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Influences of Operating Variables on Hydrodynamic Performance of Plunging Water Jet Downflow Bubble Column

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Abstract
The hydrodynamics of a co-current down flow bubble column has been investigated with air – water system. A Perspex bubble column of 5cm in diameter and 1.5m height is used as a test contactor using nozzles of 7, 8 and 9 mm diameter for air-water distributing. The column is provided with three electro-resistivity needle probes for bubble detection.

Experimental work is carried out with air flow rates from 0.09 to 0.45 m^3/hr and liquid flow rates from 0.65 to 1.1m^3/hr in order to study the effects of superficial gas velocity, nozzle diameter and liquid flow rate on the characteristics of hydrodynamic interactions viz. gas hold up, bubble diameter and bubble velocity by using two technical methods, direct height measurements for air-water mixture in the column and resistivity probe techniques.

Gas hold up is found to be progressively increased with increasing superficial gas velocity and with decreasing liquid flow rate. Lower gas hold up is obtained with smaller nozzle diameter. However, gas hold up in two-phase zone is considerably higher than the corresponding value in mixing zone.

The mean bubble velocity is increased with increasing superficial gas velocity, liquid flow rate and nozzle diameter for both mixing and two phase zones. Experimental data are found to be fairly fitted with the Drift Flux model of Zuber and Findly.

The bubble diameter is considerably increased with increasing superficial gas velocity and with decreasing liquid flow rate, whereas it is slightly influenced by nozzle diameter. However, the bubbles in two-phase zone are relatively bigger than those observed in mixing zone. Finally, mathematical correlations have been developed from the experimental data to describe the gas hold up and bubble velocity in the uniform two-phase zone.

Keywords: Gas Hold-up, Bubble velocity, Bubble size, Down-flow Bubble Column, Plunging Jet.

Introduction
Bubble column is a unit in which gas stream is dispersed in the continuous liquid phase as fine bubbles. There are many chemical and biochemical processes, viz. hydrogenation, oxidation, fermentation, petroleum refining, coal liquefaction etc, in which the overall production rate is often controlled by gas-liquid mass

transfer[1]. The reaction rate for such types of processes is proportional to the interfacial area, which is available for mass transfer. However, large interfacial area will be obtained if the gas is dispersed into the liquid phase as fine bubbles.

Numerous studies on the design and development of gas-liquid contacting equipment show that contactors or reactors belonging to the jet-mixing category with co-current or countercurrent contacting of phases such as nozzles, venturies and ejectors are going to be attractively important now a days due to the higher interfacial area and mass transfer coefficients obtained in such systems[2]. In general, the kinetic energy of fluid is used to achieve fine dispersion and mixing between phases in all co-current flow devices. The downflow bubble columns with plunging jet system have been highly recommended for chemical processes particularly wherein interfacial mass transfer area is the rate controlling step.

Considerable work has been reported by different authors on efficient dispersion of gas by liquid jet in gas-liquid two-phase co-current contactor with nozzles, venturies and ejectors as gas- liquid mixing devices[3,4].

Several aspects of gas entrainment phenomena by plunging jet are reviewed in the experimental and theoretical studies, Where the downcomer section of the plunging jet bubble column is commonly consisted of four main regions, namely the free jet, plunging jet, mixing zone and uniform two-phase flow zone[5].

Gas holdup is the most important parameter that can be used to evaluate the hydrodynamic performance of bubble columns. It represents the percentage by volume of gas in two or three phase mixture. The overall gas holdup under two phase steady operation can be calculated according to the relation:-

$$\varepsilon_g = \frac{h_m - h_l}{h_m} \qquad \ldots (1)$$

where h_m and h_l are the total gas – liquid mixing height, and the corresponding clear liquid height respectively.

Gas holdup in two phase system gives the volume fraction of each phase and also determines the interfacial area between the gas and the liquid phases. Early work for evaluation of gas holdup of co-current two phase flow in horizontal pipe under different flow conditions was carried out by Lockhart and Martinelli[6]. Gas holdup in vertical upflow of air-water mixture using either a gas or liquid jet ejector was studied by Mitra et al.[7] . However, Kazumori et al.[8] used plunging water jet in air without downcomer section and observed tow correlations of gas holdup for laminar and turbulent jet velocities. Briens et al.[9] studied venture bubble column with both downflow and upflow modes. They obtained much higher gas holdup (0.15-0.4) in downflow compared to upflow (0.08-0.12). However, Deckwer[10] showed that gas distributor and physical properties of fluid have some further effect on gas holdup. Havelka et al.[11] and Akosman et al.[12] showed that gas holdup depends upon the superficial gas velocity and physical properties of gas and liquid. On the other hand, Dema et al.[13] observed about 69-80% enhancement in gas holdup by using two venture ejectors positioned between the nozzles.

The present work is aimed to study the air entrainment by using plunging water jet in co-current downflow bubble column. Gas holdup, bubble velocity and bubble diameter have been evaluated for air-water system under different rates of both phases using nozzles of 7, 8 and 9mm diameters.

2. Experimental

2.1. Experimental Apparatus

A Perspex column of 50 mm diameter 152 cm length is used as a test contactor as shown in Figure 1. The upper end of the contactor is connected to the brass nozzle. Three electro-resistivity probes, P_1- P_3, are located at different positions along the column, in order to give hydrodynamic measurements for the three expected zones throughout the column. The electrodes are connected to the interface to compute of number, size, and the speed of the bubbles at specified working time.

A rectangular Perspex vessel of 320 × 320 mm size and 900 mm height has been used as air-liquid separator. The separator is provided with two outlet points for venting gas and draining liquid. A straight glass tube, L, has been connected between the lower part of the separator and the top of the column for measuring the height of clear liquid corresponding to the two-phase height in the column.

2.2. Experimental Procedure and Observations

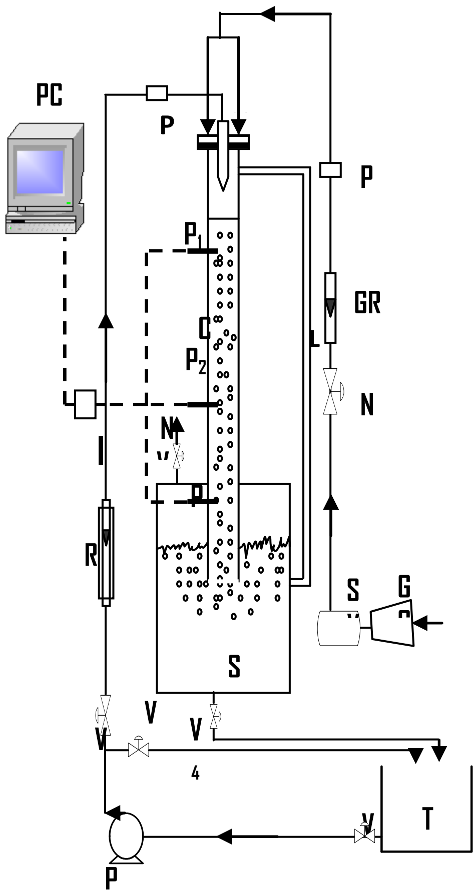
The selected nozzle and extended pipe line contactor is fitted properly before each experimental run to achieve an axially symmetric liquid jet. Water is pumped from the storage tank (T) via a centrifugal pump and it is substantially emerged from the nozzle and flows downward through the center of the pipeline contactor. The valves V_4 and NV_2 are initially kept fully open and liquid jet directly hits the bottom of the air-water separator and then returns to the storage tank through valve V_4.

The valve V_4 is then closed and the liquid is allowed to flow as a jet and accumulated in the separator. When the separator is filled with water to a certain height, the valve opening is adjusted to maintain the height in the separator at the desired fixed level. This process is continued so that the liquid level is increased until touches the end of the column, and a sudden change in the process is reached. Meanwhile, the accumulated liquid in the column is directly increased and the liquid level could be fixed by adjusting the valve V_4. Two distinct zones are clearly observed, viz., the intense gas-liquid mixing zone followed by a downflow fine bubble zone. The pressure in the upper space of the separator can be increased by controlling the gas-flow rate from the separator through the valve V_4 so that the level of gas-liquid mixture in the column should be still at the desired point. The operation range is limited so that the height of the two-phase mixture should not go up to the end of the contactor.

When the operation is at steady state at certain gas and liquid flow rates the overall gas holdup can be obtained by measuring the total gas-liquid mixing height (h_m) in the contactor and the corresponding clear liquid height (h_L) in the arm (L), as in equation (1).

2.3. Bubble Monitoring and Analyzing System

The resistivity probe technique [14,15] is used in the present work to measure the local gas void fraction, bubble velocity, number and the size of bubbles. This measuring system consists of double sensor probe, interface, computer and software program (Visual Basic).



C	Contactor
N	Nozzle
GR	Gas rotameter
L	Arm of clear liquid
PG_1-PG_2	Pressure gauges
Pu	Pump
R	Rotameter
SE	Separator
NV_1-NV_2	Needle valves
T	Storage tank
V_1-V_4	Valves
SV	Stabilizer vessel
P_1-P_3	Probes
GC	Compressor
IF	Interface
PC	Computer

Fig. 1: Schematic diagram of the down-flow bubble column

3. Results and discussion

3.1. Gas holdup by height measurements

The effect of superficial gas velocity on the overall gas holdup with nozzles of 7, 8 and 9mm diameter are properly illustrated in Figures 2, 3 and 4 respectively. These figures indicate that gas holdup is progressively increased with increasing superficial gas velocity at a constant liquid flow rate. The values of gas holdup are generally ranged between (0.2-0.53) which are significantly higher than those reported by other workers [9,16].These results may be explained according to the fact that the gas bubbles in co-current downflow system are moved against their higher buoyancy force, so that the bubbles have a higher slip velocity compared to upflow system. Accordingly, the bubbles have longer residence time and hence higher gas holdup is observed.

It is worthy to mention that, bubble flow is characterized by the dispersion of bubbles and the free space available between them. If the bubbles are small enough (d_B< 3 mm) and there is sufficient free spaces between them, the bubbles population increases considerably with increasing gas flow rate and hence gas holdup increases [12].

However, the gas flow rate has little effect on bubble population beyond certain limit and coalescence of bubbles is significantly increased, so the gas holdup remains constant [17, 18, 19].

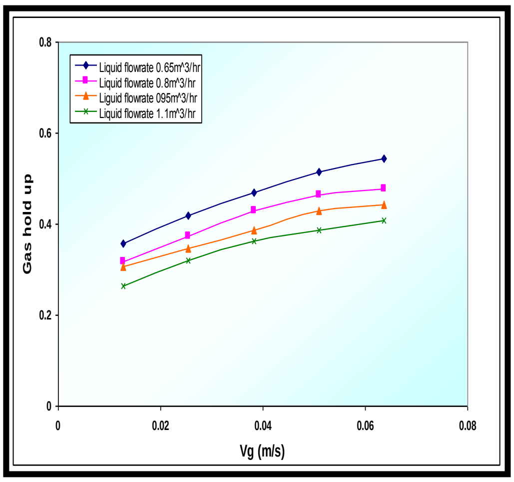


Fig. (2) Variation of overall gas holdup with superficial gas velocity at different liquid flow rates with nozzle of 7 mm diameter

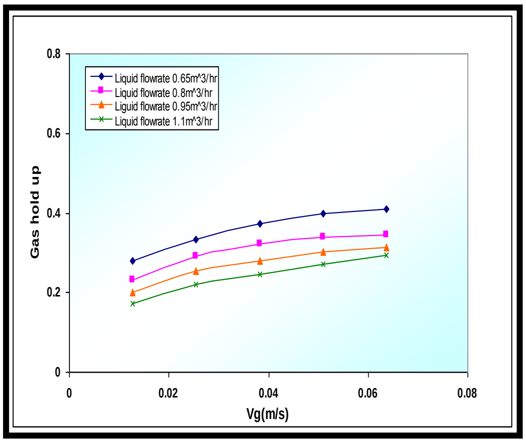


Fig. 3 Variation of overall gas holdup with superficial gas velocity at different liquid flow rates with nozzle of 8 mm diameter

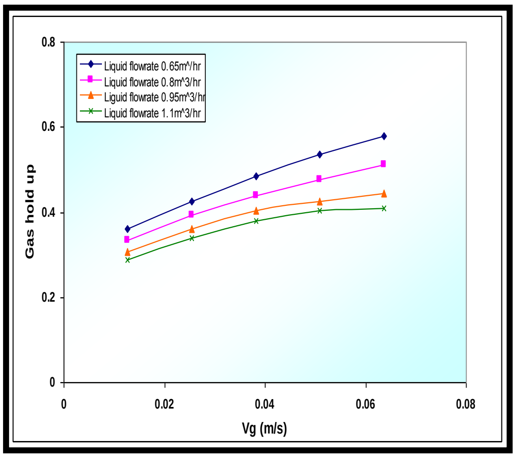


Fig.4 Variation of overall gas holdup with superficial gas velocity at different liquid flow rates with nozzle of 9 mm diameter

It is important to note that bubbles observed in the present work for co-current downflow system, are relatively of bigger size (around 2-4 mm diameter) compared to those observed by other workers (around 1-2 mm diameter) which can be related to the higher duration of bubbles in the column. Consequently, higher gas holdup will obtain. A comparative view of gas holdup obtained by different authors with different flow arrangements and over ranges of gas and liquid flow rates is given in Table 1. It is important to note from Table 1. that the values of the average gas holdup in upflow system are comparatively lower than those reported for downflow arrangement. Bando et al.[20] observed a maximum gas holdup around 0.32 in downflow bubble column with simultaneous gas-liquid injection nozzle system. Similarly, Yamagiya et al.[21] and Ohkawa et al.[22] obtained the maximum value of around 0.4 in downflow system but the superficial gas velocities in their experiments were much higher than those observed in the present work. However, a relatively higher gas holdup (0.2-0.53) was obtained in the present work with air-water system even at low gas flow rate. Hence, this system with proposed operating conditions can be alternatively used in chemical processes wherein interfacial area plays a dominating role. Moreover, it is clearly seen from the aforementioned plots that gas holdup is progressively increased with decreasing liquid flow rate at the same superficial gas velocity. This is obviously related to the higher slip velocity of the bubbles and increasing their residence time in the contactor, so that the gas holdup increases to a certain limit.

Table (1) Comparative view of gas holdup of the present work with the results of other studies

Authors	**Type of flow**	**Column diameter (m)**	**Liquid tested**	**Liquid velocity V_L (m/s)**	**Gas velocity V_G (m/s)**	**Gas holdup ε_g (-)**
Godbole et al. [17]	Batch with gas upflow	0.1	CMC	-	0.05-0.3	0.1-0.28
Schumpe &	Co-current	0.1-0.14	CMC, (0-	-	0.003-	0.03-0.2

Deckwer[18]	upflow		1.8) wt %.		0.025	
Khatib & Richardson[23]	Co-current upflow	0.039	Kaolin suspension	0.305-0.61	0.3-3.5	0.15-0.55
Ohkawa et al. [22]	Co-current downflow	0.02-0.026	Water	0.05-0.2	0.05-0.2	0.01-0.4
Bando et al. [20]	Co-current downflow	0.07	Water	0.10-0.2	0.01-0.10	0.01-0.32
Yamagiya et al. [21]	Co-current downflow	0.034-0.07	Water	0.4-0.912	0.1-0.5	0.15-0.4
Das et al. [24]	Co-current horizontal	0.019	CMC (0.5-1.00) kg/m^3	0.141-1.00	0.067-1.55	0.1-0.4
Das et al. [25]	Co-current upflow	0.019	CMC (0.5-1.00) kg/m^3	0.29-1.00	0.17-1.6	0.12-0.45
Zahradnik et al. [19]	Co-current upflow	0.29	Water	0.008-0.029	0.004-0.076	0.05-0.24
Present work	Co-current downflow	0.05	Water	0.091-0.155	0.012-0.063	0.2-0.53

On the other hand, the effect of nozzles diameter on the overall gas holdup at different liquid flow rates are shown in Figures (4) and (5) for lower and higher liquid flow rates respectively. However, results are found with higher flow rates represented elsewhere[26]. It is clearly observed from these plots that lower gas holdup is obtained with smaller nozzle diameter throughout the operating range of gas velocity under any specific liquid flow rate. This result can be explained according to the fact that nozzle with smaller diameter produces bubbles of relatively small size which move downward faster due to lower buoyancy force. This result is in quite agreement with the observation of Bando et al.[20] who showed similar trend of gas holdup variation in co-current down flow bubble column with simultaneous gas liquid injection nozzle system.

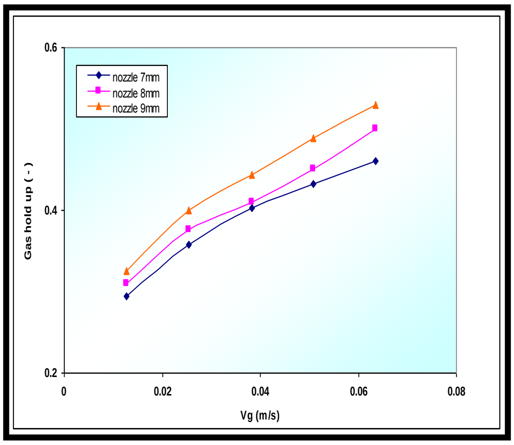


Fig. 4 Variation of overall gas holdup with superficial gas velocity using nozzles of different diameters at liquid flow rate of 0.65 m^3/hr.

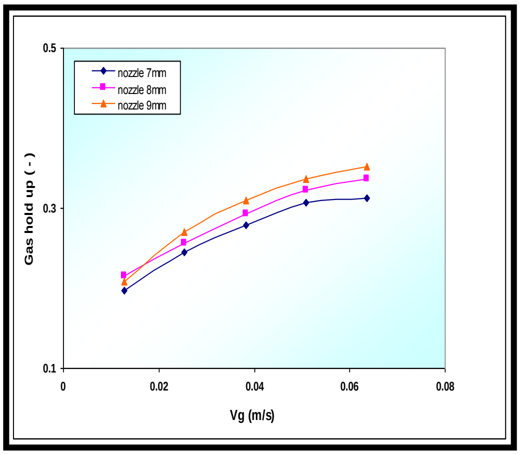


Fig.5 Variation of overall gas holdup with superficial gas velocity using nozzles of different diameters at liquid flow rate of 1.1 m^3/hr

3.2. Gas holdup by resistivity probe technique

The variation of local gas holdup, measured by resistivity probe, with superficial gas velocity at different liquid flow rates with nozzle of 9mm diameter for both mixing and two-phase flow zones are shown in Figures (6)-(7) respectively. However, plots are observed with other nozzles and presented elsewhere [26]. These plots are apparently similar to those of overall gas holdup, i.e. the local gas holdup is progressively increased with increasing the superficial gas velocity. On the other hand, it is important to note that the values of gas holdup in two-phase zone is considerably higher than the corresponding values in mixing zone for specific gas and liquid flow rates. This observation is attributed to the generation of fine and relatively small size bubbles in mixing or intensing zone so that the buoyancy force and the residence time is being lower in this zone. Accordingly, the gas holdup in mixing zone is evidently lower than the corresponding values in two-phase zone.

It is important to mention that the overall gas holdup represents the average of the values of local gas holdup throughout the column. Consequently, the values of the overall gas holdup determined by height measurement are approximated and in between the values of local gas holdup obtained by the resistivity probe technique.

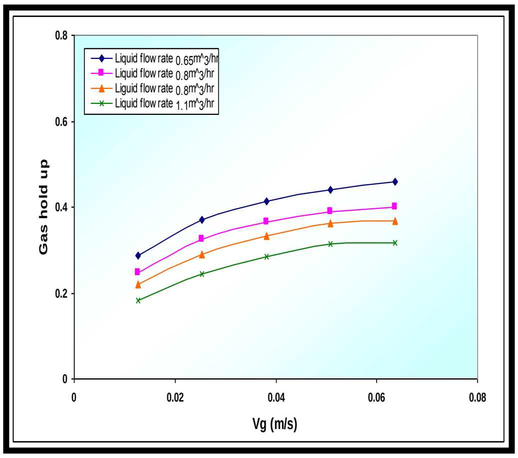


Fig. 6 Variation of local gas holdup measured by resistively probe with superficial gas velocity for *"mixing zone"* at different liquid flow rates with nozzle of 9mm diameter

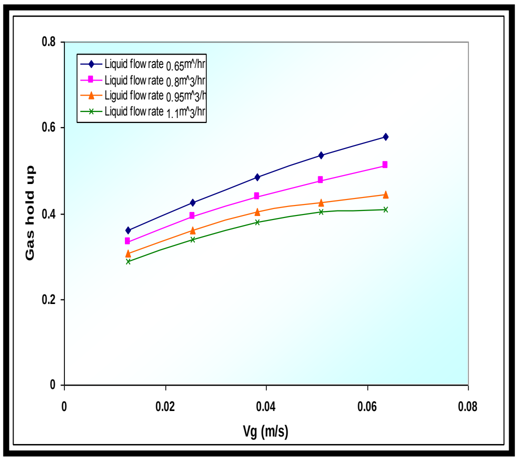


Fig.7 Variation of local gas holdup measured by resistively probe with superficial gas velocity for *"two-phase zone"* at different liquid flow rates with nozzle of 9mm diameter

3.3. Correlation of gas holdup

Mathematical correlation has been developed from the experimental data by dimensional analysis to predict gas holdup within dispersed phase in terms of physical, dynamic and geometric variables of the system.

Gas holdup can be expressed as function of the following parameters;

$$\varepsilon_g f\left(V_L, V_g, \rho_L, \mu_L, \rho_g, \mu_g, d_c, d_n, h_m\right) \quad \text{... (2)}$$

The following correlation for estimation of gas holdup is observed:-

$$\varepsilon_g = 102.16(\mathrm{Re}_{Ln})^{-0.699}(\mathrm{Re}_g)^{0.277}(a_r)^{-0.144} \quad \text{... (3)}$$

where:- Re_{Ln} is Reynolds number of liquid based on nozzle diameter, Re_g is Reynolds number of gas based on column diameter, and a_r is nozzle to column area ratio.

Numerical analysis is applied on the experimental data observed under different operating conditions and statistical evaluation parameters of equation (3) gives correlation coefficient = 0.995412 with variance of 0.974982.

The calculated values of gas-holdup according to equation (3) are plotted against the corresponding experimental values obtained by height measurement technique as shown in Figure (8).

This plot evidently indicates that most of the data are fitted the suggested correlation, and are exactly located on the diagonal with uniform and very little scattering.

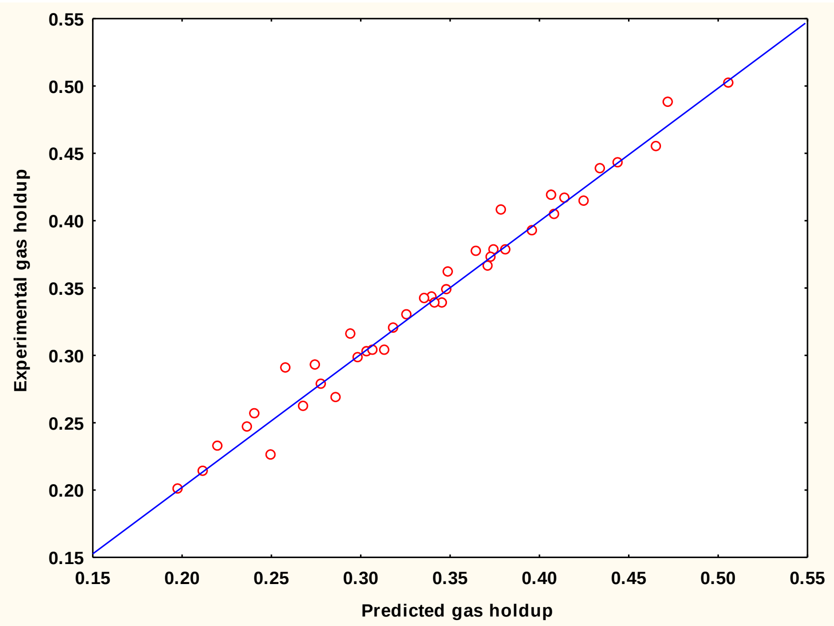


Fig.8 Graphical evaluation of gas holdup correlation

3.4. Bubble Velocity

The variations of bubble moving velocity with superficial gas velocity at different liquid flow rates with 9mm nozzle diameter for both two-phase and mixing zones are shown in Figures (9) and (10) respectively. However, similar results have been obtained with other nozzles, and were presented elsewhere [(26)].

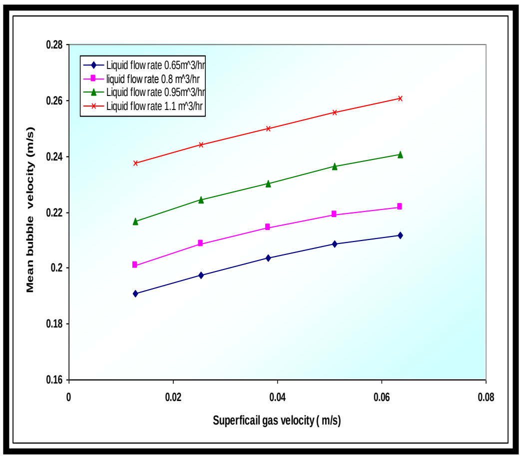


Fig.9 Variation of mean bubble velocity with superficial gas velocity at different liquid flow rates with nozzle of 9mm diameter for (two-phase zone region)

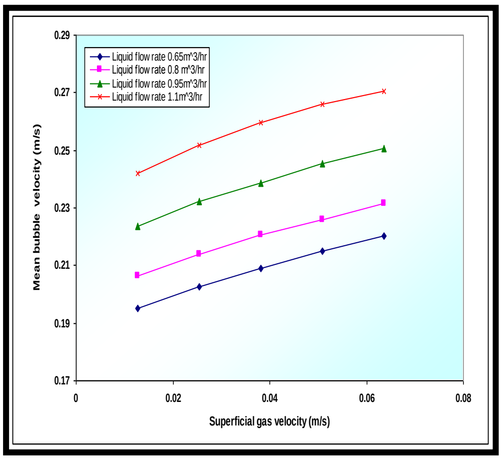


Fig.10 Variation of mean bubble velocity with superficial gas velocity at different liquid flow rates with nozzle of 9mm diameter for (mixing zone region)

These figures indicate that bubble moving velocity is linearly increased with increasing superficial gas velocity. This increasing is analyzed by the well known Drift Flux Model of Zuber and Findlay[27]. This model is expressed as:-

$$u_B = C_0\left(V_g + V_L\right) + V_d \qquad \dots (4)$$

where u_B is the mean bubble velocity.

V_g & V_L are the superficial gas and liquid velocities respectively.

V_d is the mean gas drift velocity.

C_o is the distribution parameter.

The sum of superficial gas and liquid velocities is often termed as gas-liquid mixture velocity (V_m). Accordingly, equation (4) can be rewritten as:-

$$u_B = C_0 V_m + V_d \qquad \dots (5)$$

Zuber and Findlay equation represents a straight line relationship between the mean bubble velocity (u_B) and the gas-liquid mixture velocity (V_m). The slope of this line represents the distribution parameter, C_0, which accounts the effect of the flow irregularity and concentration profiles; whereas the intercept is being the mean gas drift velocity, V_d which accounts for the effect of local relative velocity and it is equivalent to unhindered bubble rise velocity, $V_{b,\infty}$.

The experimental data for the uniform two-phase zone are plotted according to equation (5) as shown in Figure (11). The data points seem to be uniformly distributed around the straight line with little scatter. Statistical evaluation (5) gives correlation coefficient of 0.926 with variance of 0.857. The distribution parameter (C_0) is 0.76 and the mean gas drift velocity (V_d) is 0.0622 m/s. According to these correlation coefficients, an empirical equation has been found as:

$$u_B = 0.76\, V_m + 0.0622 \qquad \dots (6)$$

Where: u_B is the mean bubble velocity and V_m is superficial velocity of gas – liquid mixture.

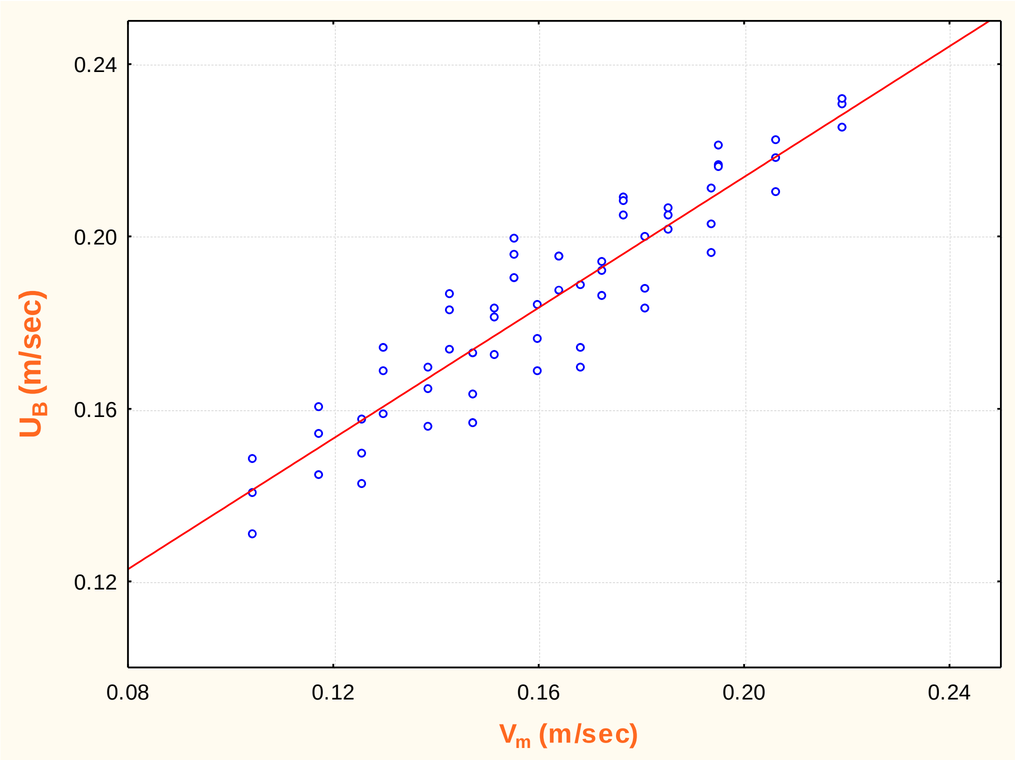


Fig.11 Zuber – Findaly Correlation plot for the uniform two – phase zone

This result proves that the present data satisfactorily fits Zuber and Findlay equation. It is clearly shown that the distribution parameter is less than unity i.e. 0.76. Zuber and Findlay[27] agreed that velocity profile becomes flatter with increasing liquid flow rate which leads to reduction in the value of C_o so that its value is apparently less than unity. This result indicates that gas phase is uniformly dispersed across the column area. On the other hand, when the value of C_o becomes more than unity, it gives an indication for parabolic velocity profile so that the centerline velocity is higher than the mean velocity and the gas phase is preferably aggregated in the center of the column.

Numerous investigators [28, 21, 25, 29] have attempted to quantify the values of the distribution parameter and drift velocity. Their results are found to be in good agreement with the results of the present work. On the other hand, Figures (9) to (10) reveal that bubble velocity in mixing zone is higher than that in the uniform two-phase zone at any specific liquid flow rate and nozzle diameter. This result is undoubtedly ascribed to the bubble size. It is well known that smaller bubbles are often found in mixing zone which has lower buoyancy force and higher velocity comparing with relatively larger bubbles that commonly exist in uniform two-phase flow zone.

3.5. Bubble Diameter

The variations of bubble diameter with superficial gas velocity at different liquid flow rates with nozzle 7mm diameter shown in Figures (12) and (13) for both two-phase and mixing zones respectively. However, same plots are appeared with other nozzles, and are presented elsewhere [26].

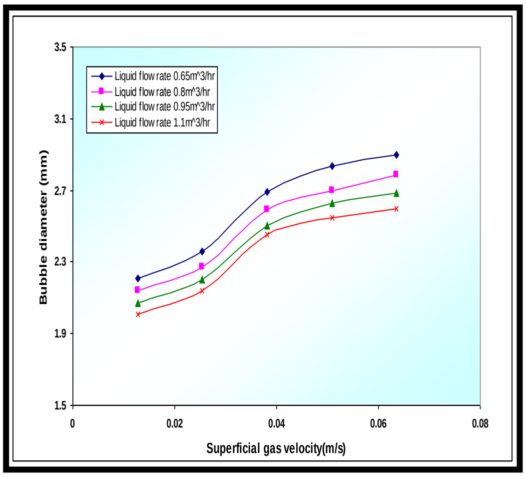


Fig. 12 Variation of bubble diameter with superficial gas velocity at different of liquid flow rates with nozzle of 7mm diameter for two-phase zone

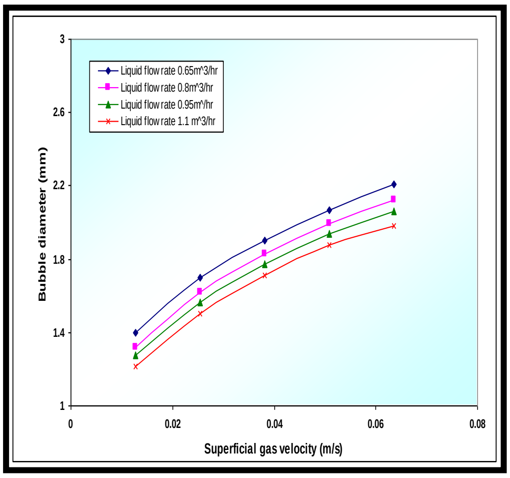


Fig. 13 Variation of bubble diameter with superficial gas velocity at different liquid flow rates with nozzle of 7 mm diameter for mixing - zone

It can be seen from these plots that the measured bubble diameter is considerably increased with increasing superficial gas velocity at any specific liquid flow rate. The result is attributed to the increasing of the volume of mixing zone with increasing the superficial gas velocity, while the energy input from the jet remains constant. Accordingly, the energy input per unit volume will be decreased and resulting an increase in the bubble diameter.

Moreover, these figures indicate that bubble diameter is progressively increased with decreasing liquid flow rate. This result is ascribed to the energy input at plunging jet which is proportional to the square of the jet velocity. The inertia force created by the jet is intuitively increased with increasing liquid flow rate. The higher downflow liquid stream prevents the bubble coalescence, so that small bubbles will be produced.

On the other hand, these plots reveal that bubble diameter is slightly influenced by the nozzle size. However, nozzle with small diameter relatively produces small size bubbles. This is ascribed to the higher energy input by the jet stream with small size nozzle at any specific liquid flow rate i.e. more inertia will input to the system which generates small bubbles.

However, an interesting observation can be seen in Figure (12) for two-phase flow region. The increasing of bubble diameter with increasing superficial gas velocity is being more significant at intermediate range of velocity, viz., between 0.02-0.04 m/sec. Thereafter, the increasing in bubble diameter is extremely limited i.e. the bubble diameter seems to be stable and less dependent on superficial gas velocity. This result can be explained according to the fact that bubble diameter increases to a certain limit with increasing superficial gas velocity. Population of bubbles facilitates the coalescence with each other to give comparatively larger bubbles. The bubbles become more stable beyond a specific size due to the balance exists between the buoyancy force of bubbles and the force of downflow stream of liquid, hence the coalescence of bubbles is highly restricted.

The values of bubble diameter observed in the present work are ranged between 2-4 mm which are similar to the result of Jonathan et al.[30], who observed bubbles of 3-4 mm diameter using an acoustic monitoring technique for air-water system. However, Mandal et al.[(29)] observed bubbles of larger size, viz., 3-5 mm diameter for down flow bubble column with air-water system.

Conclusion

Three dispersion zones are distinguished according to the visual observation: (1) the top mixing zone where the liquid jet plunges with the simultaneous formation and the breakup of gas bubbles, (2) the middle zone of a homogeneous bubbly flow and higher gas hold up of bigger bubbles, and (3) the bottom zone wherein a decrease in bubble population and gas hold up. Gas hold up is found to be progressively increased with increasing superficial gas velocity and with decreasing liquid flow rate. Lower gas hold up is obtained with smaller nozzle diameter at any specific liquid flow rate throughout the operating range of gas velocity. A significant back mixing is particularly observed in the plunging zone which leads to the regeneration of fine gas bubbles and enhances the overall gas hold up compared with up flow systems. The following correlation is developed by dimensional analysis for estimation of gas holdup:

$$\varepsilon_g = 102.16\,(Re_{Ln})^{-0.699}\,(Re_g)^{0.277}\,(A_r)^{-0.144}$$

The mean bubble velocity is linearly increased with increasing superficial gas velocity. The data satisfactorily fit the Drift Flux model of Zuber and Findlay according to the relation:

$$u_B = 0.76\,V_m + 0.0622$$

The mean bubble velocity is apparently increased with increasing liquid flow rate and nozzle diameter for both mixing and two – phase zones. However, it is found that the mean bubble velocity in mixing zone is evidently higher than that in two-phase zone at any specific liquid flow rate and nozzle diameter. The bubble diameter is considerably increased with increasing superficial gas velocity, whereas it is slightly influenced by nozzle size. Moreover, bubble diameter is found to be progressively increased with decreasing liquid flow rate. However, the bubbles in the two-phase zone are relatively bigger than those observed in the mixing zone.

Notation

ar	nozzle to column area ratio, $(d_n/d_c)^2$, Dimensionless
dc	diameter of the Column , m.
dn	diameter of the nozzle , m.
hm	total gas-liquid mixing height, m.
Qg	volumetric flow rate of gas , m^3/s.
QL	volumetric flow rate of liquid , m^3/s.
Qr	volumetric flow rate of gas to liquid , Q_g/Q_L ,Dimensionless.
Re g	Reynolds number of gas based on column diameter, $\frac{\rho_g V_g d_c}{\mu_g}$, Dimensionless.
Re Ln	Reynolds number of liquid based on nozzle diameter, $\frac{\rho_L V_{Ln} d_n}{\mu_L}$, Dimensionless.
uB	mean bubble velocity, m/s.
Vg	superficial velocity of gas phase , m/s
VL	superficial velocity of liquid phase, m/s.
Vm	superficial velocity of gas – liquid mixture, m/s.

Greek Symbols

ε_g	Gas holdup , dimensionless
μ_g	Viscosity of gas , kg/m.s
μ_L	Viscosity of liquid , kg/m.s
ρ_g	Density of gas , kg/m^3
ρ_l	Density of liquid , kg/m^3
ε_g	Gas holdup , dimensionless

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