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Investigation of Gas Hold-up and Bubble Behavior in a Split-Cylinder Airlift Reactor: Pseudo-Plastic Non-Newtonian Fluids

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Abstract

In this research, the effect of carboxy methyl cellulose (CMC) addition into pure water as pseudo-plastic non-Newtonian fluid and its concentration on bubble diameter and gas hold-up were investigated. For this purpose, four different concentrations of CMC (0.05, 0.1, 0.15 and 0.2%w/v) as the non-Newtonian fluid and five different superficial gas velocities (0.2, 0.4, 0.6, 0.8 and 1 cm/s) as the gas phase were examined in an airlift reactor. Bubble size distribution in the airlift reactor was measured by photography and picture analysis at various concentrations of CMC and various velocities of gas. Increasing in gas velocity created a wider bubble size distribution and thereby an increase in bubble diameter and gas hold-up in both riser and down-comer. However, the bubbles diameter in pure water was larger than those of the CMC solutions (in the riser and down-comer), but CMC concentration enhancement increased bubbles diameter and gas hold-up in the down-comer. Bubbles diameter expansion in the riser by CMC concentration enhancement took place from concentrations of 0.05 to0.15% (w/v) and then it suddenly decreased. Furthermore, gas hold-up decreased from concentrations of 0.05 to 0.15%(w/v) and increased at concentration of 0.2%(w/v).The gas hold-up increases (more than that in the concentrations of 0.1 and 0.15 %) when bubbles diameter decreases in concentration of 0.2%. The overall gas hold-up trend was similar to the gas hold-up in the riser.

Keywords: *Airlift Reactor, Bubble Diameter, Gas Hold-up, Non-Newtonian, Pseudo Plastic*

1. Introduction

Airlift reactors are the best types of two phase contactors in different processes such as wastewater treatment, animal cell culture and aerobic fermentation (production of enzymes, antibiotics, proteins, biomass and other biotechnology products) [1].

Liquid circulation in an airlift reactor is caused by the difference in gas hold-up between the riser and the down-comer zones [2, 3]. This difference and therefore, the rate of induced liquid circulation can be enhanced greatly by installing some form of a gas–liquid separator at the top of the reactor so that the liquid returning to the down-comer zone is mostly free of the entrained gas bubbles. Notwithstanding the beneficial effect of gas–liquid separators, airlift reactors are commonly operated without them as only a small difference between the gas hold-up in the riser and down-comer zones is sufficient to generate liquid circulation. In most

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applications the only controlled variable in an airlift reactor is the gas flow rate, or superficial velocity of the gas in the riser channel. Gas velocity in turn affects gas hold-up, the rate of induced liquid circulation, and the gas–liquid mass transfer rate. Gas–liquid hydrodynamics and mass transfer in airlift reactors have been studied extensively as reviewed elsewhere [4].

With the developments in mineral recovery, food processing, biomedical engineering, biochemical engineering and waste-water treatment, situations increasingly arise where the liquid behaves as non-Newtonian in nature. Although many industrial liquids which comprise solutions of low molecular weight can be considered Newtonian-like fluids, an increasing number of solutions with high molecular weights and internal structure are being used that have non-Newtonian behavior such as variable viscosity and memory effects [5]. Polymer solutions and melts, liquid crystals, gels, suspensions, emulsions, micellar solutions, slurries and foams enter into this non-Newtonian category [6]. Hence, there is a need to study airlift reactor using non-Newtonian. According to the literature, the majority of research has been conducted into the airlift reactors with Newtonian fluids and there are few researches on non-Newtonian or high viscosity liquids though there are numerous applied non-Newtonian systems.

Metkin and Sokolov [7] derived the wall shear stress for a pseudoplastic non-Newtonian gas-liquid dispersion as a function of the Reynolds number and rheological parameters. The authors identified the transition from laminar to turbulent flow condition using the critical Reynolds number.

Shi *et al*. [8] evaluated the average shear rate for non-Newtonian liquids in an external loop airlift reactor by analogy with the down-comer liquid velocity of Newtonian liquids, and proposed an empirical quadratic equation for the effective shear rate. They found that effective shear rate is lower in airlift reactors than in bubble columns.

Li *et al*. [9] studied the influence of non-Newtonian fluid on the hydrodynamics and mass transfer using a wide range CMC concentration of 1–4% in an internal airlift reactor. However, this work was limited to experimental determination of kla and the reactor used a single-hole sparger, which is not usually adopted in industrial reactors. Al-Masry [10] investigated the contributions of pressure drop due to wall frictional losses to the total gas holdup of two phase viscous non-Newtonian systems (xanthan gum and carboxymethyl cellulose) in a circulating bubble column. Pressure drop due to wall shear stress was found to significantly contribute to the total gas holdup by 10-70%.

Kilonzo and Margaritis [11] have reviewed the effects of non-Newtonian fermentation broth viscosities on gas–liquid mixing and oxygen mass transfer characteristics in airlift bioreactors.

Mouza *et al*. [12] studied the effect of liquid properties (surface tension and viscosity) on the performance of bubble columns. They also developed a correlation for gas holdup in the homogeneous regime.

Khamadieva and Böhm [13] investigated the effect of liquid viscosity on mass transfer at the wall of packed and un-packed bubble columns using Newtonian and non-Newtonian liquids.

Cerri *et al*. [14] evaluated oxygen transfer in three internal-loop airlift reactors of different working volumes and similar geometric configuration utilizing eight Newtonian and five non-Newtonian fluids. They reported that the effects of the superficial gas velocity and liquid viscosity had opposite effects on the volumetric oxygen transfer coefficient and the viscosity effect on oxygen mass transfer cannot be neglected.

Deng *et al*. [15] studied hydrodynamics and mass transfer in a 5m internal-loop airlift reactor with non-Newtonian CMC solution. They found that increase in aeration velocity or CMC concentration led to a wider bubble size distribution and an increase in the bubble Sauter diameter. The volumetric mass transfer coefficient increased with an increase in aeration velocity and a decrease in CMC concentration.

Velez Cordero and Zenit [5] measured mean rise velocity of bubbles warm ascending in shear-thinning fluids in a rectangular bubble column. They found that the mean rise velocity of the bubbles was larger than that of an individual bubble, in accordance with previous studies. Furthermore, they found that the appearance of clusters produced a dramatic increase of the agitation within the column.

In another research, Anastasiou *et al*. [16] studied the performance of a bubble column equipped with metal porous sparger when the liquid phase is a shear-thinning non-Newtonian fluid. Their aim was to formulate a generalized correlation that can be applied for predicting the average gas of these kinds of bubble column.

In this research, the effects of carboxy methyl cellulose (CMC) as pseudo-plastic non-Newtonian fluid as criteria of fungal fermentation broth and its concentration onthe bubble diameter and gas hold-up in the riser and down-comer of split cylinde rairlift reactor were investigated.

2. Experiment

2-1. Materials

CMC was purchased from Merck Company (Germany) and its various concentrations (0.05, 0.1, 0.15 and 0.2 %w/v) were locally prepared. Table 1 shows the physical properties of the liquid phase used in this work. The liquid densities were measured using a balance hydrometer and none were found to be significantly different from that of water because the concentration of these polysaccharides is very low.

Table 1. Physical properties of the aqueous solutions of carboxymetyl cellulose.

CMC (%w/v)	Density (kg/m^3)	Surface tension(mN/m)
0% (tap water)	998.2	72.75
0.05%	1000.5	73.00
0.10%	1001.2	74.00
0.15%	1001.3	74.50
0.20%	1001.5	75.50

The CMC solutions were assumed to be pseudo-plastic and were characterized by the Ostwald-de-Waele power law model, $\tau = K\gamma^{n}$, where τ, γ, K and n are shear stress, average shear rate, power law fluid consistency index and flow behaviour index, respectively.

The effective viscosity was defined as, $\mu_{eff} = K\gamma^{n-1}$. The K and n values of the CMC solutions and viscosity of the other liquids (μ_L) were determined with the help of a Brook field cone plate viscometer (Model RVTDV-ICP) from the respective flow curves. The values of the power law constants K and *n* were obtained by regression of the viscometer results. Each CMC solution was used within a day to ensure that viscosity had not been reduced by biodegradation.

In order to calculate the effective viscosity of a non-Newtonian fluid in the airlift reactor the effective average shear rate in the airlift reactor, the relation for the shear rate proposed by Nishikawa *et al*. [17] for bubble columns for this purpose was used. According to Nishikawa *et al*. [17] the relation is:

$$\gamma_{ave} = 5000\ U_g$$

Where γ_{ave} is average shear rate and U_G is the superficial velocity (m/s), respectively.

Surface tension (σL) reported in Table 1 was measured using a ring tensiometer (model K10ST, KRUSS GmbH, Hamburg, Germany). Compressed and oil-free air was used as the gas phase in all experiments. The air superficial velocity was varied from 0.2 to 1ms^{-1}.

2-2. Equipment

The split-cylinder airlift reactor used in this research is schematically shown in Fig. 1. The reactor includes a glass column with 1.3 m height and 0.136 m diameter. A rectangular Plexiglas baffle with 0.129 m width, 1.0 m height and 0.005 m thickness was inserted in the glass column to divide the cross section into riser and down-comer zones (with the riser area of 86.115 cm^2 and down-comer area of 40.299 cm^2 and the ratio of 2.136 riser per down-comer). The baffle was also located at 0.1 m from the bottom of the reactor. The gas free liquid height in the column was about 1.23 m for all experiments. The gas sparger was a 0.015 m diameter sintered ceramic ball located at the bottom of the riser. The volumetric flow rate of air in the riser zone was controlled using a regulating valve and a calibrated rotameter. The inverted U-tube manometers were used for hold-up measurement. All experiments were carried out at ambient conditions (atmospheric pressure and (25±0.5) °C).

2-3. Measurement methods

The steady-state bubble diameter size was determined with photographic technique by a digital camera (Panasonic, model: DMC-FS20 with resolution of 10M pixels). The moving average method [18] was used to count number of bubbles which was more than 300 bubbles that were randomly chosen in 10 pictures. Then, bubble diameters were measured by WINDIG software (version 2.5).

In spherical bubbles d_1 is equal to d_2 (so d_v=d_1=d_2), while elliptical bubbles were measured according to the maximum and minimum diameters of bubbles as follows:

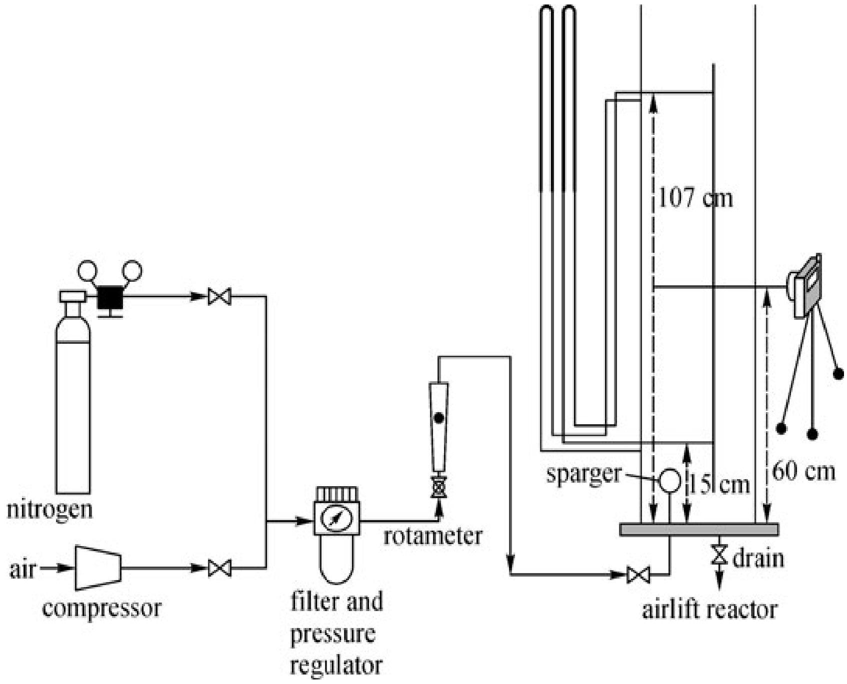

Figure 1. Schematic diagram of the split-cylinder airlift reactor.

$$d_v = \sqrt[3]{d_1^2 d_2} \tag{1}$$

Where d_1, d_2 and d_v are the maximum, minimum and equivalent diameters, respectively.

The average of bubbles diameter (d_{ave}) is calculated as:

$$d_{ave} = \frac{\sum_{i=1}^{i=N} d_i^3}{\sum_{i=1}^{i=N} d_i^2} \tag{2}$$

d_i is bubble diameter and I is the number of bubbles.

The volume expansion method was applied to measure the overall gas hold-up during the steady state condition. The overall gas hold-up (ε_g) was recorded by visual measurements using the volume expansion method [19].

The gas hold-up in the riser and down-comer sections was determined with manometric technique [20] by measuring the differential pressure between two inverted U-tube manometers (Fig. 1). The gas hold-up (ε) was calculated as:

$$\varepsilon = \frac{\rho_L}{\rho_L - \rho_G} \frac{dh_M}{dz} \tag{3}$$

where, ρ_L is the density of fluids, ρ_G is the density of air, dh_M is the manometer reading and dz is the space between the manometer taps.

3. Results and discussion

3-1. Bubble diameter measurement

As shown in Table 1, the surface tension increased by increasing the CMC solution concentration. With decreasing the surface tension, bubble volume will increase [21]. CMC molecules accumulate on the bubble surface and increase the required energy for bubble breakup when the bubbles move within the continuous phase. Furthermore, the densities of solutions increase with CMC concentrations. This increases the buoyancy force. Therefore, liquid phase mass increases with the growth of bubble and drag force enhances, consequently the growth time of bubble is extended [22]. When bubbles coalescence rate increases and instability occurs up to a critical point, instability reaches a maximum amount. At this condition, bubbles breakup rate suddenly rises to overcome the bubbles coalescence [23, 24]. As shown in Fig. 2 bubbles breakup and critical point in the riser was obtained at concentration of 0.15 %. Therefore, bubbles breakup rate increased and bubble diameter decreased in this concentration. As illustrated in Fig. 3, bubble diameter increases with concentration enhancement in the down-comer because the bubbles diameter cannot reach a maximum amount. Since the buoyancy force decreases in comparison with the drag force for smaller bubbles, these bubbles can import the down-comer. The bubbles average is shown in Figs. 2 and 3 as plotted by MATLAB software (version 7.6.0.324). The bubbles diameter increased in the riser (Fig. 2) and down-comer (Fig. 3) when aeration velocity increased.

The CMC concentration enhancement increases the viscosity of solution. Therefore, the drag force of fluid around the bubble obstructs the bubble rising [25]. This causes surface tension to increase for CMC solutions against water and large bubbles break. The bubble diameter increases when surface tension of solution increases. The surface tension, viscosity and density increase when CMC is added to water. So,

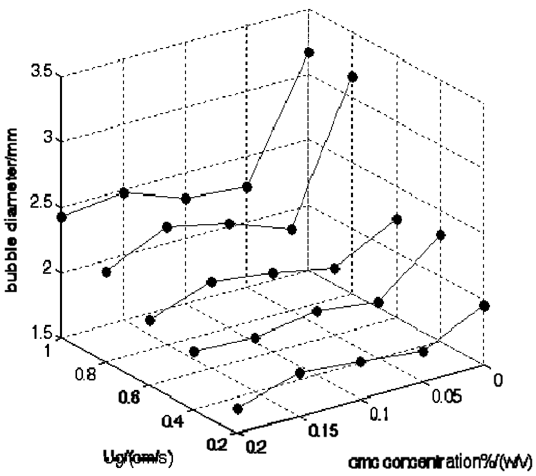

Figure 2. Average of bubbles diameter with different aeration velocity and CMC concentration in the riser.

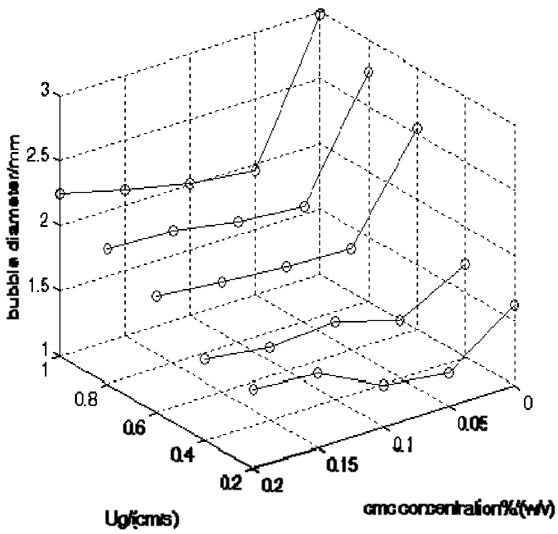

Figure 3. Average of bubbles diameter with different aeration velocity and CMC concentration in the down comer.

the bubble diameter initially reduces and then rises. The turbulence in the riser decreases and the bubble size distribution increases when the solution viscosity increases [20]. Figs. 4 and 5 show the influence of CMC and its concentration enhancement on bubble diameter in the riser and down-comer, respectively. As illustrated in these figures, the bubble diameter reduced from 1.8 to

1.5mm and 2.3 to 1.2mm in the riser and down-comer, respectively when CMC concentration and thereby density and viscosity increased. Around 10% of bubbles in pure water have the same size while about 20 and 30% of bubbles in the riser and down-comer respectively have the same size for CMC solutions.

Figs. 6 and 7 show the effect of aeration velocity on bubble diameter in the riser and down-comer for CMC concentration of 0.15% (w/v). The bubble diameter was about 1.8 and 1.6mm in the riser and down-comer, respectively, in the high aeration velocities.

The number of bubbles which are smaller or bigger than the average diameter size is almost equal in the riser. They also have various diameters (from 1.2 to 2.3mm). Furthermore, there are not much bigger bubbles in the down-comer (from 1.4 to 2mm).

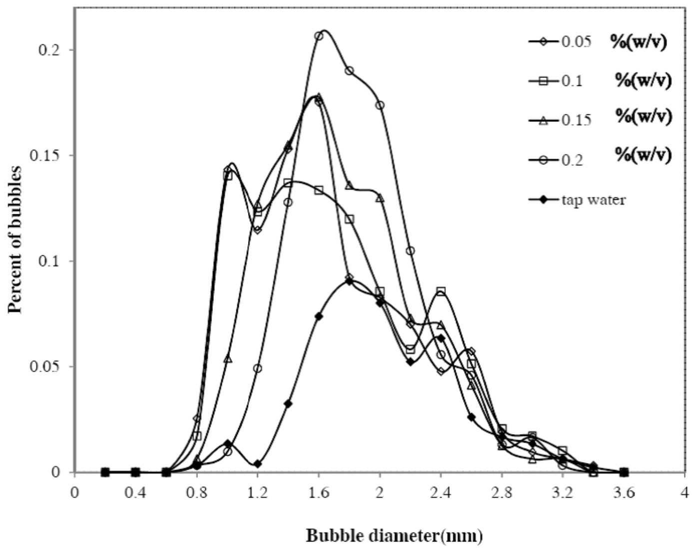

Figure 4. Percentages of bubbles with the same diameter in the riser with different CMC concentration at the Ug of 0.4cm/s.

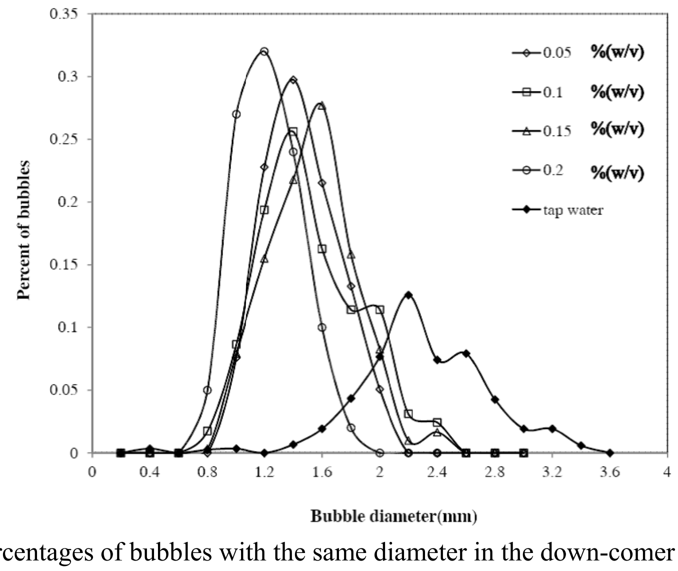

Figure 5. Percentages of bubbles with the same diameter in the down-comer with different CMC concentration at the Ug of 0.4cm/s.

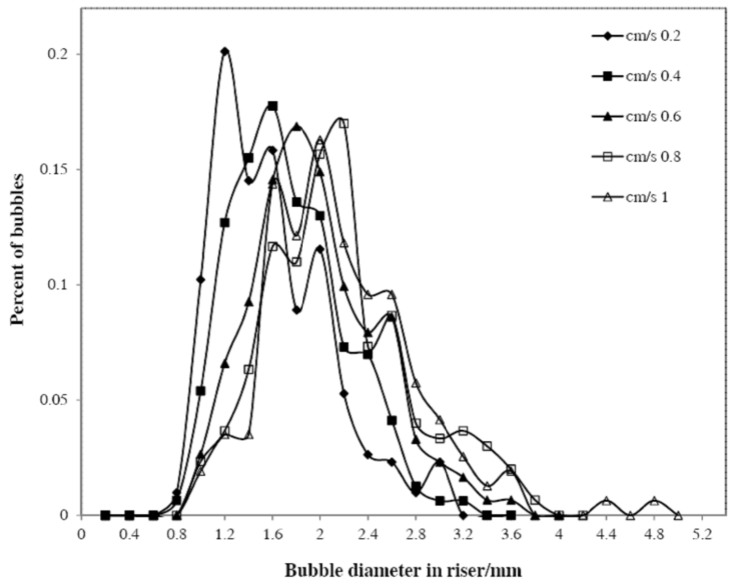

Figure 6. Percentages of bubbles with the same diameter in the riser with different Ug at the CMC concentration of 0.15 (w/v).

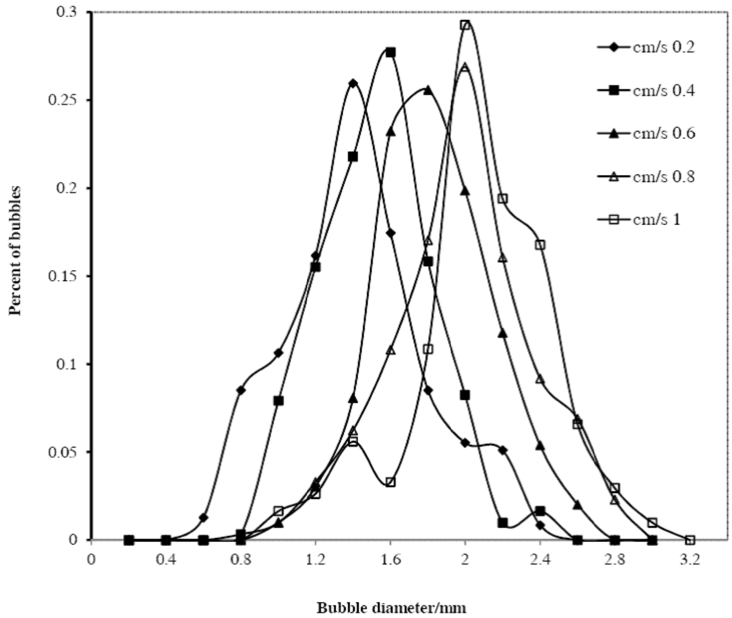

Figure 7. Percentages of bubbles with the same diameter in the down-comer with different Ug at the CMC concentration of 0.15 (w/v).

In accordance with Denget. al [15], our results show that bubble diameter increases by increasing the gas velocity, although Yoshimoto *et al*. [26] found that increasing bubble diameter is approximately independent of aeration velocity.

Furthermore, bubble diameter increases by increasing the CMC concentration. Deng et. al [15] reported that the increase in CMC concentration (liquid viscosity) led to a decrease in turbulence intensity, which in turn decreased the bubble breakup and coalescence. Further, the bubble breakup was more sensitive to liquid viscosity than bubble coalescence, thus the bubble size distribution shifted to larger size with an increase in CMC concentration [15].

3-2. Gas hold-up

Figs. 8 and 9 show gas hold-up versus superficial gas velocity (U_g) for various CMC solutions in the riser and down-comer, respectively. Fig. 10 also shows overall gas hold-up versus superficial gas velocity. According to these figures, gas hold-up increased by increasing the superficial gas velocity. This result is in agreement with the literature [15].

Furthermore, the gas hold-up in the riser and overall gas hold-up linearly increased with U_g while it initially rose slightly and then increased sharply against gas hold-up in the down-comer. The gas velocity enhancement increases the small bubbles dispersion in the reactor. Moreover, it can produce the larger bubbles. Therefore, the gas hold-up increases [27].

Figs. 8 and 10 also show that the gas hold-up reduces by increasing the CMC concentration. Since the big bubbles in the viscous solutions have short residence time, the gas hold-up decreases. The gas hold-up increases (more than that in the concentrations of 0.1 and 0.15 %) when bubbles diameter decreases in concentration of 0.2 %. Hwang and Cheng [28] showed that the gas hold-up in the down-comer for 0.8 wt% CMC solution was higher than that for 0.5 wt% CMC solution. Fransolet *et al.* showed that the gas hold-up was close for the 4 wt% and 5 wt% xanthan gum solutionsin U_g less than 8 cm/s [29]. They reported two mechanisms for this result: one was that decreasing the bubble increases velocity that resulted in longer gas holdup, and the other is, increasing the bubble increases velocity with an increase in bubble diameter.

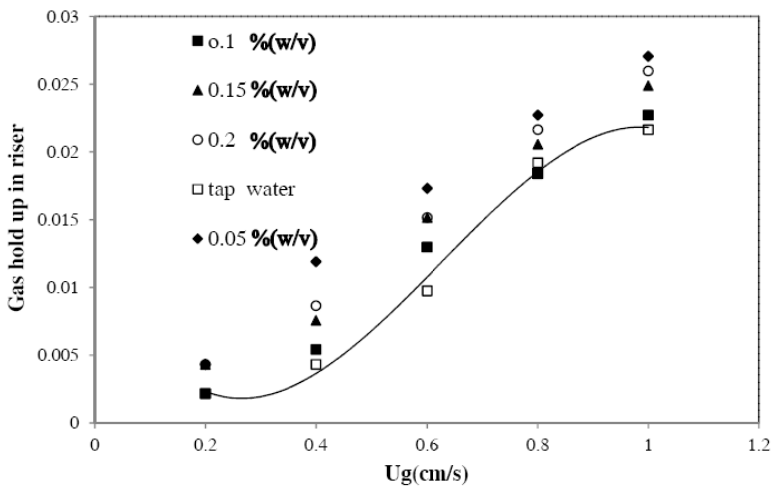

Figure 8. Gas hold-up versus superficial gas velocity in the riser.

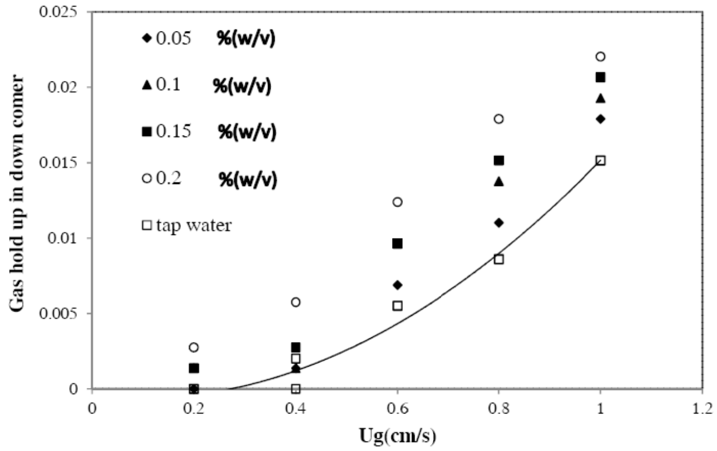

Figure 9. Gas hold-up versus superficial gas velocity in the down-comer.

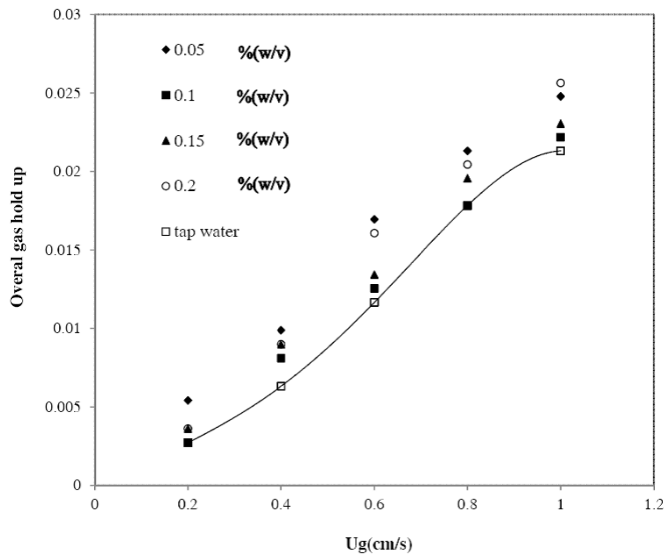

Figure 10. Overall gas hold-up versus superficial gas velocity.

Deng *et al*. [15] showed that the gas hold-up for the 0.40 wt% and 0.45 wt% CMC solutions were quite similar in U_g less than 6 cm/s. The bubble rise velocity reduction (with liquid viscosity increment) causes longer bubble residence time and gas hold-up increment. Further, the bubble velocity enhancement (with bubble diameter reduction) increases the liquid viscosity due to the bubble break up reduction.

The bubble diameter increases with CMC concentration increment in the down-comer and gas hold-up increases. According to the current research, since the gas aeration velocity was in the limited range, the number of bubbles that move into the down-comer will be few. Fig. 11 shows a comparison for gas hold-up in the riser and down-comer.The gas hold-up difference in the riser and down-comer creates the driving force for liquid circulation velocity.

Therefore gas hold-up rises with superficial gas velocity increment in the riser and down-comer. A similar trend is observed for overall gas hold-up, as well. However gas hold-up in the riser decreased with Figure increasing the

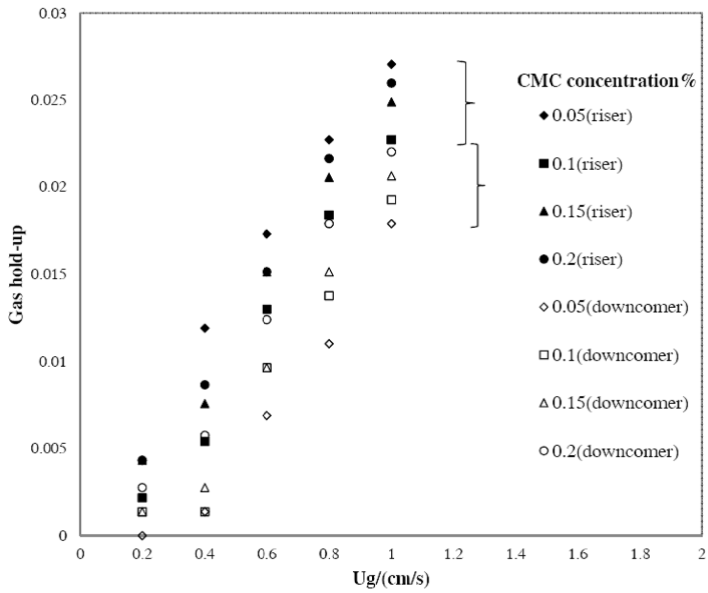

Figure 11. Effect of CMC concentration (0.5-2g/l) on the gas holdup in the riser and down comer zones at various aeration velocities.

CMC concentration, but opposite trend was observed at concentrations of 0.2 and 0.1%. Furthermore, gas hold-up increased by increasing the bubble diameter in down-comer when CMC concentration enhanced.

4. Conclusions

The effects of CMC addition into water and its various concentrations on the bubbles diameter and gas hold-up in a split-cylindrical airlift reactor were investigated. Average bubbles diameter as a function of various superficial gas velocities for the CMC solutions was also calculated. The results show that gas velocity increment increases the bubble diameter and gas hold-upin the riser and down-comer. Further, the surface tension increases when the CMC solution is concentrated. Therefore, bubble diameter increased gas hold-up decreased in the riser. In this study, the bubble break up and critical point were obtained in the riser at concentration of 0.15 %. So, bubbles break up rate increased, bubble diameter decreased and gas hold-up increased at this concentration. Moreover, bubble diameter enhanced with CMC concentration increment in the down-comer. Therefore, gas hold-up increased.

Nomenclature

d_1	maximum bubble diameter [mm]
d_2	minimum bubble diameter [mm]
d_v	equivalent of bubble diameter [mm]
d_{ave}	average diameter of bubbles [mm]
N	number of bubbles
U_g	superficial aeration velocity in the riser zone [m/s]

Greek symbols

ε	overall gas hold-up in bioreactor
ε_d	average gas hold-up in down-comer zone
ε_r	average gas hold-up in riser zone
ρ_G	density of air [kg/m^3]
ρ_L	density of water [kg/m^3]

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