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Gas-Liquid Mixing in Dual Agitated Vessels in the Heterogeneous Regime

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Abstract

Gas-liquid multi-phase processes are widely used for reactions such as oxidation and hydrogenation. There is a trend for such processes to increase the productivity of the reactions, one method of which is to increase the gas flow rate into the vessel. This means that it is important to understand how these reactors perform as high gas flow rates occurs well into the heterogeneous regime. This paper investigates the mixing performance for the dual axial radial agitated vessel of 0.61 m in diameter. 6 blade disk turbine (Rushton turbine) below a 6 Mixed flow Up-pumping and down-pumping have been studied at very high superficial gas velocities to understand the flow regimes operating at industrial conditions. Electrical Resistance Tomography have been used to produce the 3D images using Matlab, along with analysing the mixing parameters such as Power characteristics, gas hold-up and dynamic gas disengagement. Minimal difference between the two configurations have been reported in terms of gas hold-up, however with the choice of upward and downward pumping impeller power characteristics show significant difference at very high gas flow rates. Also at these high superficial gas velocities, this report introduces a 3rd bubble class, as seen in dynamic gas disengagement experiments, which corresponds to very large slugs of gas.

Highlights

- At high gas flow rates, a maximum hold-up is observed
- Power for 6MFD/6MFU problematic at high gas flow rates
- The gas liquid hold-up results show minimal difference between the two configurations
- Large slugs observed as a third bubble classes at very high gas flow rates

Keywords

Gas-liquid mixing; Gas Hold-Up; Gassed Power; Heterogeneous regime; Axial-radial dual system; Electrical Resistance Tomography

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Nomenclature

Latin symbols	
C	Clearance [m]
D	Agitator diameter [m]
H	Height of liquid [m]
h_{db}	Dished base height [m]
M	Torque [Nm]
N	Agitator speed [rps]
NLL	Normal Liquid level [m]
P	Power [W]
Q_g	Volumetric gas flow rate [m^3 s^{-1}]
r	Radius of tank [m]
T	Tank diameter [m]
T	Time [s]
u_b	Bubble rise velocity [cm s^{-1}]
V	Volume [m^3]
V_D	Volume of dispersion [m^3]
V_L	Volume of liquid [m^3]
V_G	Volume of gas [m^3]
v_s	Superficial gas velocity [m s^{-1}
Greek symbols	
ε_G	Gas Hold-up [%]
P_T	Specific total energy input per liquid mass [W kg^{-1}]
Σ	Conductivity (normalised) [mS cm^{-1}]
ρ	Density of fluid [kg m^3]
Dimensionless numbers	
Fl_G	Gas flow number
Fr	Froude number
P_g/P_u	Power gas factor (Gassed Power number/Ungassed Power number)
P_o	Power number
Re	Reynolds number
Subsideces	
a	Multiplicity factor
b	Exponent of ε_T

c	Exponent of v_s
Cd	Completely dispersed
F	Flooded
G_p	Gas phase
L	Loaded
L_p	Liquid phase
l	Large (bubbles)
R	Recirculation
s	Small (bubbles)
sl	Slug
Abbreviations	
6BDT	6 Blade Disc Turbine (Rushton Turbine)
6MFD	6 Mixed Flow Downward pumping
6MFU	6 Mixed Flow Upward pumping
DAS	Data Acquisition System
DGD	Dynamic Gas Disengagement
ERT	Electrical Resistance Tomography
FEM	Finite Element Method
ITS	Industrial Tomography System
NaCl	Sodium Chloride
RPM	Revolutions Per Minute
VVM	Volume of gas per volume of liquid per minute

1. Introduction

Mechanical agitation in vessels is among the most commonly used method of mixing in the chemical process industry. To achieve desired process results, mixing is important to reduce inhomogeneity of single or multiple phases. Multiphase mixing including gas-liquid, liquid-liquid, solid-liquid, and gas-solid-liquid are important unit operations used in major industries. Gas-liquid is one of the most important multiphase mixing processes, used in oxidation, hydrogenation, and biological aerobic fermentation etc. Power draw is an important variable in process mixing industry as it defines the energy requirement for the movement of fluid within a tank by mechanical agitation. The cost associated with power draw is substantial as it contributes to the overall operational cost of industrial plant. With gas-liquid systems, the gas is often introduced to the vessel at high pressure to reduce the volumetric flow rate and to increase the driving force for gas-liquid mass transfer. This compression also contributes to the operational cost. This cost can be reduced by reducing the pressure of the gas phase, which results in increasing the superficial gas velocity, which can effect key operating parameters for the system.

The scale-up for gas liquid mixing is often at geometric similarity and constant VVM (volume of gas per volume of liquid per minute). This results in a linear increase of superficial gas velocity (V_S) with vessel diameter. As the power input by the gas is proportional to superficial gas velocity a lot of mixing occurs by gas at large scale and at very large scale bubble columns are often used (effect of agitator power is negligible). A lot of work has been done in past for gas liquid mixing (Cooke, 2005; Hari-Prajitno et al., 1998; Nienow, 1998; Vrábel et al., 2000), but this earlier work mainly involved lower superficial gas velocities as scale-down has generally been done on gas demand (VVM). However this small-scale work does not reflect the increased energy input of the gas on the large scale and is only indicative of operation in the homogeneous (bubble regime) whereas large scale operation is generally in the heterogeneous (churn-turbulent) regime. In a recent study (Nauha et al., 2015), for the same reasons mentioned it was reported that scaling up by constant VVM will produce different hydrodynamic conditions in different scales and is therefore not recommended. The change in key variables on scale-up can be seen in Figure 1. With the economy of large scale operation, the need to understand the hydrodynamics of gas-liquid mixing at high gas superficial velocities is becoming more apparent; as is the need to investigate gas liquid mixing at realistic gas flow rates extending well into the heterogeneous regime.

Apart from homogeneous and heterogeneous regime, the gas liquid mixing flow regimes can also be identified as flooded, loaded and completely dispersed. These are the three main regimes which require much attention when conducting gas liquid mixing (Grenville and Nienow, 2004). Figure 2 depicts the flow pattern of the three regimes for single Rushton Turbine; increasing agitator speed at constant gas flow rate gives completely dispersed flow which is the most favourable regime. However, the diagrams are only truly representative of low superficial gas velocities associated with small mixing vessels. At high superficial gas velocities, the gas energy is sufficient to completely disperse the gas, even in what should be flooded conditions. The boundaries between the key flow regimes can be given by the equations below (Lee and Dudukovic, 2014; Nienow et al., 1985)

$$(\mathrm{Fl}_G)_{F\rightarrow L} = 30\left(\frac{D}{T}\right)^{3.5}\mathrm{Fr} \tag{1}$$

$$(\mathrm{Fl}_G)_{L\rightarrow CD} = 0.2\frac{D}{T}^{0.5}\mathrm{Fr}^{0.5} \tag{2}$$

$$(\mathrm{Fl}_G)_{\rightarrow R} = 13\left(\frac{D}{T}\right)^{5}\mathrm{Fr}^{2} \tag{3}$$

Where $F\rightarrow L$ is the transition from flooded to loaded, $L\rightarrow CD$ is the transition from loaded to completely dispersed, and $\rightarrow R$ is the transition to intense recirculation. The dimensionless flow regime map created by equations (1)-(3) is shown in Figure 3. The transition between the flooding and loading regime is the most important, as operation under flooding conditions are not desirable

(Cooke, 2005). Gezork et al. (2000) reported that the observation of flooding to loading and to completely dispersed transition shows good agreement with the above empirical equations for gas velocity up to 20 VVM for Rushton Turbine in a 0.29 m diameter vessel. These equations are scale independent; however they have not been tested at scales above 1.83 m vessel diameter (Cooke, 2005).

An important design parameter in gas liquid agitated vessels is the gas hold-up as this determines the total vessel volume, and is important for both operational and modelling purposes. Smith et al. (1977) and Davies (1986), along with other workers (Nienow et al. 1997) have developed correlations to fit air-water gas hold-up data with specific power input and superficial gas velocity given by the form of equation (4).

$$\varepsilon_G = a{P_T}^b v_s^c \tag{4}$$

ε_G is the gas hold-up, P_T is the specific power input (i.e from both the agitator and the gas) per liquid mass, in W kg^{-1}, v_s is the superficial gas velocity in m s^{-1} and *a*, *b*, and *c* are multiplicity factors and exponents. The *a*, *b* and *c* values can be obtained using the multi linear least square regression method, for example Cooke (2005) used this method to determine the values as 76.6, 0.39, and 0.56 respectively to fit gas hold-up data for v_s values ranging from 0.0017 to 0.127 m s^{-1}. These factors only work for the given system, hence cannot be applied universally (Nienow et al., 1997). The value of *b* and *c* varies from 0.2 – 0.7 (Moucha et al., 2003; Nienow et al., 1997) in literature, for example *Table 1* shows some reported values for *c* with different v_s. Nauha et al. (2015) actually produce a double fit for the hold-up across superficial gas velocity, which defines clear difference between the two regimes. The rate of increase in hold-up reduces at the higher superficial gas velocities; however, even with this the hold-up is over predicted at the higher superficial gas velocities.

Table 1. c values reported in literature for different superficial gas velocities.

c	v_s (m s^{-1})	Author
0.56	0.0017 to 0.127	(Cooke, 2005)
0.776	up to 0.048	(Bujalski et al., 1988)
0.67	0.019	(Chapman, 1981)
0.4	0.005 - 0.05	(Smith et al., 1977)

The use of impellers is another important parameter for gas liquid mixing. The use of both radial and axial impellers is reported to be more efficient compared to just multiple radial impellers in a stirred tank (Xie et al., 2014). Sardeing et al. (2004) reported that the single axial impellers in the upward pumping direction give better performance in gas liquid dispersion than the single downward pumping. In the same report, Rushton Turbine showed higher gas hold-up capacity but higher energy requirement, thus it was reported as economically inefficient. Hudcova et al. (1989) studied dual impeller Rushton Turbines (D/T = 1/3) in a 0.56 m diameter vessel and was the first time where each impeller was measured individually by strain gauging/telemetry method that allowed to link the power measurements with bulk flow patterns and cavity structure. They reported that aerated conditions created more complexity as different gas filled cavity structures were observed, upper impeller was reported to draw more power and much less prone to flooding conditions. However, in later studies on dual Rushton Turbines, Taghavi et al. (2011) found lower impeller to draw more power in aerated conditions. It is important to mention here that, every system behaves differently even if they have same impellers in common, for instance spacing between impellers is another important factor that can have a significant impact on power number and flow behaviour. Hudcova et al. (1989) studied various spacings between the impellers and stated that for a given speed, the individual power numbers for the dual impellers are a function of spacing. Thus one could conclude that changing impeller spacing results in different flow patterns; as from the same study they managed to safely conclude that increasing spacing with one vessel diameter between the impellers maximises the potential for mass transfer. It was also reported that for one vessel diameter spacing, the two impellers behave independently of each other under unaerated conditions, however with aerated conditions the lower impeller behaves much as a single impeller and therefore the design equations for single impeller can be applied. Similarly, Taghavi et al. (2011) also reported that the bottom impeller behaves similar to a single impeller system. This paper studies two configurations of the dual impellers; a radial (Rushton Turbine) and an axial (Mixed Flow upward/downward) as depicted by Figure 4, the same figure also show the flow pattern effect of increasing spacing between the impellers. Although the distance between the impellers in this study is kept constant throughout i.e. less than one vessel diameter, it is to note that the ratio of the distance between the impellers to height is almost the same as used by Hudcova et al. Regardless the two studies can’t be compared directly as they use different upper impeller therefore it is not necessary that they will have same effect.

Dynamic gas disengagement allows the examination of the performance of the gas liquid mixing systems by allowing identification of the bubble sizes. Schumpe and Grund (1986) outlined a method to model 2 bubble classes defined as large and small, which can be measured from disengagement data. This is important to bubble column studies, as now the use of bubble columns in heterogeneous regime in industries is becoming more common (Kantarci et al., 2005). The design and operation of the bubble column heavily relies on global and local properties such as the hold-up (ε_G), bubble rise velocity (u_b), local void fraction ($\varepsilon_{G,local}$) and bubble size distributions. A few attempts have been made to investigate heterogeneous regime where more than two bubble classes have been discovered with this technique (Hashemi et al., 2016; Yang et al., 2010). The third bubble class is usually referred as slugs which is the largest bubble class in terms of size. From the bubble size distribution analysis it has been reported in literature that large bubbles occur above the transition superficial gas velocity and that the transition from homogeneous to heterogeneous occurs slowly (Yang et al., 2010). The homogeneous to heterogeneous transition point has been reported at 0.035 m s^{-1} (Schumpe and Grund, 1986), 0.03 m s^{-1} (Besagni and Inzoli, 2016; Nauha et al., 2015) and, 0.032-

0.056 m s^{-1} over a range of liquid viscosity from 32.52 to 1.02 mPa s respectively (Yang et al., 2010). Yang et al. (2010) showed increasing liquid viscosity had a clear impact on the reduction of the transition superficial gas velocity. Similarly, (Gezork et al., 2000) reported that at the highest agitator speed the transition from homogeneous to heterogeneous was found to occur at lower superficial gas velocity values in coalescence-inhibited solutions than in the coalescing water. Again, factors such as additives may also effect the flow behaviours significantly. Cooke et al. (2008) showed that salt addition reduces the bubble size as the small bubble class rise velocity decreased to 1.2 cm s^{-1} from 10.8 cm s^{-1}. It has been reported in literature (Besagni and Inzoli, 2016; Schumpe and Grund, 1986) that the rise velocity of the large bubble class increases with increasing gas flowrate and their contribution to the total holdup, whereas small bubble class generally show a constant effect on the rise velocity. Gezork et al. (2000) also reported the transition point at 0.03 m s^{-1} and that the bubble size increases slightly until slugs (very large bubbles) are formed. These slugs are presumed to increase in size until their velocity is so great that a maximum gas hold-up is achieved (Nauha et al., 2015). The bubble size is usually determined from the bubble rise velocity and varies with different systems, as in literature a wide range has been reported; small bubble class ranges from 1.2 – 21 cm s^{-1} and large bubble class from 17 to 160 cm s^{-1} (Besagni and Inzoli, 2016; Cooke et al., 2008; Schumpe and Grund, 1986). There are many other studies in literature for bubble columns but limited to homogeneous regime, the focus should rather be on heterogeneous regime as multiphase bubble columns are more commonly used in industries (Kantarci et al., 2005).

Electrical resistance tomography (ERT) is an imaging technique that has been greatly developed since its first industrial application was reported in the 1980s (York, 2001). The technique has multiple advantages over other imaging techniques, such as high speed, low cost, non-intrusive, and no radiation hazard. Thus ERT has become one of the most promising tomography techniques in monitoring and analysing various industrial flow processes (Aw et al., 2014; Rodgers et al., 2011; Yenjaichon et al., 2011). The fundamental of ERT is to take numbers of surface voltage measurements from multiple electrodes mounted in an arbitrary geometric pattern, generating an image showing distribution of electrical resistivity or conductivity. ERT is an established useful tool for estimating the gas hold-up in any location inside the vessel based on conductivity measurements. Fransolet et al. (2001) compared gas hold-up data obtained by pressure transducers and optical probe to ERT method and found ERT as a potential tool for providing quantitative and qualitative analysis. Razzak et al. (2009) found a good agreement between ERT, pressure transducer and optical fibre technique for solid liquid mixing systems, and reported the exceptional capabilities of ERT in determining the local, cross sectional average of solid phase hold-up and distribution. ERT has been used with Dynamic gas disengagement (DGD) studies to analyse the bubble rise velocity, bubble size, and gas holdup for gas liquid mixing systems (Babaei et al., 2015; Fransolet et al., 2005; Hashemi et al., 2016; Jin et al., 2007). It is evident that ERT is a reliable tool and can provide valuable findings to gas liquid mixing systems. The measured data from ERT can be processed in MATLAB to provide 3-D image reconstruction, a detailed description of the method is discussed in section 2.3 and can also be found in literature (Adler and Lionheart, 2006; Lionheart et al., 1999; Polydorides and Lionheart, 2002; Ren et al., 2017).

This paper investigates the gas liquid mixing system of air and water; it examines the power data for two configurations, a 6 blade disk turbine (Rushton turbine) with a 6 Mixed flow down-pumping impeller and a 6 blade disk turbine with a 6 Mixed flow up-pumping impeller, at different agitation speeds and gas flow rates. For the first time, these set of agitator configurations are investigated at high superficial gas velocities, which allows studying the flow regimes at superficial gas velocities similar to industrial conditions. Moreover, flow maps have been created to further analyse the flow regimes for flooded, loaded, and completely dispersed conditions; this is important as it can help in determining the optimal mixing conditions for gas liquid systems. Gas hold-up is also investigated and compared to literature using theoretical equations, with a new equation suggested which works for the full range of superficial gas velocities. As well as quantitative, qualitative analyses have also been made using 3D ERT images. At these high superficial gas velocities, this paper introduces a 3rd bubble class, as seen in dynamic gas disengagement experiments, which corresponds to very large slugs of gas.

2. Material and Method

2.1. Experimental Set up

The equipment consisted of a Perspex cylindrical vessel (0.61m diameter) with a ring air sparger installed in a way that the air flows from the underside through 20 mm holes. The vessel had a dished base with the cylindrical part fitted inside a square jacket for the purpose of a distortion free viewing window for flow visualisation. The agitator shaft was continuous and fitted into a PTFE (Polytetrafluoroethylene) bearing located in the centre of the dished base. A Ferro-magnetic proximity sensor coupled to a compact Micro 48 tachometer was used to monitor the shaft rotational speed. Astech Rotary Telemetry System was used to transmit the sensor data from rotating components to a stationary demodulator (receiver) using inductive coupling. The sensor data included torque, strain, and temperature. The tachometer input was connected to a computer to enable the calculation of shaft power. The demodulator was connected to a Pico logger data acquisition unit which was connected to the computer for the transmission of data into the Pico logger software. Dual torque strain gauges with power pick loop were attached to the shaft to measure the shaft power, one pair above the top agitator and one pair between the two agitators, allowing simultaneous measurement of both agitators. The temperature of the vessel was monitored using a temperature probe inserted via the analysis block. The other end was connected to Pico PT-104 temperature unit, which was connected to computer for the transmission of data measurements onto the Pico Logger software. The shaft was calibrated using weights and a lever arm.

Two agitators as shown by Figure 5 were used together; Rushton Turbine (RT) also known as 6 Blade Disc Turbine (6BDT) at the bottom and Mixed Flow Upward/Downward Pumping (6MFU/6MFD) at top. The diameter for both agitators was 0.3048 m; standard for 6BDT and across flats, not tip to tip for the 45^{o} pitch blade 6MF. The agitator dimensions along with other apparatus dimensions are listed in *Table 2*. Rotating shaft clockwise and anticlockwise provided different combination of agitators i.e. 6BDT plus 6MFD and 6BDT plus 6MFU respectively. The air sparge ring was placed under the bottom agitator. The configuration shown by Figure 6 was used to investigate gas-liquid mixing parameters for both sets of agitator combinations i.e. 6BDT+6MFD and 6BDT+6MFU (clockwise and anticlockwise respectively). The experiment conditions varied from 0.0029 to 0.42 m s^{-1} superficial gas velocity with agitator speed range from 30 to 300 rpm.

Table 2. Key measurements for the system used.

Dimension	6BDT	6MF
Tank diameter (T) / m	0.6096	
Cross sectional area / m^2	0.29	
Height of liquid (H) / m	0.7625	
Dispersion Volume / m^3	0.208	
Sparge pipe diameter / m	0.02	
Sparger clearance from base / m	0.08	
Diameter (D) / m	0.3048	0.3048
Blade length / m	0.076	0.12
Blade width / m	0.06	0.085
Clearance (c) / m	0.23	0.53
Pitch / ^{o}	90	45

2.2. Power Measurement

For the ungassed power numbers, the tank was filled up to 1 metre to prevent surface aeration, this level change had no effect on the power measured. The gassed power was collected at the true normal liquid level (NLL) with the corresponding gas flow rate. The power numbers were obtained from the raw voltage data collected from the Astech unit and recorded in PicoLog Data Logging software. The voltage measurement for all conditions were averaged over 120 seconds. The voltage difference between the running voltage and the base line (zero rpm) is multiplied by the strain gauge factor to calculate torque (M). The strain gauge factor was calculated by applying known loads onto the shaft before installation into the vessel. The shaft is installed with 2 sets of strain gauges, one above both agitators, and one in between the agitators; this allowed simultaneous measurement of the torque applied on each agitator. The torque measurements are then used to calculate agitator power by the following equation,

$$P = 2\pi N\mathrm{M} \tag{5}$$

The equation for Power number, Po, is given by

$$P_o = \frac{P}{\rho N^3 D^5} \tag{6}$$

This means that for all the data collected a power number can be calculated for each condition as,

$$P_o = \frac{2\pi NM}{\rho N^3 D^5} = \frac{2\pi M}{\rho N^2 D^5} \tag{7}$$

2.3. ERT Gas-Liquid Hold-up Measurements

The vessel was installed with 8 rings of 16 equally spaced electrode (3 cm high and 2 cm wide) arrays in a baffle cage configuration, however only 7 planes were used as the liquid level covered up to 7 planes. The electrodes were connected to the Data Acquisition System (DAS) which was connected to computer. The Industrial Tomography System, ITS P2000 software (ITS P2+ v7.3) was used to obtain voltage data generated from the electrodes. The DAS settings on ITS P2000 software are shown by

Table 3.

Table 3. ERT DAS Settings used for Gas Liquid Holdup

ERT DAS Settings	
Sampling time interval	200 ms

Number of sensing planes	7
Injection current	15 mA
Frequency of current	9600 Hz
Samples per frame	8
Delay cycles	5
Frames per download	1

The ITS P2000 software offers normal adjacent measurement strategy where the input current is injected over pairs of adjacent electrodes and the voltage is measured on pairs of adjacent electrodes on the same plane. A reference measurement was taken with just water agitated at 60 rpm. This reference is essential as it is required for the means of comparison and is used for every run for that condition for consistency. Measurements were collected for different gas flow rates at 180 and 300 rpm. The measured data were processed in MATLAB for 3-D image reconstruction. A 3-D vessel model was generated in Electrical Impedance Tomography and Diffuse Optical Tomography Reconstruction Software (EIDORS), which is a MATLAB toolkit for 2D and 3D ERT, providing accurate ERT sensor and electrode modelling using finite element method (FEM) to solve forward problem (Adler and Lionheart, 2006; Polydorides and Lionheart, 2002; Vauhkonen et al., 1999). The vessel model consists of 7 rings of 16 electrodes, fitting with two disc agitators in the centre and four blades at side, regarding to the actual geometry. The imaging region was divided into 17979 elements. However, a 7 ring of 16-electrode ERT sensor only gives 728 independent measurements and need to solve 17979 unknowns, which is ill-posed and ill-condition. To solve this problem, L-curve Tikhonov regularization method has been employed for image reconstruction in this paper (as outlined in Daisy paper). Tikhonov regularization method is used to determine a set of solution using prior constraint information and then to choose one solution from this set. The success of a regularization method is dependent to the selection of the regularization parameter, because a too small regularization parameter value would cause a noisy image while a too large value will make the solution over regularized and not fit the measured voltage data. L-curve method is to plot the size of the regularized solution versus the size of the corresponding residual, which would form an ‘L’ shape, and to optimise the regularization parameter by finding the corner of the ‘L’. For more details, see literatures (Hansen, 2007; Hansen and O’Leary, 1993). To avoid noise, 3-D images showing the conductivity distribution were reconstructed in MATLAB by averaging 100 frames of voltage data at the same condition.

2.4. Gas hold-up measurements

The gas-liquid hold-up is the rise in liquid surface level at aerated conditions. The difference between the aerated and unaerated level measured by eye view is used to calculate the gas hold-up volume. The volume of agitator, shaft and pipes inside the vessel are not taken into account. The following equation are used to calculate the gas hold-up volume,

$$\varepsilon_G = \frac{V_G}{V_D} 100 \tag{8}$$

$$V_D = V_G + V_L \tag{9}$$

$$V_G = \pi r^2 \, (aerated\ level - unaerated\ level) \tag{10}$$

$$V_L = \frac{2}{3} r^2 \pi h_{db} + r^2 \pi \, (unaerated\ level - h_{db}) \tag{11}$$

where V_G is the volume of gas, V_L is the volume of liquid, V_D is the dispersion volume and h_{db} is dished base height. The gas hold-up was investigated in the range of 0.0029 to 0.42 m s^{-1} superficial gas velocity and 30 to 300 rpm agitation speed. 500 ml of 0.86 molar solution of NaCl (50 g l^{-1}) was added to the vessel (much lower than the concentration of NaCl required to significantly affect coalescence) to increase the contrast between the liquid conductivity and the air. The average temperature was also measured for every condition. These conditions were investigated for both set of agitators; 6BDT+6MFD and 6BDT+6MFU (clockwise and anticlockwise respectively).

2.5. Dynamic Gas Disengagement

The dynamic gas disengagement data was taken to estimate the bubble rise velocity of the different bubble classes. The method used was proposed by Sriram and Mann (1977), which involves measuring the liquid level drop against time after simultaneously switching off the agitator and gas supply. A Canon 6D DSLR camera fixed on a tripod was used to record the level drop. The tripod was placed and position was marked for consistency. In addition, the level seen from a parallel position was calibrated against level seen on video to ensure accuracy. The camera view was set around the NLL level so that the complete level drop is captured. The experiment conditions were investigated across 0.0029 to 0.42 m/s superficial gas velocity at two agitation rates, 180 and 300 rpm. The software 'ffmpeg' (FFmpeg, 2017) was used to split the video in series of still frames, allowing 22 frames per second. Each frame was analysed by finding the height level, which was plotted against time to determine the rate of level drop, an example is shown by Figure 7.

The method yields the hold-up fractions of the different size bubbles, and allows calculation of the bubble rise velocities using the method described by Schumpe and Grund (1986). However, for these results the model needed to be extended to 3 bubble classes as at the extreme conditions used, most of the results had 3 clear classes rather than 2. The Schumpe and Grund (1986) model is modified which is illustrated by Figure 8 and 9. The gas hold-up, superficial gas velocity and bubble rise velocities of the three bubble classes can be determined by the following equations represented in

Table 4.

Table 4. Key equations for the gas disengagement analysis for 3 bubble classes.

Gas Hold-up		
Slugs, $\varepsilon_{G_{sl}}$	$(H_0 - H_{13})/H_0$	(12)
Large, ε_{G_l}	$(H_{13} - H_{45})/H_0$	(13)
Small, ε_{G_s}	$(H_{45} - H_2)/H_0$	(14)
Large and Small, $\varepsilon_{G_{l+s}}$	$(H_{13} - H_2)/H_0$	(15)
Superficial gas velocity		
All Gas, υ_G	$-\left(\frac{H_1 - H_0}{t_1 - t_0}\right)$	(16)
Large and Small, $\upsilon_{G_{l+s}}$	$\left(\frac{H_4 - H_1}{t_2 - t_1}\right)\frac{(H_{13} - H_4)(H_1 - H_4 t_1/t_2)}{H_0(H_1 - H_4)}$	(17)
Slug, $\upsilon_{G_{sl}}$	$\upsilon_G - \upsilon_{G_{l+s}}$	(18)
Small, υ_{G_s}	$\left(\frac{H_2 - H_4}{t_3 - t_2}\right)\frac{(H_{13} - H_4)(H_1 - H_4 t_1/t_2)}{H_0(H_1 - H_4)}\frac{(H_{45} - H_2)(H_4 - H_2 t_2/t_3)}{H_1(H_4 - H_2)}$	(19)
Large, υ_{G_l}	$\upsilon_{G_{l+s}} - \upsilon_{G_s}$	(20)
Bubble rise velocity		
Slug, $u_{B\,sl}$	$\upsilon_{G_{sl}}/\varepsilon_{G_{sl}}$	(21)
Large, $u_{B\,l}$	$\upsilon_{G_l}/\varepsilon_{G_l}$	(22)
Small, $u_{B\,s}$	$\upsilon_{G_s}/\varepsilon_{G_s}$	(23)
Large and small, $u_{B\,l+s}$	$\upsilon_{G_{l+s}}/\varepsilon_{G_{l+s}}$	(24)

3. Results and Discussion

3.1. Power data

Table 5. Ungassed power numbers of the two set of agitators

Impeller Configuration	Ungassed Power number	
6BDT + 6MFD	6BDT	4.6
	6MFD	1.6
6BDT + 6MFU	6BDT	4.5
	6MFU	1.8

The ungassed power numbers are calculated using equation (7), and are listed in *Table 5*. Figure 10 show ratio of gassed to ungassed power number against Flow number for individual agitators for both configurations; 6DBT + 6MFD and 6DBT + 6MFU. The relative standard deviation was calculated over the measured power data and was found to be within ±10% of error range. According to Nienow (1998), for standard Rushton turbine, at constant agitator speed the power drawn decreases as the gas flow rate increases due to the growing cavity size in low viscosity systems. The gassed power number will also decrease under constant aerated conditions with increasing agitator speed due to increased amounts of gas recirculation, and increase gas hold-up. The results in Figure 10 show similar pattern for the Ruston Turbine agitator upon increasing the agitator speed, i.e. the power decreases with both configurations (6MFD and 6MFU). However, at constant agitator speed, when increasing the gas flow rate, the power drawn decreases to a minimum value before increasing again. The power gas factor for the Rushton Turbine in both configurations is essentially the same at equal speed and gas flow rate. This means that the direction of pumping from the mixed flow impeller has little effect on this, at the examined agitator spacing. The power draw increase is at high gas flow numbers where the high gas flow rates dominate the agitator.

The power draw results for 6MFD and 6MFU (with Rushton Turbine) in figure 10 show differences. This is due to the direction of the pumping of the agitators. For the down pumping configuration a pattern similar to the 6BDT is seen, with a decrease with gas flow followed by an increase. For the up pumping configuration a plateau is reached at intermediate gas flow numbers, but then the decrease in power continues. The difference between the two configurations is more pronounced when shown at full scale, in Figure 11. From Figure 11, it can be seen at the lower agitation speeds, the largest differences occur. For the down pumping system the power draw increases significantly needing in some cases about 6 times the ungassed power. This is due to the large force being supplied by the gas pushing on the blades of the top agitator, due to their 45^{o} angle. However, for same agitation speeds the power drawn decrease to a negative value of Pg/Pu for 6MFU. Negative power is of course a strange concept? What is happening in this case is that the large gas flow rate is pushing up on the 6MFU blades, this force is then converted into angular momentum of the 6MFU due to the 45^{o} pitch of the blades. For the upwards flow conditions, the direction of agitator rotation is the same as the direction pushed by the gas. This means than the power needed to turn the 6MFU is lower as the gas is helping. At the very high gas flow numbers, this effect is so significant that the gas overwhelms the power draw of the 6MFU and actually starts pushing the 6BDT as well. This is of course less surprising when noting that the highest gas flow rate studied in this paper has about 120 times the power of the 6MFU at the lowest speed (based on the ungassed power) and in fact about 36 times the power of both agitators combined at the lowest speed (based on the ungassed power). As a confirmation of this fact, if the agitators are turned off, and the maximum gas flow rate is applied, then the agitators slowly start to spin due to the gas turning the 6MFU. It is important to try to understand the effect of the gas regime on the gassed power, i.e. is there a regime change which causes the power of the 6BDT to start to increase again with increasing gas flow?

All gas regimes were visually observed and plotted on the graph represented by Figure 12, as Froude number against gas flow number. This allowed a check to see if flow regime map based on equations (1) to (3) agree with visual flow observations. The observed L-CD transition points (in black circles) fit close enough to the predicted transition line for completely dispersed – loaded. Although the observed flooded loaded points (black triangles) are not far enough from Nienow's predicted F-L line, the F-L transition was observed at slightly higher agitator speed. A new transition combination was found above 0.5 gas flow number, where the transition was observed from Completely-dispersed to Flooded directly, so visually no loaded regime was observed. Thus the F-L predicted line actually becomes F-CD line above 0.5 gas flow number in case of this investigation.

It is also interesting to examine the location of the equality of power between the agitators and the gas, as it is difficult to imagine that the gas can be dispersed by the agitators if it contains more power. From a simple analysis taking the equality of the gas power and the agitator power we get,

$$\mathrm{Fl}_{[P_{gas}=P_{agi}]} = a\mathrm{Po}\left(\frac{D}{T}\right)\mathrm{Fr} \tag{25}$$

where a is the power gassing factor (P_g/P_u) when the gas power is equal to the agitator power, this value is dependent on Fr. The equality of power condition can be plotted as a function of the gas flow number for the Froude number as in Figure 13. It is interesting to note that there is no difference between the two configurations studied, most likely as the large differences in gassed power occur at higher gas flow numbers than these conditions. This analysis provides us with two key equations, one when considering both agitators together,

$$\mathrm{Fl}_{[P_{gas}=P_{agi}]} = 1.02\mathrm{Fr}^{0.71} \tag{26}$$

and one for just the power of the 6BDT,

$$\mathrm{Fl}_{[P_{gas}=P_{agi}]} = 0.66\mathrm{Fr}^{0.71} \tag{27}$$

The exponent on equations (26) and (27) when fitted independently are very similar in value and the confidence intervals overlap, therefore it was chosen to fit the values to be the same, this value had a standard error of 0.012. The standard error for the pre-exponential factor is 0.029 and 0.015 respectively.

The position of equation (26) can be plotted on the 6BDT results from Figure 10, Figure 14. It can be seen that the line of equal power is a good dividing line between the different trends in the variation of the gassed power. This means that instead of correlating the power gassing factor for different gassing regions, e.g. flooding, loading, and completely dispersed, it could be better to correlate for the two regions based on the ratio of the gas to agitator power. Using multi-linear least squares regression, the two regions can be fit with simple power law models. For gas power greater than the agitator power:

$$\frac{P_g}{P_u} = 0.222\ \mathrm{Fl}^{0.086}\mathrm{Fr}^{-0.334} \tag{28}$$

For gas power less than the agitator power:

$$\frac{P_g}{P_u} = 0.091\ \mathrm{Fl}^{-0.398}\mathrm{Fr}^{-0.315}(\mathrm{Fl}+1)^{1.253} \tag{29}$$

Figure 15 shows the quality of the fit of equations (28) and (29) for the power gassing factor. Both equations predict the true values well for the data collected, with an average error of 5.3%. For equation (28) the standard errors for the fitted values are 0.0032, 0.0070, and 0.011 respectively. For equation (29) the standard errors for the fitted values are 0.0042, 0.0091, 0.011, and 0.099 respectively.

At the very high power gassing factors (almost 1) in the agitator power greater than the gas power regime, the fit is the worst. This is expected as equation (29) does not tend to 1 as Fl tends to 0 as it is an empirical fit for the data collected. Very low gas flow rates are not the focus of this study, but could be predicted with equations of the style as presented by Taghavi et al. (2011).

3.2. Gas Liquid hold-up

The gas hold-up data for the two configurations is compared in Figure 16 which shows no major differences. With 6MFU agitator, the gas hold-ups are slightly higher compared to 6MFD and that is likely due to the slightly higher ungassed power number of the 6MFU and that the 6MFU works together with 6BDT to push the flow upwards, which creates a lot of fluctuation in level and at the same time, builds the gas hold-up with higher peaks. In general, for both set of agitators, the gas hold-up starts to level out at higher superficial gas velocity, especially at higher agitator speeds. Also for both configurations, with increasing agitation speed there is an increased gas hold-up, due to the faster movement hence larger force applied to move the gas allowing it to remain in the system for longer. It was observed that for lower agitation speeds as the gas flow rate increases the flow regimes is shifted to flooded which lowers the level of dispersion around the tank walls.

Cooke (2005) examined the hold-up with a very similar system, which included a 6BDT with 4MFD agitators at superficial gas velocities ranging from 0.0017 to 0.127 m s^{-1} with 0.1 to 6.7 W kg^{-1} of specific power input; notably these conditions are before the plateau in hold-up seen at high gas rates. Using a multi-linear least squares regression on the data to determine the best fit for the constants a, b, and c for equation (4), producing,

$$\varepsilon_G = 76.6{\varepsilon_T}^{0.39} v_S^{0.56} \tag{30}$$

Figure 17 shows the results from the combined data of Cooke (2005) and the data collected here for the 6BDT + 6MFD and 6BDT + 6MFU against the prediction from equation (30). It can be seen that the data from this investigation levels out at the higher hold-up values meaning the prediction is poor. However, at the lower gas hold-ups the prediction is appropriate.

As mentioned in the introduction, the value of the exponent on the superficial velocity varies depending on the range of superficial velocities, while the exponent on the total power is generally about 0.4 for this system type. Also at high superficial velocities there is a plateau in the hold-up, while at lower superficial velocities there is a variation in the hold-up. This means that the variation in hold-up with superficial velocity should not be a simple power law, but start with a large dependency and end with limited dependency. The best fit expression for this is,

$$\varepsilon_G = 867{\varepsilon_T}^{0.40} \frac{v_S^{0.95}}{1 + 58{v_S}^{0.95}} \tag{31}$$

Figure 18 shows the result of the predicted hold-up against the measured hold-up based on equation (31). All the data points fit well on the line apart from one or two anomalies, average error of 15.1%, thus this is a good correlation to combine all data including the very high superficial velocities. For equation (31) the standard errors for the fitted values are 22.5, 0.0093, 0.011, and 2.70 respectively (867, 0.40, 0.95, 58).

3.3. ERT gas hold-up distribution

Figure 19 shows ERT images, obtained from MATLAB for the 6BDT + 6MFU configuration at 180 rpm agitation speed across different gas flow rates. To produce Figure 19, 20 vertical cut planes were generated from the full 3D solution to the data, equally angularly spaced. These cut planes were then averaged to produce a single representation of the variation in the vessel. With increasing gas flow rate, the conductivity shows a general decrease as the blue regions gets larger representing higher gas dispersion. An odd result is observed at 0.12 m s^{-1} with the conductivity lower than expected; the possible reason could be that the dispersion level may have been slightly lower than the normal liquid level during the measurement. Initially the gas is mostly seen around the walls of the vessel with no gas below the agitator, which suggests loaded regime, as the gas flow rate increases at constant speed, the flow regime shifts to flooded from loaded, at around 0.12-0.16 m s^{-1}. This was also confirmed by visual observation. At the higher gas flow rates, it can be seen that due to very high gas flow rates and the configuration being 6MFU above 6BDT, the gas is circulating more; the 6MFU pushing the gas in upwards direction and the 6BDT pushing the gas in both upwards and downwards direction. For this reason, a little gas is observed below the 6BDT impeller as most of the gas is recirculating between the two impellers due to the nature of their design.

It is expected that the standard deviation of the conductivity within the tank will change with the gas flow rate, and that there might be a difference between loaded and flooded regimes. This means that the standard deviation of the conductivity could be used to determine the flow regime in the vessel. Figure 20 shows the standard deviation in the conductivity for the two configurations at 180 rpm plotted against superficial gas velocity. The flooding-loading boundary predicted by equation (1), the visually observed transition, and the point of equal gas to agitator power are also shown on Figure 20. At the point around the visual observations and the equality of power, the standard deviation changes from an increasing trend with increasing gas flow, to an approximately constant value. The values for the low flow rates are in the loaded regime and as the gas flow rate increases the standard deviation increases as the gas is becoming less evenly distributed. As the flow regime shifts to the flooded regime the standard deviation becomes flat, due to areas of high gas and less gas, and the gas flow just dominating the distribution.

The gas void fraction, ε, can be approximated as 100 x (1-σ), where σ, is the normalised conductivity, i.e. the conductivity of the mixture divided by the conductivity of the pure liquid. Smith (1992), based on the idea of Hassan and Robinson (Hassan and Robinson, 1977), proposed that the gas void fraction correlates linearly with the volumetric gas rate and the square of the impeller speed. Smith found linear relationship for single impellers; however for multiple impellers the correlation did not show linear relationship going through the origin for lower measured gas fractions (below at about 6%). Montante and Paglianti (2015) suggested that the approximate gas void fraction above could be correlated for a single impeller system with,

$$\varepsilon = A(\mathrm{ReFrFl})^{0.35}(D/T)^{1.25} \quad (32)$$

The above correlation fit was plotted for the data from this report for dual impeller systems shown by Figure 21. It shows a good linear relationship for multiple impeller systems, i.e. for both configurations investigated in this report at different speeds. The A constant value ranges from 0.625 for Rushton Turbine to 0.33 for Up-pumping 6 Pitched Blade Turbine in the literature (Montante and Paglianti, 2015). For this report the A constant value is calculated to be around 0.5 for both dual impeller systems (Rushton Turbine and 6MFD/U) which is within this in the range. Furthermore, these results are extended over a much higher value of $(\mathrm{ReFrFl})^{0.35}(D/T)^{1.25}$, up to 35 as compared to the original results which were only up to a value of 5.

3.4. Dynamic Gas Disengagement

The DGD studies were investigated and compared for the same two configuration systems. The 6MFD above 6BDT configuration was repeated at higher agitation speed i.e. 300 rpm. Figure 22 shows total hold-up for the three conditions against superficial gas velocity. The highest hold-up curve is seen for 300 rpm; it is understood as higher agitation speed increases gas hold-up. In comparison to 6MFD and 6MFU at constant speed, 6MFU shows higher gas hold-up for the reasons described earlier in this report as the dispersion flows in same direction by both agitator and gas power.

Figure 23 shows a graph of bubble rise velocity as a function of superficial gas velocity for the 6BDT and 6MFD configuration at 180 rpm. At very high superficial gas velocity, 3 bubble classes are identified; small, large and slugs. The third new bubble class represent very large bubbles, reported as slugs, formed due to higher gas flow rate. The bubble rise for these slugs increases with increasing gas flow rate. The large and small bubbles show an average bubble rise velocity for all conditions. It is difficult to identify the bubble class for points at 0.002 and 0.04 m s^{-1}, which indicates most likely to be anomalies. The results for the other two configurations are represented by *Table 6*. The average bubble rise velocity for large and small classes shows very similar results for 6MFU 180 rpm and 6MFD 300 rpm. In comparison with 6MFD at 180 rpm, these values are relatively low; as with 6MFU 180 rpm there is more gas recirculation due to the nature of the flow pattern described earlier, hence more bubbles break down into smaller size. At higher agitation speed of 300 rpm, the impellers move faster so again more break down of bubbles, therefore the bubble rise velocities are smaller.

Table 6.Bubble rise velocity for large and small bubbles at different conditions

Impeller Configuration	Agitator Speed (rpm)	Average Bubble rise velocity (cm s^{-1})	
		Large	Small
6BDT + 6MFD	180	41.4	13.1
6BDT + 6MFD	300	19.3	3.2
6BDT + 6MFU	180	19.0	3.2

4. Conclusions

The individual agitator power numbers within dual radial axial stirred tank show interesting results. Where Ruston Turbine has no significant effect on power drawn with in both configurations, the mixed flow downward/upward pumping show opposed results at higher gas flow rates. Downward pumping can use up to 5 times more power in gassed condition as compared to ungassed condition at higher gas flow rates. For the other configuration gas power dominates the agitator power, where 6MFU loses its effectiveness at high gas flow rates.

Flow regime maps are useful in identifying the flow regimes; completely dispersed, loaded and flooded, for known gas flow rate and speed. The Nienow et al. (1985) equations are valid up to 0.5 gas flow number, beyond this point, the transition seems to occur from completely dispersed directly to flooded.

The gas liquid holds up results show minimal difference between the two configurations and reaches its maximum. Homogeneous and heterogeneous data can be correlated using the equation introduced in this study.

With regards to the agitator choice; the results from this study indicate little or no difference between up flow pumping and down flow pumping. Overall it is very difficult to predict the better performing configuration out of the two studied in this investigation as not much difference is observed.

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