CHARACTERIZATION OF GAS-LIQUID PARAMETERS IN A DOWN-FLOW JET LOOP BUBBLE COLUMN

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Abstract - In order to characterize the gas-liquid parameters of a down-flow jet loop reactor, experiments were carried out to study the gas holdup, bubble sizes and their distribution, mass transfer area and mass transfer coefficient. The experiments were performed in the bubbly flow regime because of its stability and uniformity. Analysis of experimental data showed an unambiguous relationship between gas holdup, bubble size, bubble size distribution and interfacial area. It has also been found that gas holdup and interfacial area is a strong function of slip velocity. The variation of interfacial area and mass transfer coefficient were also analyzed and expressed as a function of operational and geometric parameters.
Keywords: Bubble column; Down-flow; Bubble size; Interfacial area; Mass transfer coefficient.

INTRODUCTION

Because of simple construction and low operating costs, bubble columns are widely used in various fields, which include petrochemical, chemical, biochemical and pharmaceutical industries, and water pollution abatement. Although numerous studies have been devoted to the fundamental understanding of complex behavior in bubble column reactors, their design and scale-up mostly relies on semi-empirical relations.

Studies on the two-phase co-current jet loop bubble column have gained considerable importance in recent years, in view of its inherent advantages, viz., finer and uniform bubble size, negligible coalescence of bubbles, homogenization of the phases, higher residence time of the gas bubbles and efficient dispersion of a very small amount of gas in a large amount of liquid. A literature review reveals that studies on the dispersion of gas in two-phase downflow have been carried out either with plunging jet or with sparger type systems. In a sparger type

system, the gas sparger is fixed at the top of the column. Liquid with high velocity is forced through the column and, as it moves, it shears the gas from the sparger in the form of bubbles(Shah et al., 1983; Kulkarni and Shah, 1984). The concept of plunging jet in a downflow column comes from the studies of gas entrainment and dispersion by a liquid jet impinging into a pool of liquid (Van de Sande and Smith, 1973; McKeogh and Ervine, 1981). However, this offers poor gas-liquid contacts due to lower residence time of the gas bubbles. Therefore, studies have concentrated on attempts to increase the gas-liquid contact time. Higher gas-liquid contact can be obtained by increasing bubble penetration, but it requires enough downward liquid flow to overcome the buoyant force of the bubbles produced by the liquid jet. In another way, a higher residence time of the entrained bubbles can be obtained by plunging a vertical liquid jet in a vertical column (Ohkawa et al., 1986; Bando et al., 1988; Yamagiwa et al., 1990). Milenković et al. (2007) reported the characteristics of large coherent structures in

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turbulent, periodically-excited, as well as naturally-developing bubbly jets and the interactions between bubbles and large-scale vortices.

In a plunging jet system having a downcomer, the liquid jet coming out through the nozzle directly impinges into the liquid inside the downflow contactor and forms an intense mixing zone, where the entrained gas breaks into fine bubbles by shearing forces of the jet. The gas and liquid then flow cocurrently down the column where they are discharged at the bottom. As the gas bubbles are forced to move downward against their upward buoyant force, the bubbles spend more time in the column compared to an upflow system.

The performance of a jet loop bubble column is best characterized by the available gas-liquid interfacial area (a), especially in the regime in which absorption is accompanied by a fast chemical reaction. Again, for gas-liquid processes that take place in the absorption regime with slow reaction, the mass transfer properties can be characterized by the volumetric mass transfer coefficient ($k_L a$). Thus, a precise knowledge of available gas-liquid interfacial area and volumetric mass transfer coefficient is very important for designing and scaling up a gas-liquid reactor. Efficient gas dispersion in a cocurrent downflow bubble column with ejector type gas distributor leads to an enhanced mass transfer. Investigations on mass transfer in bubble columns with liquid driven ejectors have been reported by several authors. Atkinson et al. (2003) studied the bubble size distribution and energy dissipation rate measured in a plunging liquid jet bubble column. They predicted the maximum stable bubble diameter in the mixing zone based on Weber number analysis. Uchiyama et al. (2008) reported the gas holdup distribution in a gas bubble dispersed phase generated by an impinging gas–liquid two-phase jet measured with an optical fiber void probe. They also studied the effects of structural factors of nozzle and column diameter on gas holdup distribution. Huynh et al. (1991) studied mass transfer characteristics in an upward Venturi/bubble column combination. They reported that gas holdup was increased by 50% to 150% and the overall volumetric mass transfer coefficient was tripled when Venturi was used as "gas distributor" instead of a porous distributor. Manish and Majumder (2009) investigated the quality of mixing in a modified downflow bubble column, which they analyzed by using information entropy theory. They also developed empirical models for a downflow system based on parameters that affect the quality of mixing and mass transfer efficiency.

Havelka et al. (2000) measured the volumetric liquid-side mass transfer coefficient in the cocurrent ejector-distributor bubble column. Significant works have also been done on mass transfer in downflow system with efficient dispersion, e.g., liquid jet ejector (Dutta and Raghavan, 1987), Venturi-bubble column (Briens et al., 1992) and plunging jet bubble column (Evans et al., 2001). However, comparison of these different systems is difficult, because of their different ejector-contactor configurations and operating ranges of liquid and gas flow rates.

The present paper deals with the performance characteristics of a jet loop bubble column, with special emphasis on gas entrainment, gas holdup, bubble sizes, interfacial area and mass transfer coefficient. The novelty of the system is that no extra energy is required for gas dispersion and gas holdup is significantly high. The experimental results are analyzed and elucidated as a function of operational and geometric parameters.

EXPERIMENTAL

A detailed experimental set up for the ejector induced downflow system is shown in Fig. 1. It consists of an ejector assembly, an extended pipeline contactor, a gas liquid separator, and other accessories. The ejector assembly and extended contactor are made of transparent Perspex for visual observation of the flow and mixing patterns. The inside diameter of the column is 51.6 mm and length of the column is 1.5 m. A high velocity liquid jet is produced through the nozzle of the ejector system, which entrains air along with it. The liquid jet with surrounding air plunges at the top of the column, forming fines bubbles, which are moved downward by the higher momentum of the liquid jet prevailing over the upward buoyancy forces of the bubbles. The column is then filled with gas-liquid mixture by proper adjustment of pressure and liquid level inside the separator through the valves V_4 and V_5, respectively. Two distinct zones are observed in the column at this stage: an intense mixing zone at the top, followed by a zone of two-phase bubbly flow. The length of the intense mixing zone depends on the operating and geometric parameters of the system. It has been found that the length of the intense mixing zone increases with decreasing nozzle diameter at constant operating conditions. For the same nozzle diameter, it again increases with increasing liquid flow rate. Its length generally varies from 30 mm to 45 mm. In the intense mixing zone, the jet energy is utilized to form fine gas bubbles. The experiments are carried out with water as motive fluid and air as secondary fluid that entrains through the secondary entrance of the ejector. The detailed experimental procedure has been discussed in an earlier paper (Mandal et al., 2004).
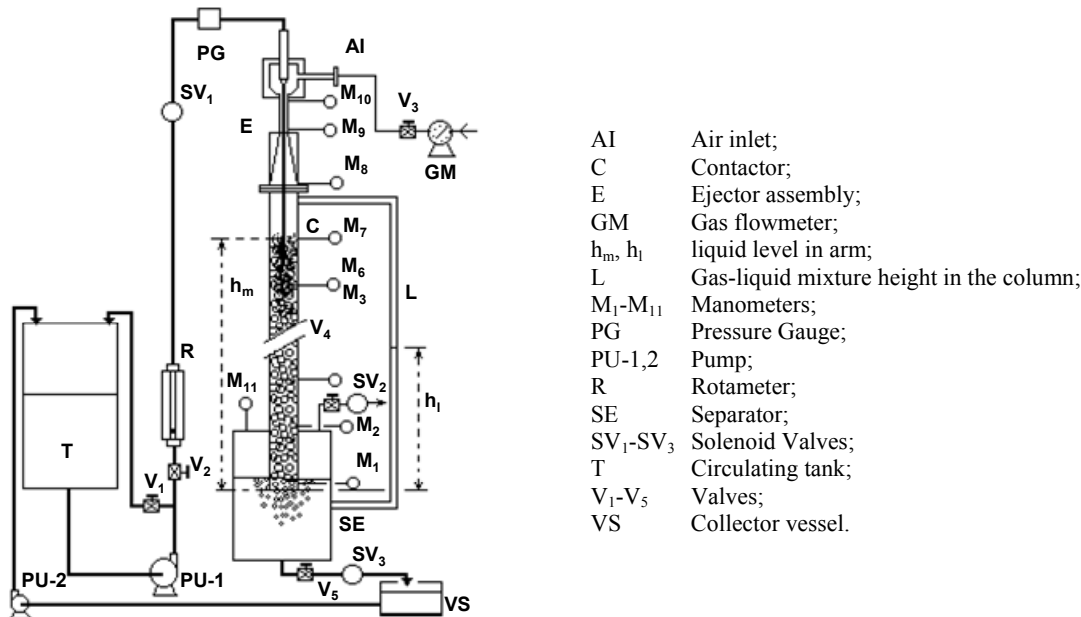
Figure 1: Schematic diagram of the experimental set-up

AI	Air inlet;
C	Contactor;
E	Ejector assembly;
GM	Gas flowmeter;
h_m, h_l	liquid level in arm;
L	Gas-liquid mixture height in the column;
M_1-M_{11}	Manometers;
PG	Pressure Gauge;
PU-1,2	Pump;
R	Rotameter;
SE	Separator;
SV_1-SV_3	Solenoid Valves;
T	Circulating tank;
V_1-V_5	Valves;
VS	Collector vessel.

Gas Hold-Up

Gas holdup for the present system has been measured by the "flow isolating technique" as reported in the literature (Watson and Grainger-Allen, 1974; Kundu et al., 1995). When a steady state condition of the system is attained, the total height of the gas-liquid mixture in the column is noted. Then the three solenoid valves, SV_1, SV_2 and SV_3 (Fig. 1) are switched off simultaneously. This causes an immediate termination of flow of both the fluids. The liquid gas mixture inside the column gets arrested and is allowed to settle for some time, whereby all gases get separated. The clear liquid height inside the column is then noted. The difference between the gas liquid mixing height (h_m) and the corresponding clear liquid height (h_l) gives the holdup of gas in the column. Thus, the average gas holdup can be represented as:

$$\varepsilon_g = \frac{h_m - h_l}{h_m} \tag{1}$$

Bubble Size

Bubble sizes are measured by a photographic method (Schafer et al., 2002). The column is surrounded by a 150 mm×150 mm clear Perspex jacket with flat surfaces and the gap between the column and the jacket is filled with distilled water. This arrangement is made to avoid the optical distortions caused by the curved surface of the Perspex column. The rear part of the column is illuminated by a 500 W halogen lamp and the opposite wall is covered with a diffusion screen to avoid undesired reflections and refractions as much as possible. Then images are taken at different heights of the bubbly flow zone under different operating conditions in the column with a high-speed digital camera. Good quality images are then analysed by image processing software (Olympus Microsuite [TM] Basic) to obtain the bubble sizes. More than 100 bubbles are considered in each experimental condition to obtain the size distribution.

Interfacial Area

Interfacial area can be obtained both by physical and chemical methods. The chemical method of interfacial area determination is based on gas absorption followed by chemical reaction. In general, when physical absorption is followed by a chemical reaction, the overall rate is governed by both the physical rate of absorption and the kinetics of the reaction. Brian (1964) has discussed the theory of absorption followed by a chemical reaction of general order. The liquid phase reaction between dissolved carbon dioxide and sodium hydroxide satisfies the condition for a pseudo-first order reaction at certain alkali concentrations and was therefore selected for the present study. This reaction is said to take place as,

$$CO_2 + OH^- \rightarrow HCO_3^- \tag{2}$$

$$HCO_3^- + OH^- \rightarrow CO_3^{-2} + H_2O \qquad (3)$$
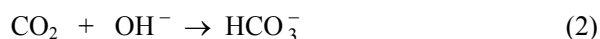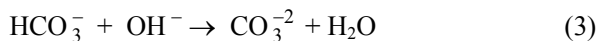
Since the second reaction is ionic, it takes place at a much faster rate than the first one; therefore, the first reaction step is rate controlling. The second order rate constant, k_2, was determined by the equation suggested by Porter et al. (1966). If the concentration of OH^- remains essentially constant in the bulk of the solution, the equation becomes pseudo-first order. The condition to be satisfied for this is given by the following condition (Danckwerts, 1970),

$$\left[\frac{D_A k_2 C_B^o}{k_L^2} \right]^{\frac{1}{2}} \quad << \quad 1 + \frac{C_B^o}{2 C_A^*} \qquad (4)$$

D_A, the diffusivity of carbon dioxide in aqueous solutions was obtained from the equation reported by Nijsing et al. (1959)], It has been shown by Sharma and Danckwerts (1970) that the specific rate of absorption per unit area for a second order reaction satisfying the condition of Equation (4) is,

$$R = C_A^* \sqrt{D_A k_2 C_B^o + k_L^2} \qquad (5)$$

If further, $\left[\dfrac{D_A k_2 C_B^o}{k_L^2} \right]^{\frac{1}{2}} > 3$, $\qquad (6)$

the second term within the square root of Equation (5) becomes small compare to the first term and, therefore, the specific rate is given by the simplified Equation (7),

$$R = C_A^* \sqrt{D_A k_2 C_B^o} = C_A^* k_{LR} \qquad (7)$$

When the conditions given by Eqs. (4) and (6) are satisfied, the rate of absorption will be a function of the physical-chemical factors and will be independent of the hydrodynamic conditions in the system.

If further the gas phase resistance can be neglected, the concentration of carbon dioxide at the interface, C_A^*, is related to its partial pressure in the gas phase by,

$$p_A^* = H_e C_A^* = p_A = y P' \qquad (8)$$

There are some experimental evidences in the literature (Porter et al., 1966; Voyer and Miller,

1968) to suggest that gas phase resistance does not exceed about 10 percent of total resistance and can be neglected. Therefore, following the method suggested by Voyer and Miller (1968), one can write the rate of absorption in a differential volume dv,

$$R a dv = a C_A^* k_{LR} dv \qquad (9)$$

From a material balance equation on a solute free basis, one can derive (Mandal, 2003) the following expression for interfacial area:

$$a = \frac{G' H_e}{k_{LR} P' V} \ln \left[\frac{1 - y_o}{y_o} \frac{y_i}{1 - y_i} \right] + \left[\frac{1}{1 - y_i} - \frac{1}{1 - y_o} \right] \quad (10)$$

Eq. (10) can be used to determine the specific interfacial area, provided the conditions for the pseudo-first order reaction are satisfied. Experimental results for the present system were found to satisfy the conditions for "fast pseudo-first order reaction" as given by Eqs. (4) & (6). The values of specific interfacial area were obtained by using Eq. (10) with knowledge of D_A, k_2 and H_e together with the data of inlet and outlet concentration of carbon dioxide, volume of gas-liquid mixture in the column and gas flow rate. The Henry's law constant, H_e, in ionic solution was directly obtained from solubility, S. The diffusivity of carbon dioxide in aqueous solution, D_A, was calculated from equations suggested by Nijshing et al. (1959), while the second-order reaction rate constant, k_2, was determined by the equation suggested by Porter et al. (1966).

Liquid-Side Volumetric Mass Transfer Coefficient

Similar to interfacial area, experimental measurements of volumetric mass transfer coefficient are also categorized as physical and chemical methods. Measurements taken in the absence of simultaneous chemical reaction are commonly used to determine the volumetric mass transfer coefficients in bubble columns. Investigations involve the purely physical mass transfer between the gas and the liquid phase. Oxygen is used as the exchange gas in the majority of the studies, as its partial pressure in the liquid phase is easy to determine electrochemically using polarographic electrodes. Van de Sande and Smith (1975), Bin and Smith (1982) have used the above method to study the mass transfer characteristics.

The volumetric mass transfer coefficients for the present systems were measured by a chemical method. Among different chemical methods,

absorption of CO_2 into a carbonate- bicarbonate buffer solution is a convenient system, and is represented by the second-order reaction Equation (11).

$$CO_2 + OH^- \rightarrow HCO_3^- \qquad (11)$$

Under certain conditions the above reaction is sufficiently fast, so that the concentration of CO_2 in the bulk of the liquid phase is equal to zero. The condition to be satisfied for this is,

$$k_L a << \varepsilon_l k_2 C_B^o \qquad (12)$$

Further, the reaction rate is such that no appreciable amount of reaction takes place in the diffusion films, and the condition to be satisfied for this is given by,

$$\frac{D_A k_2 C_B^o}{k_L^2} << 1 \qquad (13)$$

When the conditions given by Equations (12) and (13) are satisfied, the rate of mass transfer is given by (Danckwerts, 1970)

$$Ra = k_L a C_A^* \qquad (14)$$

The ratio of CO_3^{2-} to HCO_3^- in the buffer solution was maintained within 1.6-4.3 to satisfy the conditions given by Eqs. (12) and (13). By measuring Ra, the rate of mass transfer per unit volume and interfacial concentration of carbon dioxide ($C_A^* = HeP_A$), the mass transfer coefficient, $k_L a$, can be determined.

RESULT AND DISCUSSIONS

Flow Rates and Regimes

Flow conditions have significant effects on momentum, mass and energy transfer in gas-liquid contacting systems and, hence, it is desirable to maintain a constant flow regime throughout. Depending on the flow conditions, there are mainly four types of flow, viz., homogeneous bubbly flow, heterogeneous churn flow, slug flow and annular flow. Generally flow regimes are largely dependent upon the superficial gas velocity and column diameter. But for the present system of an ejector-induced downflow bubble column, where gas is sucked in by a high velocity liquid jet, the flow regimes depend primarily on liquid flowrate at constant column diameter. As liquid flowrate increases, there is a natural tendency of the gas bubbles to coalesce, forming large bubbles. Under this condition, large bubbles tend to move upward due to higher buoyant forces and, consequently, heterogeneous churn-turbulent or slug flow results. Kulkarni and Shah (1984) reported a flow regime map in terms of superficial gas and liquid velocities for possible operation in a cocurrent downflow system. With a sparger type gas distributor they found that operation is possible only in regime A; regime C is undesirable practically and operation is not possible in regime B. However, from Fig. 2 it is found that operation is also possible in regime B with an efficient gas-liquid distributor like an ejector-nozzle system. The Fig. 3 also shows that the gas to liquid flow ratio in the present work is much higher than that employed by Bando et al. (1988) with a similar system.
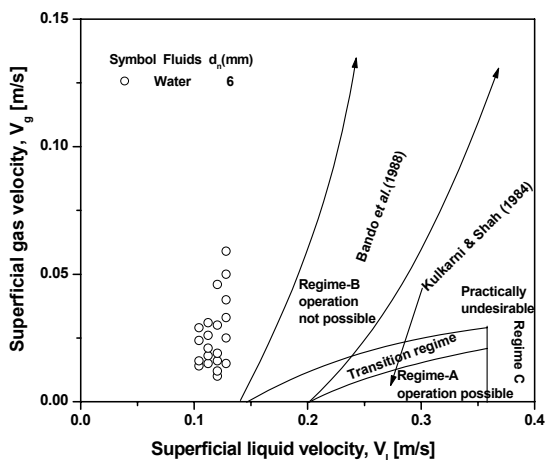

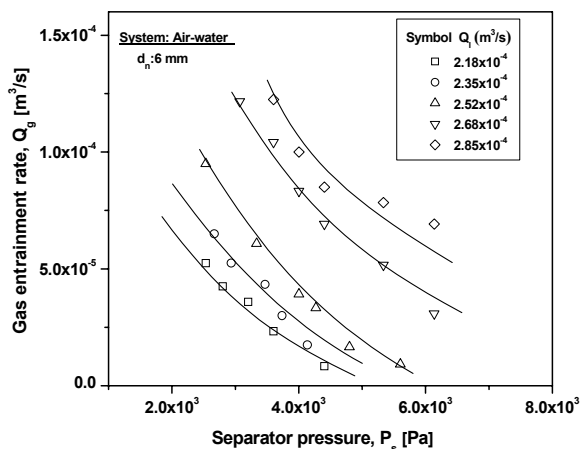**Figure 2:** Flow regime map of stable co-current downflow

Figure 3: Effect of separator pressure on gas entrainment at different liquid flowrates for the air-water system

Gas Entrainment by Liquid Jet Ejector

The downflow bubble column with gas entrainment by a liquid jet ejector system conforms to a new type of cocurrent bubble column, which can be successfully used in chemical, fermentation and waste treatment processes because of its self aeration characteristics. Entrainment occurs due to characteristics of the plunging liquid jet and the rate of entrainment is mostly controlled by the jet velocity. However, for a plunging liquid jet with a downcomer, entrainment also depends on the two-phase gas-liquid mixing height inside the downcomer. The entrained gas disperses into the liquid as fine bubbles in the intense mixing zone and the dispersed gas bubbles are then carried downward by the high momentum of downflow liquid. A minimum liquid flowrate is required to move the bubbles in the downward direction by prevailing over their upward buoyant force, which depends on the bubble size and physical properties of the fluids. Significant works have been reported on gas entrainment by plunging liquid jets through nozzles, Venturies and ejector system with and without downcomers.

The variations of gas entrainment rate (Q_g) with separator pressure for air–water at different liquid flow rates are presented in Fig. 3. It may be seen from the figure that, at constant liquid flow rate, gas entrainment rate decreases with separator pressure, P_s. As the separator pressure increases, the entrained gas bubbles face more resistance to move downward and, consequently, lower entrainment results. The effects of motive liquid flow rate on air entrainment rate at constant P_s are presented in Fig. 4. It may be seen from the plots that entrainment rate increases with liquid flow rate for a particular nozzle, due to the increase in the average velocity of the air film around the liquid jet (Evans et al., 1996). Entrainment rate generally increases with a decrease in nozzle diameter for constant liquid flow rate due to increased jet velocity. Further, on extrapolating the plots of Q_g vs Q_L of Fig. 4, the liquid flow rate corresponding to zero gas flowrate gives the minimum entrainment flow rate. For the present system, the minimum entrainment velocity based on nozzle diameter was found to vary between 5 m/s and 5.5 m/s.
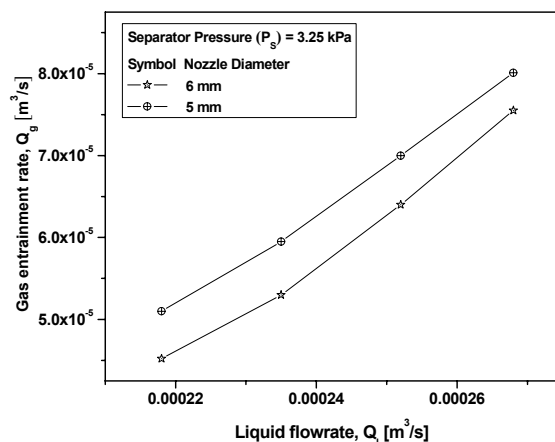

Figure 4: Effect of motive liquid flowrate on the air entrainment rate
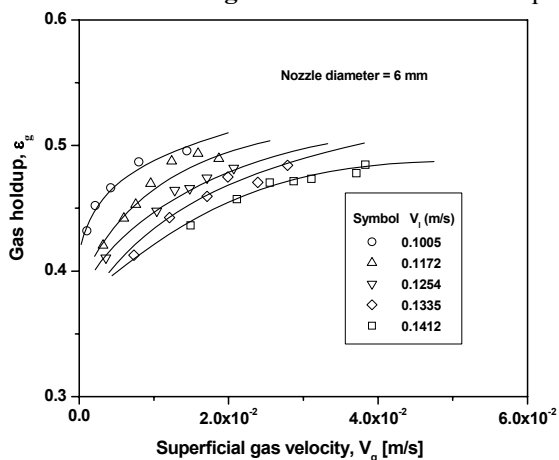

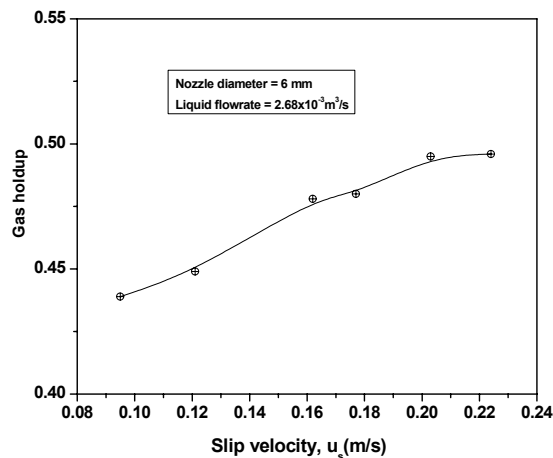**Figure** 5: Variation of overall gas holdup with gas flowrate

Figure 6: Variation of gas holdup with slip velocity

Gas Hold-Up

In cocurrent downflow, the gas bubbles move against their high buoyant force, having a higher slip velocity compared to an upflow system. Thus, the bubbles have a longer residence time and hence higher gas holdup. Fig. 5 shows the variation of overall gas holdup with gas flowrate at constant liquid flowrate. At a constant liquid flowrate, gas holdup increases with increase in gas flowrate because of increased bubble population. However, the range of gas holdup variation is very small, because even at very low gas flow rate a higher gas holdup was observed due to a higher residence time of the bubbles. It may also be seen from the plots that gas holdup decreases with liquid flow rate at the same separator pressure, due to the increase in momentum imparted by the liquid on the gas bubbles.

The slip velocity model, first introduced by Behringer (1936), has been used to analyze the experimental holdup data for two-phase flow in the present system. The slip velocity of bubbles relative to the surrounding liquid is defined as;

$$u_s = \pm [u_g - u_l]$$

$$= \pm \left[\frac{V_g}{\varepsilon_g} - \frac{V_l}{1 - \varepsilon_g} \right] \quad (15)$$

where the negative sign has been incorporated to consider the downflow system. Thus, gas holdup data can be correlated as a function of V_g and V_l. The variation of gas holdup with slip velocity is shown in Fig. 6. The gas holdup increases with increase in slip velocity because of the increasing number of bubbles with higher residence time in the column.

Bubble Size

In the two-phase gas-liquid system, the simultaneous bubble coalescence and breakup was considered to be largely responsible for the variation in bubble sizes throughout the column. The bubble size distribution was obtained at different axial positions of the column under steady state of a homogeneous bubbly flow. The sampling of bubbles was taken at different axial positions of the column, when the column is completely filled with the gas liquid mixture. Typical bubble size frequency distribution curves are shown in Fig. 7. The size distributions of the bubbles were found to deviate from a normal distribution with a skewness towards the larger bubble sizes. The Sauter mean bubble diameter was then calculated from the sampled bubble's size and, consequently, interfacial area was calculated from Eq. (16).

$$a = \frac{6\varepsilon_g}{d_s} \quad (16)$$

where d_s is the Sauter mean bubble diameter and is defined as,

$$d_s = \frac{\sum n_i d_i^3}{\sum n_i d_i^2} \quad (17)$$

where n_i is the number of bubbles of diameter d_i. The distribution of d_s along the column is presented in Fig. 8. Using this value of Sauter mean bubble diameter, one can calculate the interfacial area from Eq. (16). The calculated values of interfacial area are shown as a function of slip velocity in Fig. 9. Interfacial area increases with slip velocity because of increasing gas holdup.
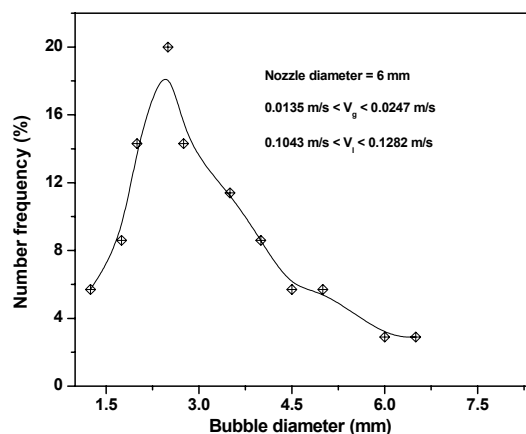

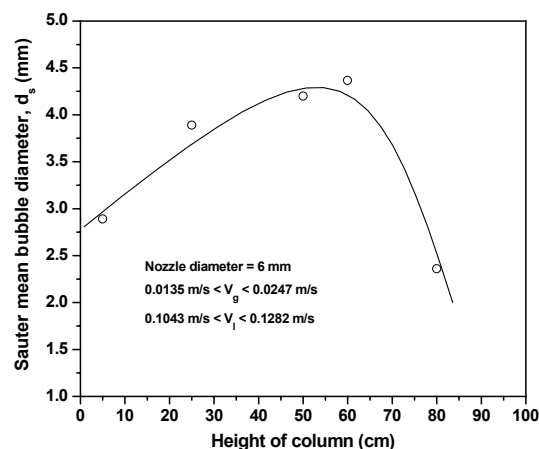**Figure 7:** Distribution of bubble size at the middle of the column.

Figure 8: Variations of Sauter mean bubble diameter at different positions of the column.
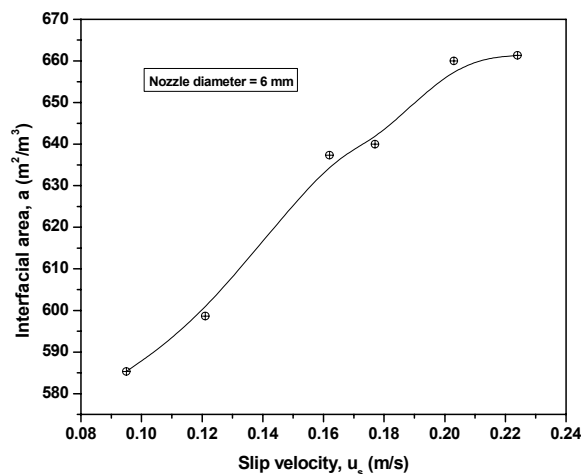
Figure 9: Variation of interfacial area with slip velocity

Measurement of Interfacial Area by Chemical Method

In the chemical method, the interfacial area was measured by absorbing carbon dioxide into an aqueous solution of sodium hydroxide. For the present system, there is an unambiguous relationship between the gas flow rate and interfacial area. The variations of interfacial area are shown graphically with superficial gas velocity, V_g in Fig. 10. With an increase in gas flow rate, both bubble population and gas holdup increase in the column. Therefore, higher interfacial area was observed at higher gas flow rate. The interfacial area obtained by the carbon dioxide absorption method in the present system was 100-300 m^2/m^3.

Mass Transfer Coefficient

The product of the liquid-side mass transfer coefficient k_L and the specific interfacial area a is known as the volumetric mass transfer coefficient ($k_L a$). The product $k_L a$, which is easy to determine experimentally, is the only essential feature for a reaction that is slow enough in comparison with mass transfer (Schumpe and Deckwer, 1980). In the present system, $k_L a$ was measured by absorbing CO_2 into a carbonate-bicarbonate buffer solution. The variations of $k_L a$ with simultaneous superficial liquid and gas velocity are shown in Fig. 11. $k_L a$ increases with superficial gas and liquid velocity due to increased gas bubble population and turbulences of the system, respectively.

If we assume that the interfacial area, a, measured by absorption of carbon dioxide in dilute aqueous solution of NaOH is more or less the same in the carbonate-bicarbonate buffer system, the true liquid-side mass transfer coefficient, k_L, can be calculated from Eq. (18).

$$k_L = \frac{k_L a}{a} \qquad (18)$$

A series of empirical and semi-empirical correlations have been proposed for liquid-side mass transfer coefficients in gas-liquid dispersions (Calderbank, 1959; Akita and Yoshida, 1974). Essentially, k_L values are a function of bubble size and liquid flow pattern around the bubbles. Calderbank and Moo-Young (1961) carried out detailed and extensive research on the mass transfer coefficient k_L as a function of bubble diameter in stirred tank reactors and bubble layers. A rule of thumb for low-viscosity media is that k_L values range between 0.015-0.035 cm s^{-1}.

The k_L values for the present system, ranging between 0.025 and 0.05 cm/s, are in good agreement with those found in the literature (Ohkawa et al., 1987; Bouaifi et al., 2001). A comparative picture of gas holdup, interfacial area and mass transfer coefficient, along with the operating conditions for different gas-liquid contactors, is given in Table 1. The present system offers significantly higher gas holdup and comparable interfacial area and mass transfer coefficient.
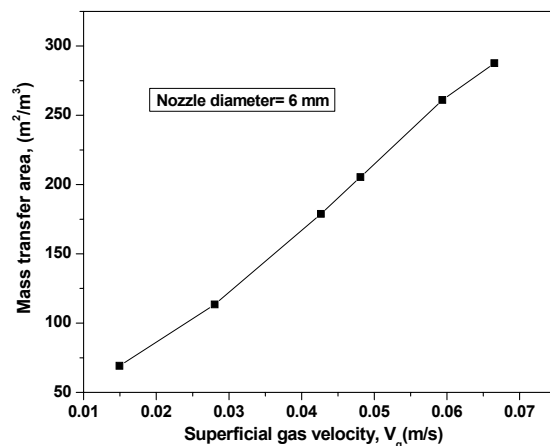
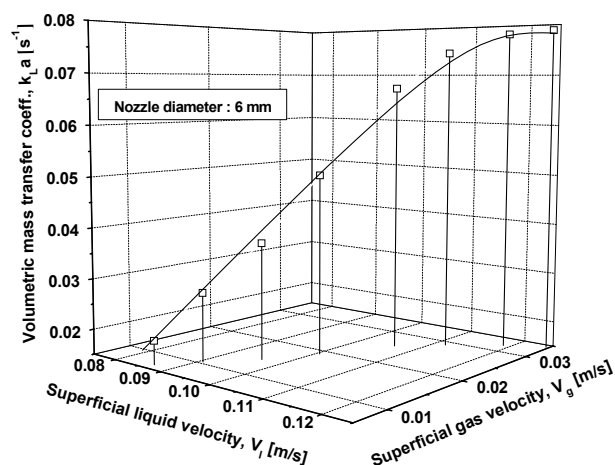**Figure 10:** Variation of interfacial area with superficial gas velocity

Figure 11: Variation of $k_L a$ with superficial liquid and gas velocity

**Table 1: Comparison of gas holdup, interfacial area and mass
transfer coefficients for different gas-liquid contactors**

Author	Type of reactor	Superficial Liquid velocity, V_l, m/s	Superficial Gas velocity, V_g, m/s	Gas holdup, dimen-sionless	Interfacial area a (m^2/m^3)	Mass transfer coeffi., $k_L a$ (s^{-1})
Ohkawa et al. (1987)	Downflow liquid jet bubble column	0.30-0.60	0.001-.02	0.10-0.40	250-450	0.02-0.10
Dutta and Raghavan (1987)	Jet loop reactor	0.001-0.005	0.001-0.006	0.02-0.15	30-500	0.01-0.08
Havelka et al. (2000)	Upflow ejector loop reactor	0.017-0.025	0.02-0.09	0.15-0.35	180-450	0.03-0.20
Bouaifi et al. (2001)	Bubble column	—	0.0025-0.04	0.01-0.30	2-130	0.005-0.10
Ide et al. (2001)	Plunging jet absorber	0.028-0.071	—	0.09-0.35	80-800	0.02-0.20
Cerri et al. (2009)	Concentric-tube airlift reactors	—	0.001-0.10	0.01-0.16	25-250	0.01-0.07
Present system	Downflow ejector bubble column	0.05-0.11	0.01-0.09	0.40-0.60	100-300	0.01-0.10

CONCLUSION

Experiments were carried out to measure the gas-liquid parameters of a down-flow jet loop bubble column reactor. Based on the experimental results and analysis of the data, the following conclusions may be drawn:

▪ Gas holdup obtained in the present system is significantly higher compared to other reported gas-liquid systems because of the fine dispersion of gas bubbles with the plunging jet and the higher residence time of the gas bubbles in the downflow column.

▪ Gas holdup and interfacial area is a strong function of gas-liquid slip velocity.

▪ Interfacial area obtained from gas holdup and Sauter mean bubble diameter was found to be higher than that from the chemical method because, the bubbles in the present case are mostly distorted and have substantial common interfaces between them.

▪ Mass transfer coefficients are functions of both liquid and gas flow rate, as both are responsible for bubble population and turbulence of the system. However, beyond a certain flow rate, the effects gradually decrease and the mass transfer coefficient values remain constant.

NOMENCLATURE

a	specific interfacial area per unit volume	m^2/m^3
C_A^*	concentration of dissolved gas at the interface, in equilibrium with gas at the interface	$gmol/cm^3$
C_B^o	concentration of reactant B in the bulk of the solution	$gmol/l$
D_A	diffusivity of reactant A inthe liquid phase	cm^2/s
d_n	diameter of the nozzle	m
d_s	Sauter mean bubble diameter	m
G'	molar flowrate of inert gas,	$gmol/s$
h_l	clear liquid height, corresponding to total gas-liquid mixing height in the column	m
h_m	gas-liquid mixing height in the column	m
He	Henry's law constant	$cm^3.atm/gmol$
k_2	second order reaction rate constant	$cm^3/gmol.s$
k_L	liquid-side mass transfer coefficient	cm/s
k_{LR}	coefficient of chemical absorption	$(D_A k_2 C_B^o)^{1/2}$, cm/s
p_A	partial pressure of A	atm
p_A^*	equilibrium partial pressure of A	atm
P'	system pressure	$atm.$
P_s	separator pressure	Pa
Q_g	volumetric flow rate of gas	m^3/s
Q_l	volumetric flow rate of liquid	m^3/s
u_s	slip velocity of gas and liquid	m/s
V	volume of gas-liquid mixture inside the column	m^3
V_g	gas superficial velocity	m/s
V_l	liquid superficial velocity	m/s
y	composition of CO_2 in gas phase	$mol\%$
y_o	composition of CO_2 in gas phase at inlet	$mol\%$
y_i	composition of CO_2 in gas phase at outlet	$mol\%$
ε_g	gas holdup	dimensionless
ε_l	liquid holdup	dimensionless

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