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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.

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

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.

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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

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 Fig. 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.

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

c	v_s ($m\,s^{-1}$)	Author
0.56	0.0017–0.127	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)

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). Fig. 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)

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

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

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

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 Eqs. (1)–(3) is shown in Fig. 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 Eq. (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 to 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