysis may be necessary to achieve the optimum.

However, the exponents on power per unit volume and the flow ratio \((Q_G/(Q_G+Q_L))\) are similar to those seen in a pure water system (Equation 1). This is not surprising due to a number of the data points representing an air-water system.

Therefore, force fitting these exponents and the constant to give a correlation consistent with previous work and applicable across the range of viscosities gives:

\[
k_L a = 2.2 \times 10^{-3} \left( \frac{P_T}{V_d} \right)^{0.6} \left( \frac{Q_G}{Q_G + Q_L} \right)^{0.55} \mu_L^{-0.18} \quad \text{(Eqn. 42)}
\]

\(R^2 = 0.401\)

for \(Q_G/Q_L \leq 1.0\) and \(\mu_L\) is in mPas.

Despite a much smaller value of \(R^2\), the fit to this correlation is similar to Equation 41, as seen in Figure 69. Therefore, due to consistencies with previous work (Z2, Z3 and Equations 1 and 2), Equation 42 may be used to estimate \(k_L a\). However, because of a better fit (giving a greater accuracy), demonstrated by the higher value of \(R^2\), Equation 41 is recommended for use in \(k_L a\) estimation for the design of a gas-liquid ejector to be used for gas contacting for liquid viscosities up to 10 mPas.

6.6 SPECIFIC INTERFACIAL AREA AND LIQUID-SIDE MASS TRANSFER COEFFICIENTS

Specific interfacial area was calculated using the overall Sauter mean diameter and overall gas holdup in the mixing tube, using Equation 17. Overall mass transfer coefficients were obtained as detailed in Section 6.5. By using these results, liquid-side mass transfer coefficient could be calculated. Table 16 shows limited results for \(a\) and \(k_L\).
Chapter 7 - CONCLUSIONS AND RECOMMENDATIONS

Experimental tests reported in this thesis were carried out in three types of systems: coalescing (air-water), non-coalescing (air-Magnesium Sulphate solution) and viscous (air-< 10 mPas CMC solution). The experiments in the coalescing system measured gas holdup and bubble size and the effects of gas and liquid flowrates on these parameters. This extended the previous work done in an air-water, gas-liquid ejector by Zhu (Z2-Z5). The tests in the non-coalescing and viscous fluids measured gas entrainment, ejector efficiency, bubble size, gas holdup and most importantly mass transfer and compared these parameters to the air-water system.

This thesis has therefore considerably extended the previous work carried out by Zhu (Z2-Z5) and other workers in this field (S3, C5, C6, D1, W1 and Z7). Measurements in an air-water system on bubble size and gas holdup have been achieved, producing valuable correlations for a downflow ejector. A series of experimental tests in systems other than coalescing: non-coalescing and viscous, enabled comparisons and a broader base for industrial practice. Again, correlations were developed for use in the design of a gas-liquid ejector to be used for gas-liquid contacting.

The conclusions and main findings for the above work are split into three sections and are summarised as follows:

7.1 AIR-WATER

- Numerous photographs were taken and analysed for several experimental conditions and from this, a minimum bubble count of 200 - 250 gave an accurate Sauter mean diameter. The information from these and the remaining bubble counts showed that the bubble size distributions in air-water were close to a Gaussian distribution.
• The local Sauter mean diameter was observed to increase as the two phase flow developed down the mixing tube due to coalescence of the gas bubbles.

• An increase in gas flowrate at constant liquid flowrate and an increase in liquid flowrate at constant gas-liquid flow ratio decreased local Sauter mean diameter.

• Overall measured gas holdup is generally 20 - 30\% higher than the flow ratio $Q_g/(Q_g + Q_l)$. This flow ratio has previously been used as an estimate of gas holdup and from the results in this thesis this is shown to be an inaccurate approximation.

• An increase in gas flowrate at constant liquid flowrate increased gas holdup which stabilised at $Q_g/Q_l = 1.5$ and appeared to correlate well with $Q_g/Q_l$, independent of $Q_L$.

• The calculated slip-velocities which were higher than would be expected from bubble rise were thought to be due to the high velocity of the jet at the early stage of the mixing tube.

7.2 EFFECT OF COALESCENCE INHIBITION

• Coalescence inhibition has negligible effect on ejector pumping efficiency and gas entrainment as well as gas holdup and slip-velocities in the mixing tube.

• As seen with air-water, local $d_{32}$ increases as the two phase flow develops down the mixing tube, however, the effect is not as significant. Compared to the coalescing system, the non-coalescing system gives more spherical and smaller bubbles and shows a less significant increase in $d_{32}$ down the mixing tube. Sauter mean diameter decreases with an increase in gas flowrate, decreases with
an increase in liquid flowrate in the early part of the mixing tube (axial position 1) and is unaffected by a change in liquid flowrate at constant $Q_g/Q_L$ further down the mixing tube (axial positions 2 & 3).

- The measured $k_L\alpha$ values of the non-coalescing system are not significantly different from those of coalescing systems. This is due to the counteracting effects of higher $\alpha$ but lower $k_L$. Correlations have been proposed.

### 7.3 EFFECT OF LIQUID VISCOSITY

- Liquid viscosity has a small effect on pumping efficiency and gas entrainment and up to a liquid viscosity of 6 mPas has negligible effect on gas holdup and slip-velocities in the mixing tube.

- Again, local $d_{32}$ increases slightly as the two phase flow develops down the mixing tube and compared to the water of 1 mPas, a higher viscosity of up to 10 mPas is less significant. Sauter mean diameter is seen to increase with an increase in gas flowrate and decrease with an increase in liquid flowrate at constant $Q_g/Q_L$.

- Increasing the liquid viscosity reduces the value of $k_L\alpha$ in the gas-liquid ejector. This is despite an increase in $\alpha$ due to a constant gas holdup and reduced Sauter mean diameter hence displaying there is a significant reduction in $k_L$. Correlations have been proposed to enable design of a gas-liquid ejector for gas-liquid contacting for liquid viscosities up to 10 mPas.
7.4 POTENTIAL FUTURE WORK

It is recommended that future work may consist of the following:

- Investigation of an increased concentration of Magnesium Sulphate solution to gain a deeper understanding of the effect of the salt on $k_e\alpha$ would be advantageous.

- For a non-coalescing system, were the effects observed due to the non-coalescing nature of the system alone or combined with the specific concentration of Magnesium Sulphate? Therefore, the use of a different salt to represent a non-coalescing system is also suggested as the current work may be system specific rather than a general representation of a non-coalescing system.

- Non-coalescing systems can also be represented by organic solutes as well as inorganic, giving even further scope for future work.

- A further investigation into the effect of viscosity is also recommended due to the limited and interesting observations in this work.

- The thesis covered a gas-liquid ejector in downflow orientation and the observed effects in coalescing, non-coalescing and viscous systems may vary for a different flow orientation which may lead to extensive future work.
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H1 Heijnen, J. and K. Van’t Riet
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H2 Ho, C.S., R.F. Baddour and M.J. Stalker
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H3 Ho, C.S., L.K. Ju, and C.T. Ho
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II. Fermentation media

J1 Jamialahmadi, M. and H. Müller - Steinhagen
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J2 Ju, L.K., E. Livio and C. S. Ho
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J3 Ju, L.K. and C.S. Ho
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Sorption characteristics of slot injectors and their dependency on the coalescence behaviour of the system
TABLE 6  Measured peak to peak times and calculated gas holdup for $Q_L = 2.1 \text{l}s^{-1}$: Air-water

<table>
<thead>
<tr>
<th>$Q_L$ (ls$^{-1}$)</th>
<th>$Q_d/Q_L$</th>
<th>$t$ (peak-peak) (sec)</th>
<th>Gas holdup (-)</th>
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<tbody>
<tr>
<td>2.1</td>
<td>0.48</td>
<td>0.258</td>
<td>0.422</td>
</tr>
<tr>
<td>0.48</td>
<td>0.257</td>
<td>0.425</td>
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<tr>
<td>0.48</td>
<td>0.251</td>
<td>0.450</td>
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</tr>
</tbody>
</table>

TABLE 7  Effect of axial position on $d_{32}$

<table>
<thead>
<tr>
<th>$Q_L$ (ls$^{-1}$)</th>
<th>$Q_d/Q_L$</th>
<th>Axial position (mm)</th>
<th>$d_{32}$ (mm)</th>
<th>$\sigma$ (mm)</th>
<th>No. of bubbles</th>
<th>$d_{32}$ (mm)</th>
<th>$\sigma$ (mm)</th>
<th>No. of bubbles</th>
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<tr>
<td>1.2</td>
<td>0.48</td>
<td>1</td>
<td>0.85</td>
<td>0.07</td>
<td>329</td>
<td>0.497</td>
<td>0.008</td>
<td>335</td>
</tr>
<tr>
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<td></td>
<td>2</td>
<td>1.77</td>
<td>0.11</td>
<td>213</td>
<td>0.503</td>
<td>0.012</td>
<td>344</td>
</tr>
<tr>
<td></td>
<td></td>
<td>3</td>
<td>2.11</td>
<td>0.12</td>
<td>241</td>
<td>0.627</td>
<td>0.032</td>
<td>499</td>
</tr>
<tr>
<td>2.1</td>
<td>0.48</td>
<td>1</td>
<td>0.534</td>
<td>0.07</td>
<td>318</td>
<td>0.372</td>
<td>0.033</td>
<td>449</td>
</tr>
<tr>
<td></td>
<td></td>
<td>2</td>
<td>1.22</td>
<td>0.07</td>
<td>465</td>
<td>0.54</td>
<td>0.028</td>
<td>464</td>
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<tr>
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<td>1.34</td>
<td>0.08</td>
<td>441</td>
<td>0.629</td>
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<td>423</td>
</tr>
<tr>
<td>2.7</td>
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<td>1</td>
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<td>0.029</td>
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<tr>
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<td>2</td>
<td>0.84</td>
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<td>0.08</td>
<td>2.11</td>
<td>0.621</td>
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TABLE 8  Effect of gas flowrate on $d_{32}$

<table>
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<tr>
<th>$Q_L$ (ls$^{-1}$)</th>
<th>$Q_g/Q_L$</th>
<th>Axial position</th>
<th>Water</th>
<th>Salt solution</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td></td>
<td>$d_{32}$ (mm)</td>
<td>$\sigma$ (mm)</td>
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<tr>
<td>2.1</td>
<td>0.24</td>
<td>3</td>
<td>1.34</td>
<td>0.08</td>
</tr>
<tr>
<td>0.48</td>
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<td>3</td>
<td>0.84</td>
<td>0.05</td>
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<td>0.71</td>
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<td>0.71</td>
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<tr>
<td>0.95</td>
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<td>3</td>
<td>1.10</td>
<td>0.09</td>
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TABLE 9  Effect of liquid flowrate on $d_{32}$ at constant $Q_d/Q_L$

<table>
<thead>
<tr>
<th>$Q_L$ (ls$^{-1}$)</th>
<th>$Q_d/Q_L$</th>
<th>Axial position</th>
<th>$d_{32}$ (mm)</th>
<th>$\sigma$ (mm)</th>
<th>No. of bubbles</th>
<th>$d_{32}$ (mm)</th>
<th>$\sigma$ (mm)</th>
<th>No. of bubbles</th>
</tr>
</thead>
<tbody>
<tr>
<td>1.2</td>
<td>0.48</td>
<td>1</td>
<td>0.85</td>
<td>0.07</td>
<td>329</td>
<td>0.497</td>
<td>0.008</td>
<td>335</td>
</tr>
<tr>
<td>2.1</td>
<td>0.48</td>
<td>1</td>
<td>0.534</td>
<td>0.07</td>
<td>318</td>
<td>0.372</td>
<td>0.003</td>
<td>449</td>
</tr>
<tr>
<td>2.7</td>
<td>0.48</td>
<td>1</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>0.345</td>
<td>0.029</td>
<td>363</td>
</tr>
<tr>
<td>1.2</td>
<td>0.48</td>
<td>2</td>
<td>1.77</td>
<td>0.11</td>
<td>213</td>
<td>0.503</td>
<td>0.012</td>
<td>344</td>
</tr>
<tr>
<td>2.1</td>
<td>0.48</td>
<td>2</td>
<td>1.22</td>
<td>0.07</td>
<td>465</td>
<td>0.54</td>
<td>0.028</td>
<td>464</td>
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<tr>
<td>2.7</td>
<td>0.48</td>
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<td>0.06</td>
<td>362</td>
<td>0.507</td>
<td>0.037</td>
<td>611</td>
</tr>
<tr>
<td>1.2</td>
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<td>3</td>
<td>2.11</td>
<td>0.12</td>
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<td>0.627</td>
<td>0.032</td>
<td>499</td>
</tr>
<tr>
<td>2.1</td>
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<td>3</td>
<td>1.34</td>
<td>0.08</td>
<td>441</td>
<td>0.629</td>
<td>0.012</td>
<td>423</td>
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<td>1.21</td>
<td>0.08</td>
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<td>0.036</td>
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</tr>
<tr>
<td>1.2</td>
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<td>Overall</td>
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<td>0.563</td>
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<td>Overall</td>
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<td></td>
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<tr>
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<td>Overall</td>
<td>-</td>
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<td></td>
<td></td>
</tr>
<tr>
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<td>3</td>
<td>1.10</td>
<td>0.09</td>
<td>339</td>
<td>0.514</td>
<td>0.025</td>
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TABLE 10  Calculated $d_{32}$, $\phi$ and $a$: air-water

<table>
<thead>
<tr>
<th>$Q_L$ (ls$^{-1}$)</th>
<th>$Q_d/Q_L$</th>
<th>Overall $d_{32}$ (mm)</th>
<th>$\phi$</th>
<th>$a$ (m$^{-1}$)</th>
<th>$k_i\alpha$ (s$^{-1}$)</th>
<th>$k_i$ (ms$^{-1}$)</th>
</tr>
</thead>
<tbody>
<tr>
<td>2.1</td>
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<td>1.26</td>
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<td>2048</td>
<td>3.3</td>
<td>0.0016</td>
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</table>
TABLE 11  Calculated $a$ and $k_L$; air-salt solution

<table>
<thead>
<tr>
<th>$Q_L$  ($\text{ls}^{-1}$)</th>
<th>$Q_{d}/Q_L$</th>
<th>Overall $d_{32}$ (mm)</th>
<th>$\phi$</th>
<th>$a$ (m$^3$)</th>
<th>$k_La$ (s$^{-1}$)</th>
<th>$k_L$ (ms$^{-1}$)</th>
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<td>0.0027</td>
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TABLE 12  10 mPas $d_{32}$ results

<table>
<thead>
<tr>
<th>$Q_L$ ($\text{ls}^{-1}$)</th>
<th>$Q_{d}/Q_L$</th>
<th>Axial position</th>
<th>$d_{32}$ (mm)</th>
<th>No. of bubbles</th>
<th>$d_{32}$ (mm)</th>
<th>No. of bubbles</th>
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<td>2.11</td>
<td>241</td>
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</table>

TABLE 13  Effect of axial position on $d_{32}$

<table>
<thead>
<tr>
<th>$Q_L$ ($\text{ls}^{-1}$)</th>
<th>$Q_{d}/Q_L$</th>
<th>Axial position</th>
<th>$d_{32}$ (mm)</th>
<th>$\sigma$ (mm)</th>
<th>No. of bubbles</th>
<th>$d_{32}$ (mm)</th>
<th>$\sigma$ (mm)</th>
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<td></td>
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<td>517</td>
<td>1.34</td>
<td>0.08</td>
<td>441</td>
</tr>
<tr>
<td>overall</td>
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### TABLE 14  
**Effect of gas flowrate on $d_{32}$**

<table>
<thead>
<tr>
<th>$Q_L$ (ls$^{-1}$)</th>
<th>$Q_d/Q_L$</th>
<th>Axial position</th>
<th>$d_{32}$ (mm)</th>
<th>$\sigma$ (mm)</th>
<th>No. of bubbles</th>
<th>$d_{32}$ (mm)</th>
<th>$\sigma$ (mm)</th>
<th>No. of bubbles</th>
</tr>
</thead>
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<td>0.435</td>
<td>0.044</td>
<td>517</td>
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<td>1.34</td>
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<td>0.71</td>
<td>0.19</td>
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### TABLE 15  
**Effect of liquid flowrate at constant $Q_d/Q_L$**

<table>
<thead>
<tr>
<th>$Q_L$ (ls$^{-1}$)</th>
<th>$Q_d/Q_L$</th>
<th>Axial position</th>
<th>$d_{32}$ (mm)</th>
<th>$\sigma$ (mm)</th>
<th>No. of bubbles</th>
<th>$d_{32}$ (mm)</th>
<th>$\sigma$ (mm)</th>
<th>No. of bubbles</th>
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<td>0.435</td>
<td>0.044</td>
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<tr>
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</tr>
</tbody>
</table>

### TABLE 16  
**Calculated $a$ and $k_l$: 6mPas CMC solution**

<table>
<thead>
<tr>
<th>$Q_L$ (ls$^{-1}$)</th>
<th>$Q_d/Q_L$</th>
<th>Overall $d_{32}$ (mm)</th>
<th>$\phi$</th>
<th>$\alpha$ (m$^{-1}$)</th>
<th>$k_l\alpha$ (s$^{-1}$)</th>
<th>$k_l$ (ms$^{-1}$)</th>
</tr>
</thead>
<tbody>
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<td>2.1</td>
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</table>
Figure 1: Effect of molar concentration on coalescence frequency
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APPENDIX A
DEFINITION OF SWIRL NUMBER OF A LIQUID JET

Refer to Figure A1.

The presence of a spinner upstream of the ejector nozzle imparts a tangential velocity component to the liquid flow. This swirling characteristic of the liquid jet may be described with a swirl number, defined as the ratio of tangential to axial velocity of the jet:

\[ Sw = \frac{r_n \omega_n}{u_{z,0}} \]  

(Eqn. A1)

where \( r_n \) is nozzle radius and \( u_{z,0} \) the axial velocity at the nozzle. \( Sw \) can be related to the spinner geometry and nozzle diameter. The following summarises the derivation for this relationship made in previous work (P1).

Assuming that angular momentum is conserved about the pipe centerline when the swirl was carried from the spinner to the nozzle outlet,

\[ r_n^2 \omega_n = r_0^2 \omega_0 \]  

(Eqn. A2)

The time that the liquid spends on travelling along the spinner can be expressed as:

\[ t_s = \frac{l_s}{u_{z,0}} = \frac{\theta}{\omega_0} \]  

(Eqn. A3)

or:

\[ \omega_0 = \frac{\theta u_{z,0}}{l_s} \]  

(Eqn. A4)
Combining Equations (A1), (A2) and (A4) gives:

\[ Sw = \frac{\theta r_n}{l_s} \frac{r_0^2 u_{z,0}}{r_n^2 u_{z,n}} \]  
(Eqn. A5)

Since the second group equals 1, then \( Sw \) is given by:

\[ Sw = \frac{\theta r_n}{l_s} \]  
(Eqn. A6)
Figure A1: Spinner geometries

θ: $90^\circ, 135^\circ, 225^\circ, 360^\circ$
APPENDIX B

PERFORMANCE PARAMETERS

B.1 Power

A gas-liquid ejector utilizes the liquid kinetic energy to entrain gas and achieve two-phase mixing. The power consumed in the form of liquid pressure drop can be expressed as follows:

$$P_T = (p_s - p_2) Q_L + \frac{1}{2} \rho Q_L (u_j^2 - u_{l,2}^2)$$  \hspace{1cm} (Eqn. B1)

The gas power gain is expressed by:

$$P_G = Q_G \frac{p_s^1 + p_2}{2} \ln\left(\frac{p_2}{p_s^1}\right)$$  \hspace{1cm} (Eqn. B2)

where $Q_G$ refers to the average volumetric gas flowrate in the mixing tube and $u_j$ the liquid jet velocity. $p_s^1$ is the gas chamber pressure after taking into account the elevation and can be calculated by:

$$p_s^1 = p_s + \rho g L_m (1 - \phi)$$  \hspace{1cm} (Eqn. B3)

where $\phi$ is the volume fraction of gas in the mixing tube (gas holdup). The liquid downstream velocity, $u_{l,2}$, is given by:

$$u_{l,2} = \frac{Q_L}{A_m (1 - \phi)}$$  \hspace{1cm} (Eqn. B4)

where $A_m$ is the mixing tube cross-section area.

In the absence of gas holdup data, values of flow ratio, $Q_g/(Q_g + Q_L)$ was used as an approximation for $\phi$. 
The power used for mixing and friction is the difference between $P_T$ and $P_G$, i.e.

$$P_m = P_T - P_G \quad \text{(Eqn. B5)}$$

**B.2 Entrainment**

Entrainment performance of ejectors can usually be described by the relationship between gas flowrate and the pressure in the gas chamber ($P_g$). For a given gas chamber pressure, the higher the gas flowrate (more gas entrained) the better the entrainment performance.

The ejector's characteristic of pressure rise of the gas phase can also be described by a pumping efficiency, which is defined as the ratio of the power gained by gas phase to the liquid power input:

$$\eta = \frac{P_g}{P_T} \quad \text{(Eqn. B6)}$$

**B.3 Mass Transfer**

As in most other gas-liquid systems, oxygen transfer between water and nitrogen is liquid-film controlling (oxygen not easily soluble in water). In this case the liquid-side mass transfer coefficient, $k_L$, may replace the overall coefficient and the mass transfer performance of the equipment can be evaluated by $k_L a$, a product of $k_L$ and specific interfacial area.

Assuming that the system is in plug flow, $k_L a$ can be calculated as follows:

$$k_L a = \frac{Q_L (C_1 - C_2)}{\Delta C V_d} \quad \text{(Eqn. B7)}$$
$k_a$ is a function of temperature. Given the inevitable small variations in temperature during experiments, the values of $k_a$ obtained were corrected to 20.0°C with the following equation (S4).

$$k_{a_{20}} = \frac{k_{a_1}}{0.467 + 0.0267 \, t} \quad \text{(Eqn. B8)}$$

where $t$ is in °C. It was assumed that this relationship, obtained for air-water systems, also holds for the nitrogen-water system.
APPENDIX C

DERIVATION OF EQUATION USED TO CALCULATE GAS HOLDUP

Gas holdup was calculated using the following equation:

\[
\phi = 1 - \frac{\left( t_{g\cdot t} - \frac{V_{Li}}{Q_L} \right)}{\left( t_i - \frac{V_{Li}}{Q_L} \right)}
\]  
(Eqn. C1)

This equation can be derived as follows:

The mean residence time in the absence of gas, which was taken to be equal to the peak-peak time between the two conductivity probe traces is:

\[
t_i = \frac{V_{Li}}{Q_L} + \frac{V_2}{Q_L}
\]  
(Eqn. C2)

The mean residence time in the presence of gas is:

\[
t_{g\cdot t} = \frac{V_{Li}}{Q_L} + \frac{V_2 (1 - \phi)}{Q_L}
\]  
(Eqn. C3)

Combining Equations C2 and C3 gives:

\[
t_{g\cdot t} = \frac{V_{Li}}{Q_L} + (1 - \phi) \left( t_i - \frac{V_{Li}}{Q_L} \right)
\]  
(Eqn. C4)

This gives the equation for holdup as expressed in Equations 15 and C1.
APPENDIX D
DERIVATION OF EQUATION USED TO CALCULATE SLIP-VELOCITY

Slip-velocity in the mixing tube can be calculated using the following equation:

\[ v_{slip} = v_{slip} \left( \frac{\phi}{Q_g(Q_g + Q_l)} - 1 \right) \quad \text{(Eqn. D1)} \]

This equation can be derived as follows:

Slip-velocity is equal to the difference in liquid and gas velocities in the mixing tube:

\[ v_{slip} = v_L - v_G \quad \text{(Eqn. D2)} \]

\[ = \frac{Q_L}{A(1 - \phi)} - \frac{Q_G}{A\phi} \quad \text{(Eqn. D3)} \]

\[ = \frac{Q_L \phi - Q_G (1 - \phi)}{A\phi (1 - \phi)} \quad \text{(Eqn. D4)} \]

\[ = \frac{Q_G}{A} \left( \frac{Q_L}{Q_G} \phi - 1 + \phi \right) \quad \text{(Eqn. D5)} \]

\[ = v_{sg} \left( \frac{\phi}{Q_g(Q_g + Q_l)} - 1 \right) \quad \text{(Eqn. D6)} \]

This gives the equation for slip-velocity as expressed in Equations 22 and D1.