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BUBBLE PROPERTIES AND MIXING CHARACTERISTICS IN AN AIRLIFT BUBBLE COLUMN WITH REDISTRIBUTOR

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Abstract—The effects of gas velocity, liquid velocity, redistributor and their free surface area on bubble properties (size, rising velocity, size distribution) and mixing characteristics (mixing time, circulation time) were studied in an internal airlift bubble column. Bubble properties were measured by resistivity probe method, and mixing characteristics were obtained by salt-base impulse method in air-water system. The empirical correlations between bubble properties, mixing characteristics and various operating parameters were proposed. The disk type redistributor was more effective to redisperse coalesced bubbles than the perforated plate type. Especially, the disk type with 19.5% free surface area had given the best result.

INTRODUCTION

The micro-organisms, substrate, and oxygen must be distributed as uniformly as possible throughout the liquid phase in order to operate successfully at aerobic bioreaction such as the production of single cell protein (SCP). In addition, it is very important to avoid foaming as well as flotation and sedimentation of reaction compounds. Also economical supply of oxygen requirement and effective removal of reaction heat may be required [1]. In recent years, airlift reactors have been the object of much attention, because they fulfill these demands very well, are suited for the fermentation of sensible organism and have no problem of sealing due to the absence of moving parts.

Their configurations are a bubble column with external loop and a bubble column divided into two sections by vertical baffle or draft tube installed in the interior. As the gas is sparged into one section, the density difference between the bubbly liquids in the riser and the downcomer occurs and induces the stable liquid circulation. This circulating liquid flow enhances the heat transfer between the fluid and the column wall and makes the liquid properties homogeneous in the column [2]. However an increase of the recirculation velocity leads to lower K_La values due to reduction of ε_r. Therefore an optimum recirculation rate must exist [3].

Bubble coalescence and breakup play a significant role in the performance of fermentors so that a fine dispersion of disperse phase is an important requirement for successful operation of any bioreactor. For enhancement of oxygen transfer in airlift reactors by the prevention of bubble coalescence and the redispersion of coalesced gas bubbles, studies have been carried out.

Gasner [4] performed a theoretical analysis to improve the economical performance of concentric cylinder airlift system, which was based upon a minimization of coalescence and a maximization of bubble entrainment, so he developed a rectangular airlift reactor. Lin et al. [5] have investigated oxygen transfer and mixing in a baffled airlift tower fermentor. They have shown that the value of K_La in the baffled airlift fermentor would only be 5% greater than that in the fermentor without the baffle. The installation of the baffle section in the airlift tower would drastically prolong both mixing and circulation time of liquid medium compared to that without the baffle. In the work by Maclean et al. [6], the oxygen transfer coefficients and dispersion coefficients were determined in an airlift tower containing Koch static mixers. Their results indicate that the dispersion model is the most appropriate model for the mass transfer coefficient from the available data. Gopal and Sharma [7] have made a study of the hydrodynamics and mass transfer characteristics of bubble and packed bubble columns with downcomer. They have shown that the packed bubble column with downcomer was seen to exhibit higher values of interfacial area and K_La than those obtained in conventional bubble columns and packed bub-

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ble columns.

Although the redistributor has been employed industrially to reduce the high circulation velocity and to redisperse coalesed gas bubbles in the airlift reactor, little information has been published on bubble properties and mixing characteristics in the airlift reactor with redistributor [3, 8].

The purpose of this paper is to investigate the effects of gas velocity, liquid velocity, free surface area and type of the redistributor on the gas holdup, mixing time, circulation time, bubble properties in a concentric cylinder airlift bubble column containing redistributor.

EXPERIMENTAL APPARATUS AND PROCEDURE

The main column and draft tube in the experimental setup (Figure 1) were made of acrylic transparent pipe. The dimensions of the various parts of the equipment are listed in Table 1. Tap water and air were used as the liquid and gas phases in the experiments. The liquid and gas velocities in the experiments ranged from 0 to 2 cm/sec and from 2 to 20 cm/sec, respectively, based on

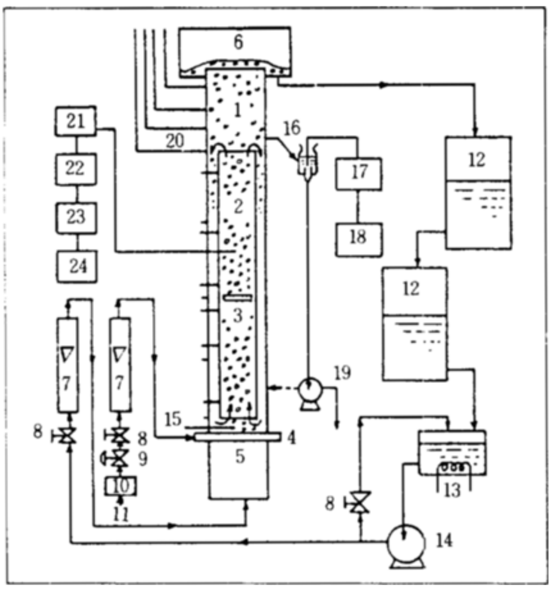

Fig. 1. Schematic diagram of experimental apparatus;

1. main column	2. draft tube
3. redistributor	4. distributor
5. calming section	6. gas-liquid separator
7. rotameter	8. valve
9. pressure regulator	10. air filter
11. air line	12. water tank
13. heater	14. pump
15. tracer injection	16. bubble separater
17. capacitance meter	18. chart recorder
19. circulation pump	20. manometer
21. resistivity circuit	22. oscilloscope
23. A/D converter	24. microprocessor.

Table 1. Dimensions of the experimental apparatus.

Main column I. D.	23.8 cm
Main column height	155.0 cm
Draft tube I. D.	14.6 cm
Draft tube height	110.0 cm
Thickness of main column and draft tube	0.6 cm
Gas-liquid separater	40×40×20 cm
Distance of draft tube from bottom	5 cm
Redistributor type	perforated, disk
Free surface area of redistributor	6.0, 13.0, 19.5, 26.0%

the cross-sectional area of the draft tube. All experimentals were carried out ambient temperature (23 to 27°C) and atmospheric pressure.

Air and water were fed into the draft tube through an air-water distributor. The water distributor was a perforated copper plate, consisting of 4mm diameter 37 holes, 2.1cm triangular pitch. Filtered air was sparged into the column through six tubes with 1mm diameter 30 holes drilled with 2.1cm triangular pitch and attached on the plate.

Two types of redistributor (perforated plate and disk) were used in this work. The redistributors were made of flat acrylic sheet 1.5cm thick and fastened across the riser section 50cm above the distributor. Their free surface areas were varied as 6.0, 13.0, 19.5, 26.0%. In the perforated plate type, 155 holes were arranged with 1cm rectangular pitch.

Water manometers were used to obtain local holdup data. There were eight pressure taps installed along the main column and six along the draft tube. For measurement of circulation and mixing time, 10ml of 4M potasium chloride solution was used as a tracer. The response of a pulse input of the tracer was measured by a capacitance meter and simultaneously recorded on a chart paper. The circulation time, t_C, and mixing time, t_m, were estimated from the response curves. The circulation time was determined as the mean time period between the two adjacent peaks. The maximum and minimum peaks of the response curve are connected by smooth curves, respectively. Then two curves encounter at a point on the response curve. The mixing time is measured as the time required to achieve 3% inhomogenity throughout the reator space. The inhomogenity is defined as the ratio of the difference between the concentration of tracer at impulse maximum, C, and the mean concentration for total tracer distributed in reactor, C_∞, to C_∞.

In order to consider the effects of the presence of redistributor on bubble properties, bubble properties had been measured by resistivity probes placed 20 cm downstream and 30 cm upstream of the redistributor. The probe consisted of two needles which were made of chromel-alumel wire 0.25 mm in diameter and coated with epoxy resin except for the needle tip. The vertical distance between the two tips was maintained 3mm. The probe was supported by a stainless steel tube which served as an opposite electrode and was placed in center of the column. A d.c. voltage was applied to the probe and the signal from the probe was amplified, fed to an A/D converter, and storaged in a microprocessor [9].

The discharge coefficient for the perforated plate type redistributor was obtained from the phase pressure drop across the plate measured by pressure taps of D_d downstream and $0.5D_d$ upstream from the redistributor.

RESULTS AND DISCUSSION

1. Gas holdup

The effect of the presence of redistributor on the local gas holdup in draft tube and annular region is shown in Figure 2. Installation of the redistributor increased the local gas holdup in draft tube and decreased it in annular region. For low gas velocity or smaller free surface area of redistributor, the local gas holdup in annular region was negligible. This means that the presence of redistributor resisted to the liquid circulation, and led to decrease the velocity of liquid circulation and entrained bubbles in the annular region.

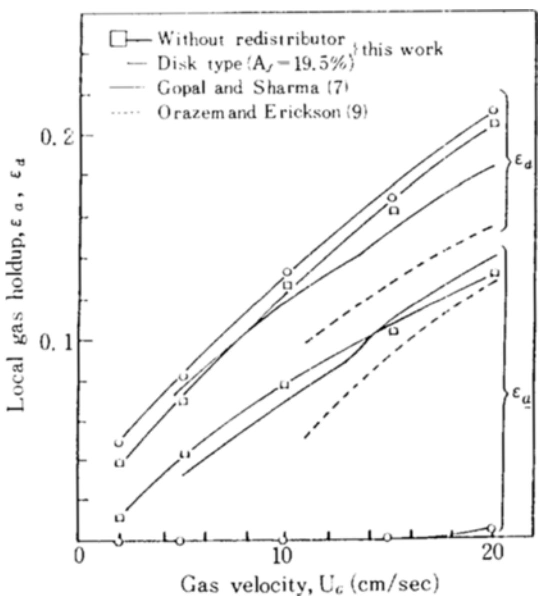

Fig. 2. Effect of the presence of redistributor on the local gas holdup in draft tube and annular region.

The values of local gas holdup reported by other investigators [7,10] are also shown in Figure 2. Their values were observed to be lower than those found in this work. This may be due to the difference of downcomer-to-riser area ratio.

The overall gas holdup was increased according to the increase of gas velocity and showed a slight decrease with respect to increase of the liquid velocity. Previous investigators also observed similar results in an airlift reactor [11,12].

As shown in Figure 3, the local gas holdup in draft tube was increased as the free surface area fraction, A_f, of redistributor decreased. The local gas holdup in each section using the perforated plate type of redistributor was higher than that of the disk type.

The bubble were entrained in the annular region when the liquid velocity was higher than the terminal velocity of the bubbles. As the gas velocity increased, more bubbles were entrained and the gas holdup in annular region increased due to the increasing circulation velocity. In this work, the overall gas holdup without redistributor was higher than that obtained with redistributor, because the area ratio of draft tube to annulus was fixed at 0.673. If the inner volume of draft tube is larger than the annular volume, the overall gas holdup with redistributor may be higher than that obtained without redistributor. For a given aeration rate the overall gas holdup using the perforated plate type of redistributor was higher than that of the disk type.

The local gas holdup in draft tube was correlated with operating parameters as in Figure 4, and the following relationship was found:

$$\varepsilon_d = 0.020\,(\exp(-0.077U_L))\,(U_G)^{0.712}(A_f)^{-0.182} \tag{1}$$

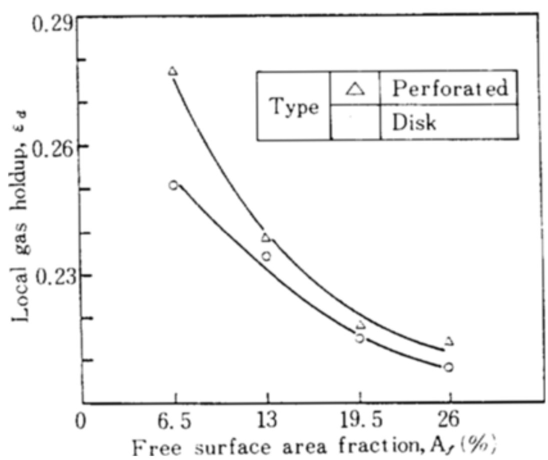

Fig. 3. Effect of the free surface area fraction on the local gas holdup in draft tube for U_L=1.5cm/sec and U_G=20cm/sec.

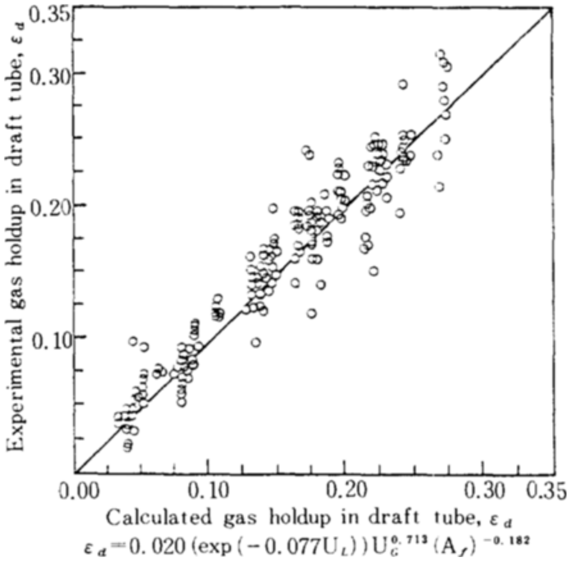

Fig. 4. Comparison of experimental local gas holdup in draft tube with calculated value. (Standard deviation: 0.016; regression coefficient: 0.977).

As shown in the correlation, the gas holdup was a strong function of the gas velocity.

2. Circulation time and mixing time

The effect of the gas velocity on the circulation and mixing time is shown in Figure 5. Without regard to the presence of redistributor, the circulation time and mixing time decreased as the gas velocity and bubble size increased. The plug flow pattern in annular region gradually diminished with the increasing gas velocity due to the increase of entrained bubbles. The presence of redistributor increased the circulation time and mixing time. The percentage increase in circulation time was found to change from about 30% at higher A_f of redistributor (e.g. 26%) to 800% at lower A_f (e.g. 6.5%). The mixing times using redistributor with A_f of 6.5% were about two times of those without redistributor. It means that the presence of redistributor interrupts the liquid circulation. Merchuk and Stein [13] reported that high liquid circulation velocities were obtained by minimizing the resistances to fluid flow in the circuit.

The reduction of the liquid recirculation velocity leads to an oxygen transfer limitation in the upper part of the airlift reactor and to considerable cell sedimentation. An increase of the recirculation velocity leads to lower K_La values. Adler and Schügerl [3] have found that an optimum recirculation rate must exist. Therefore the redistributor can be used to regulate the liquid circulation velocity.

With the redistributor, a bubbly liquid front moved progressively down the column as the gas velocity was increased. The opposite trend was observed with respect to increasing liquid velocity. At higher gas flow rate, U_G = 20 cm/sec, the front eventually reached the base of the downcomer and the onset of gas recirculation occured.

Figure 6 shows the effect of A_f of redistributor on the

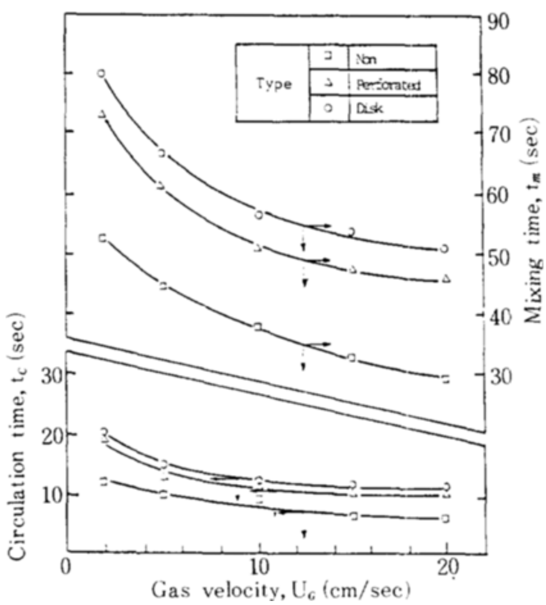

Fig. 5. Effect of the gas velocity on the circulation and mixing time for U_L=0cm/sec.

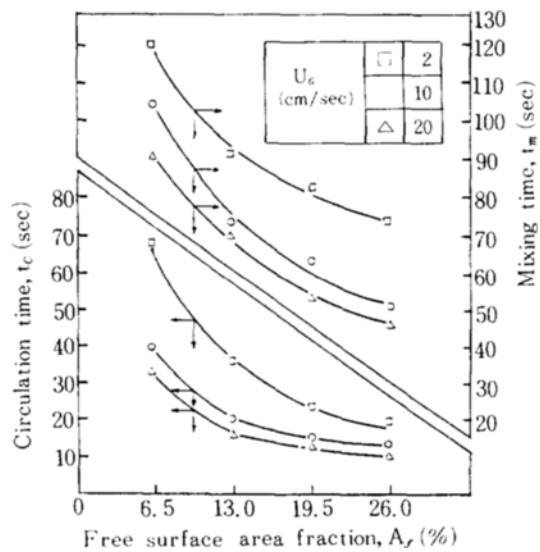

Fig. 6. Effect of the free surface area fraction of redistributor(perforated type) on the circulation and mixing time for U_L=0cm / sec.

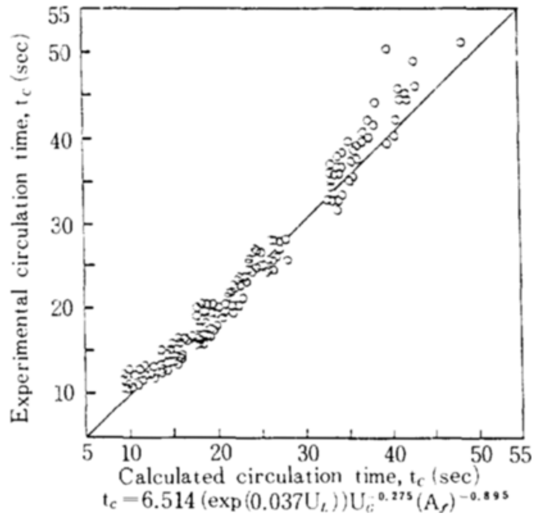

Fig. 7. Comparison of experimental circulation time with calculated value. (Standard deviation: 1.815; regression coefficient : 0.986).

circulation and mixing time. The circulation and mixing time decreased with increasing A_f. It is due to decrease of the resistance to circulation flow. The circulation and mixing time could be related to operating variables by:

$$t_c = 6.514(\exp(0.037U_L))(U_G)^{-0.275}(A_f)^{-0.895} \quad (2)$$

$$t_m = 38.24(\exp(0.001U_L))(U_G)^{-0.168}(A_f)^{-0.532} \quad (3)$$

and these correlations were shown in Figure 7,8, respectively.

3. Bubble characteristics

Without the redistributor, the effect of the gas velocity on the mean vertical bubble length as a function of bed height is shown in Figure 9. Due to coalescence, the vertical bubble length increased as the gas velocity and bed height increased.

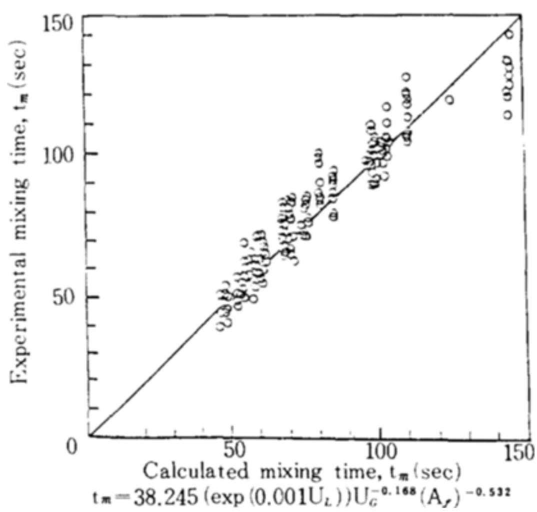

Fig. 8. Comparison of experimental mixing time with calculated value. (Standard deviation: 7.433; regression coefficient: 0.950).

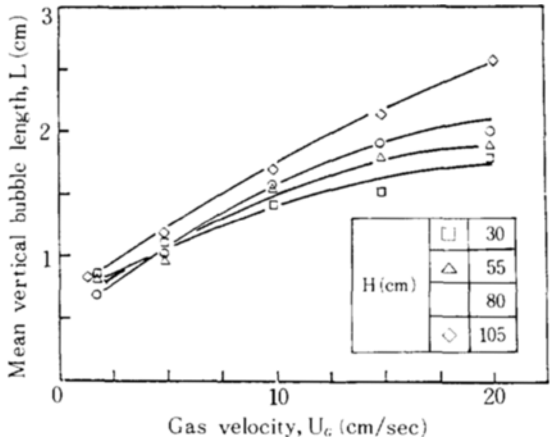

Fig. 9. Effect of the gas velocity on the mean vertical bubble length as a function of heigth for U_L=2cm/sec(without redistributor).

Without the redistributor, it was observed that bubble size distribution curve became more broad and shifted toward the large bubble region as the gas velocity and bed height increased which may be due to coalescence.

Figure 10 shows the effect of presence of the disk type redistributor on mean vertical bubble length. The mean vertical bubble length in the downstream of the redistributor was higher than that of without the redistributor, because of the flow restriction effect by the redistributor. Generally the mean vertical bubble length in the upstream of the redistributor was smaller than that in the downstream of the redistributor. The bubble size distribution in the upstream of the redistributor shifted toward to smaller bubble region than that in the downstream of the redistributor. Furthermore, the former was more uniform than the latter, and similar results were observed in case of the perforated plate type redistributor. It means that both type of redistributors are effective in reducing the size of the gas bubbles.

Figure 11 shows the effect of A_f on the mean vertical bubble length. For higher gas velocity (e.g. 15-20 cm/sec) and extream values of A_f (e.g. 6.5 and 26.0%), the perforated plate type redistributor was ineffective in size reduction of coalesed bubbles. Adler et al. [3] observed that the perforated plate with A_f of 6.52% was very ineffective aerators, because it formed large bubbles which quickly passed the liquid layer before they could be redispersed. Ineffective gas distributors cause a

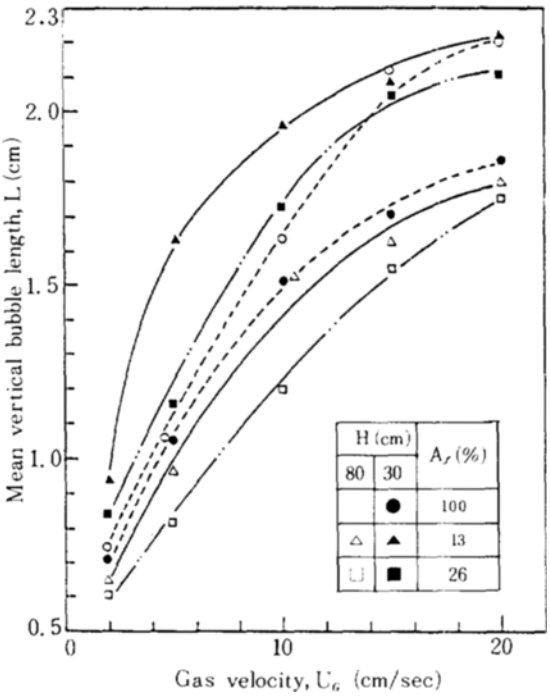

Fig. 10. Effect of presence of the disk type redistributor on mean vertical bubble length for $U_L = 0.75$cm/sec.

reduction of oxygen transfer rate, and oxygen limitation occurs. Except for 6.5% free surface area, the disk type redistributor acted as an effective redistributor, especially, the disk type with 19.5% free surface area had given the best result.

The bubble rising velocity depended on the bubble diameter and the liquid circulation velocity. Without the redistributor, the bubble rising velocity, U_b, increased as the gas velocity and bed height increased. The bubble rising velocity was increased due to increasing liquid circulation velocity as A_f increased.

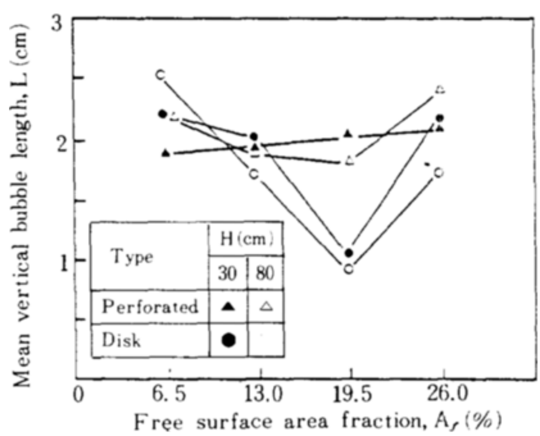

Fig. 11. Effect of the free surface area fraction on the mean vertical bubble length for $U_L = 2$cm/sec and $U_G = 20$cm/sec.

In this experimental range, the effect of the liquid velocity on the mean vertical bubble length and bubble rising velocity in the upstream of the redistributor was negligible.

Without the redistributor, the mean vertical bubble length was correlated with operation parameters as in Equation [4] and was shown in Figure 12.

$$L = 0.361\,(U_L)^{-0.033}(U_G)^{0.438}(H)^{0.101} \qquad (4)$$

Crawford [14] suggested that the log-normal probability distribution of particle diameter as:

$$P(Li) = \frac{1}{\sqrt{2\pi\beta}} \exp\left(\frac{(\ln Li - \alpha)^2}{2\beta^2}\right) \qquad (5)$$

$$\text{where: } \alpha = \ln(Lm) \qquad (5-1)$$

$$\beta = \ln(\sigma_m)$$

The relationship between bubble size and cumulative number fraction of bubbles was a linear relation when it was plotted on log-probability paper as in Figure 13. It is seen that the bubble size distributions follow the log-normal distribution function. In the investigation of bubble size distribution in a cylindrical split-column airlift fermentor using photographic methods, Glasgow et al. [15] found that the bubble size distribution was adequately described by the lognormal distribution. Fan

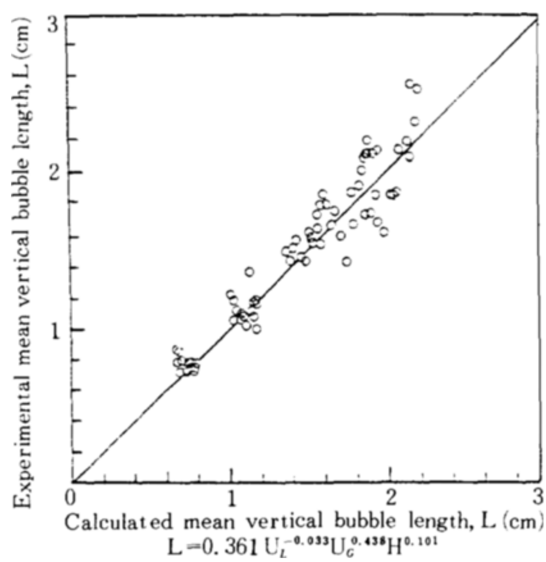

Fig. 12. Comparison of experimental mean vertical bubble length with calculated value. (Standard deviation: 0.140; regression coefficient: 0.961).

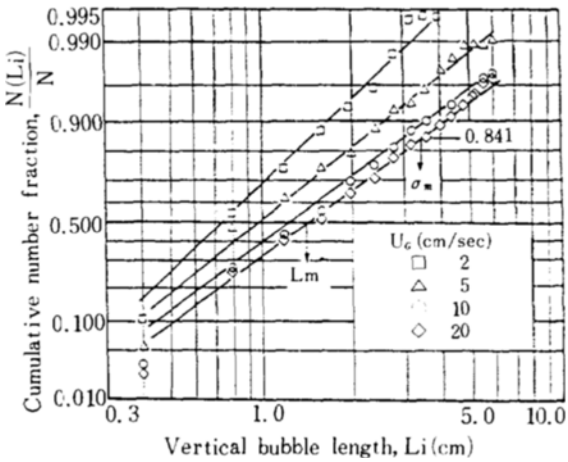

Fig. 13. Log-normal distribution of vertical bubble length for U_L=0cm/sec, H=30cm and A_f=13%(perforated type redistributor).

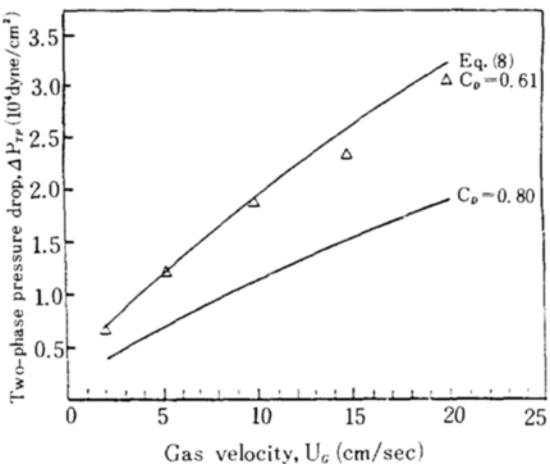

Fig. 14. Variation of two-phase pressure drop with gas velocity for perforated type redistributor, A_f=6.5% and U_L=0cm/sec.

et al. [16] observed similar results in their studies of gas-liquid-solid spouted bed with draft tube.

4. Discharge coefficient of perforated plate

For two phase flow through an orifice, assuming separated flow of the two phases, Lin [17] correlated

$$(\Delta P_{TP})^{1/2} = (\Delta P_L)^{1/2} + (\Delta P_G)^{1/2} \tag{6}$$

With $\Delta P_G \ll \Delta P_L$, ΔP_{TP} is independent of gas void fraction. Assuming homogeneous flow of the two phases, Fairhurst [18] proposed the relationship

$$\frac{\Delta P_{TP}}{\Delta P_{L0}} = \frac{r^2 \rho_L}{\varepsilon \rho_G} + \frac{(1-r)^2}{(1-\varepsilon)} \tag{7}$$

With $\rho_G \ll \rho_L$, $r \simeq 0$. Putting $\Delta P_L = \Delta P_{LO}$, Equation (7) gives

$$\Delta P_{TP} = \frac{1}{1-\varepsilon} \Delta P_L \tag{8}$$

In this work, ε is the gas holdup in the riser region, which can be calculated by Equation (1).

In Equation (8), the pressure drops for liquid flow, ΔP_L, can be calculated from the following equation (9):

$$\Delta P_L = \rho_L U_r^2 R^2 / (2 g_c C_D^2) \tag{9}$$

where U_r is the riser linear liquid velocity. U_r can be obtained from the circulation length, L_c, the experimental equation for circulation time, Equation (2), and the material balances as follows.

$$L_C = 2H_d + D_d + \frac{(D - D_d)}{2} \tag{10}$$

$$\frac{L_C}{2}\left(\frac{1}{U_r} + \frac{1}{U_d}\right) = t_C \tag{11}$$

$$U_r = U_L + \frac{U_d}{A_f} \tag{12}$$

Figure 14 shows a typical comparision of the calculated ΔP_{TP} by Equation (8) and (9) with the measured ΔP_{TP}. It is evident from these data that the discharge coefficient for co-current upward flow of bubbly air-water system through perforated plate is from 0.61 to 0.80 over a range of gas void fraction 0-0.3. This result is similar to the observation of Lewis and Davidson [20]. The large discharge coefficient may be due to gas bubbles breaking up the shear layer around the jet from the orifice which prevents full contraction of the flow.

CONCLUSIONS

Installation of the redistributor increased the gas holdup in draft tube, mixing time and circulation time. Generally, both the perforated plate type redistributor and the disk type redistributor were effective in reducing the size of the gas bubbles in an air-water bubble column. The disk type was more effective to redisperse coalesced bubbles than the perforated plate type. Especially, the disk type with 19.5% free surface area had given the best result. Bubble size distribution was adequately descirbed by the log-normal distribution. For co-current upward of bubbly air-water flow through the perforated plate, the discharge coefficient was from 0.61 to 0.80 in the range of gas void fraction 0-0.3. Empirical equations of gas holdups in draft tube, mixing time, circulation time and vertical bubble length were proposed.

NOMENCLATURE

A_f : free surface area of the redistributor (cm^2)
C : concentration of trace at impulse maximum (gmol/l)
C_∞ : mean concentration for total distributed in reactor (gmol/l)
C_D : discharge coefficient
D : inside diameter of main column (cm)
D_d : outside diameter of draft tube (cm)
g_c : gravitational conversion factor
H : bed height (cm)
K_La : volumetric oxygen-mass transfer coefficient (sec^{-1})
L : mean vertical bubble length (cm)
L_c : circulation length (cm)
Li : vertical bubble length (cm)
Lm : the diameter for which the cumulative distribution curve has a value of 0.5
N : total number of bubbles
N(Li) : number of bubbles with vertical length less than Li
ΔP : pressure drop ($dyne/cm^2$)
ΔP_D : overall pressure loss ($dyne/cm^2$)
ΔP_G : pressure loss at the vena contractor for gas only flow ($dyne/cm^2$)
ΔP_L : pressure loss at the vena contractor for liquid only flow ($dyne/cm^2$)
ΔP_{LO} : pressure loss at the total mass flow as liquid ($dyne/cm^2$)
P(Li) : log-nomal probability distribution
r : mass flow rate of gas/total mass flow rate
R : riser area/opening area = $1/A_f$
t_c : circulation time (sec)
t_m : mixing time (sec)
U_d : linear liquid velocity in the downcomer (cm/sec)
U_G : inlet superficial gas velocity based on the riser cross-sectional area (cm/sec)
U_L : inlet superfical liquid velocity based on the riser cross-sectional area (cm/sec)
U_r : linear liquid velocity in the riser (cm/sec)

Greek Letters

α : size distribution parameter defined by Equation(5-1)
β : standard deviation
ρ_G : gas density (g/cm^3)
ρ_L, ρ : liquid density (g/cm^3)
ε_a : local gas holdup in annular region
ε_d : local gas holdup in draft tube
ε_r : local gas holdup in riser section
σ_m : ratio of the vertical bubble length for which the cumulative-distribution curve has the value of 0.841 to the median vertical bubble length

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